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# CLARIFIER DESIGN

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Under the Direction of the **MOP-8 Subcommittee** of the **Technical Practice Committee**


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The WEF Technical Practice Committee (formerly the Committee on Sewage and Industrial Wastes Practice of the Federation of Sewage and Industrial Wastes Associations) was created by the Federation Board of Control on October 11, 1941. The primary function of the Committee is to originate and produce, through appropriate subcommittees, special publications dealing with technical aspects of the broad interests of the Federation. These publications are intended to provide background information through a review of technical practices and detailed procedures that research and experience have shown to be functional and practical.

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# Preface

This manual describes all aspects of all kinds of clarifiers and alternative clarifying devices from the perspective of design. In addition to documenting the current state of the art and types of clarifiers and clarifier equipment available, it will provide enough clarifier science to allow the user to make critical assessment and comparison of vendor claims. The manual is intended for designers, users, and wastewater treatment plant decision makers.

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# Chapter 1

# Introduction

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## INTRODUCTION

This initial chapter is an overview of the material that will be presented in this update to the Water Environment Federation's 1985 edition of *Clarifier Design* (MOP FD-8). This revised edition will be more up to date and covers a broader range of clarifier applications (i.e., primary, tertiary, storm water, and secondary).

This second edition of the manual provides an update of the existing text and additional chapters outlining primary clarifier design concepts and considerations, high-rate and wet weather clarifier design concepts and considerations, secondary clarifier design concepts and considerations, and tertiary clarifier design concepts and considerations. The manual also addresses topics such as modeling, field testing, circular and rectangular clarifiers, clarifier performance monitoring and control, approaches from outside of the United States, and interlocking with solids-handling facilities. This is intended to be a complete update and expansion of the 1985 edition of *Clarifier Design*, an entirely self-contained design manual, and a companion piece to other manuals such as the International Water Association's *Secondary Settling Tanks: Theory, Modeling Design and Operation* (Ekama et al., 1997*)*, which go into some aspects in more detail and at a more theoretical level.

## APPROACH

It is the intent of the authors of this manual to not only give readers a reference on current design practice but to also give them a resource to better understand vendor information and optimize designs. It is intended to give just enough basic information and science to understand clarifier design but not overwhelm the reader with theory. References are included for those wishing to understand more theory. A number of new concepts are presented, some original and some commonly used. Paramount among these is the d*esign efficiency (DE),* which is the ratio of the clarifier area required by an ideal clarifier to that of a particular design. An ideal clarifier would have a DE of 1.0 and, for example, Ozinzky et al. (1994) and Watts et al. (1996) suggest that typical shallow circular clarifiers have a DE of approximately 0.7 to 0.8 and Ekama et al. (1997) suggest that certain rectangular clarifier designs may have a design efficiency 0.8 to *more than* 1.0. In these cases, the "ideal clarifier" was one that performed according to one-dimensional flux theory. It is expected that, in the near future, vendors will include this ratio in their designs as well as documentation to support their claims.

In this manual, an attempt has been made to have a single, general approach for sizing all types of clarifiers. It is, in its simplest form,

1. Characterize the settling velocity or settling velocity distribution of wastewater the settled.
2. Select design settling velocity $V_d$ (m/h).
3. Calculate the ideal clarifier area $A_{ideal}$ (m$^2$):

$$\mathrm{A}_{ideal} = Q_m / (V_d \times 24)$$

where $Q_m$ is the maximum wastewater flow to clarifier (m$^3$/d).

4. Determine degree of nonideality expected and express it as a DE.

5. Determine design surface area $A_d$:

$$A_d = A_{ideal}/\mathrm{DE}$$

where DE is a characteristic of the particular clarifier design details.

6. Select depth and design details (inlet and outlet designs, baffling, collectors, etc.) to achieve the most cost-effective design.

Clarifiers for treating *stormwater, combined wastewater, raw wastewater,* and *secondary effluent* are primarily discussed in Chapters 2, 3, and 5. Historically these types of clarifiers are categorized as type I or type II settlers. Details of this classification may be found primarily in Chapter 4 but also in Chapters 2 and 3. In these chapters, $V_d$ is chosen by developing a distribution of settling velocities of the particles in the wastewater. This can be a cumulative frequency distribution (fraction of suspended solids settling faster than stated value; a method for accomplishing this, originally developed for Lamella separator design but applicable to any type I or II system, is in Appendix A) or, for raw wastewater, sometimes a velocity corresponding to "settleable solids". In the former case, $V_d$ is chosen to correspond to percent removal desired (i.e., fraction of solids settling faster than $V_d$). For settleable solids, $V_d$ is the settling velocity corresponding to the test procedure, typically approximately 5 m/h and percent removal is determined from this characteristic, not vice versa, as is common in current design approaches.

For clarifiers treating secondary solids (typically including chemical solids resulting from phosphorus removal and called *tertiary clarifiers*), $V_d$ is case specific and is discussed in Chapter 5.

For all of these clarifiers, it is possible to increase $V_d$ (i.e., reduce design clarifier area $A_d$) by providing flocculation and/or by adding chemicals and/or ballasting agents. Another option for these types of clarifiers is to include tubes or plates to increase the settling area available for a given footprint. Details of how to do this are discussed in each appropriate chapter and in most detail in Chapter 3.

*Secondary clarifiers* (i.e., clarifiers that are part of an activated sludge system) treat higher concentrations of suspended solids, which settle as a uniform mass at a uniform initial settling velocity (ISV). This is traditionally referred to as type III or zone settling (see Chapter 4 for more details of this designation). These solids are called mixed liquor suspended solids (MLSS). Here, $V_d$ = ISV and is a function of the biology (primarily how many and what type of filamentous organisms are in MLSS) and the concentration of the MLSS ($X$). Typically, the mixed liquor quality is represented by tests like sludge volume index (SVI) and/or settling constants like $V_o$ and $k$. Chapter 4 describes methods to measure and estimate ISV.

For MLSS, the clarifier size can be reduced by lowering the MLSS concentration ($X$) fed to the clarifier. One approach to doing this is to design the preceding aeration basin to have step-feed capability. Another approach is improving SVI (or $V_o$ and $k$) by aeration tank design (such as using selectors) and/or operation. These are discussed in Chapter 4. Yet another approach is to use chemicals and/or ballasting agents. These are discussed in detail in Chapter 3.

Chapter 7 discusses approaches to measuring how close an existing clarifier is to ideal.

Chapter 6 is devoted to modeling. These approaches allow a designer to analyze various design options and improve clarifier design to more closely approach ideal.

## TRADITIONAL AND VENDOR APPROACHES

It is recognized that not all designers will have the resources or inclination to follow the preferred approach described above. Accordingly, most chapters also include some of the traditional, more empirical approaches that appeared in previous manuals as well as approaches recommended by vendors of proprietary equipment.

## A WORD ABOUT THICKENING

When a wastewater is clarified, the collected solids are called *primary sludge* (for primary clarifiers and stormwater clarifiers); *secondary sludge* (for clarifiers that are part of secondary treatment); *tertiary sludge* (from tertiary clarifiers); and, sometimes, (particularly in Europe) *humus* when sludge is from a trickling filter or another attached-growth biological process. Historically, many clarifiers have been designed and operated to thicken primary sludge and humus to approximately 4% and secondary and tertiary sludge to more than 1%. In this manual, it is advocated that thickening and clarification be separated—conducted in separate processes—and that most sludge be drawn “thin”. Thickener designs are not part of this manual but may be found elsewhere (WEF, 1998).

## CHAPTER DESCRIPTIONS

A brief description of each of the other chapters included in this revision is as follows. Chapter 2, Primary Clarifier Design Concepts and Considerations, includes design concepts for primary clarifiers and when to use them. This chapter also features information on clarifier enhancements. Chapter 3, High-Rate and Wet Weather Clarifier Design Concepts and Considerations, includes information detailing swirl concentrators and various types of very high-rate chemically augmented clarification systems. This chapter also features information on clarifier enhancements. Chapter 4, Secondary Clarifier Design Concepts and Considerations, includes data examining flux theory and the latest approaches for sizing clarifiers for attached growth, activated sludge (suspended growth), moving bed biofilm reactors, and combined (integrated fixed-film activated sludge) systems. Chapter 5, Tertiary Clarifier Design Concepts and Considerations, covers all applicable topics involving tertiary clarifiers from a design standpoint.

Chapter 6, Mathematical Modeling of Secondary Settling Tanks, covers all of the latest information on software availability. One-, two-, and three-dimensional models are discussed in this chapter.

Chapter 7, Field Testing, details when field testing is needed or required. Within this chapter, Clarifier Research Technical Committee and other field testing procedures are presented.

Chapter 8, Circular Clarifiers, details equipment selection, "nuts and bolts", trends, and problems in reference to circular clarifiers. Chapter 9, Rectangular Clarifiers, includes detail about equipment selection, "nuts and bolts", trends, and problems in reference to rectangular clarifiers.

Chapter 10, Clarifier Performance Monitoring and Control, addresses topics that include key parameters, monitoring and control equipment, and interaction of clarifiers with other facilities.

Chapter 11, International Approaches, discusses approaches used outside of North America, focusing on European practice.

Chapter 12, Interaction of Clarifiers with Other Facilities, examines design approaches with the rest of plant in mind.

The manual describes all aspects of all kinds of clarifiers and alternative clarifying devices from the perspective of design. In addition to documenting the current state of the art and types of clarifiers and clarifier equipment available, it also provides enough clarifier science to allow the user to make critical assessment and comparison of vendor claims. The manual will also include performance data and case histories where appropriate.

The organization of this manual inevitably results in some overlap of similar topics in multiple chapters. The decision was made to leave in most of these redundancies to make the manual easier to read. The authors have attempted to reference where topics are discussed elsewhere in the manual, but the reader is encouraged to use the index to find other chapters where a given topic might be discussed in more detail or from a different perspective.

The manual is being written for use by designers who are given the choice of using traditional methods or newer approaches, depending on their particular resources and nature of their project. They are given tools to "demystify" vendor claims and improve their designs. They are also given enough "nuts and bolts" information to make detailed design decisions, information on what shape and depth a clarifier should be and what inlets and outlets should look like. Users will be able to compare their clarifiers to what others have and are given objective ways of analyzing and improving their clarifiers. Vendors are given the methodology to demonstrate the superiority of their designs. It is expected that they will start to include design efficiency as part of their literature, documented according to the procedures outlined in this manual. Wastewater treatment plant decision makers will find this a resource to better understand what designer and vendors tell them and make more informed decisions.

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# Chapter 2

# Primary Clarifier Design Concepts and Considerations


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## INTRODUCTION

Gravity separation of solids from liquid, producing a clarified overflow and a thickened solids underflow, has long been used in the wastewater treatment industry. Often, the terms *clarification* and *thickening* or *sedimentation* are used to describe gravity separation unit operations, depending on if the process focus, or objective, is on the clarified liquid or the thickened solids, respectively (Rich, 1961). Many primary clarifiers are deliberately designed and/or operated to produce a thickened primary sludge, a fact further exemplified by the practice of pumping waste activated sludge to primary clarifiers for co-thickening with primary sludge. Perhaps it is for this reason that the profession is confused about the process objective of primary clarifiers as they are just as commonly known as primary sedimentation tanks or primary settling tanks. While the solids concentration in primary sludge is an important consideration insofar as the solids treatment train is concerned, more depends from an overall plant perspective (e.g., sizing and operating expense of downstream units) on the quality of primary effluent than the solids concentration of primary sludge. Thickening sludge in primary clarifiers brings more detriment to the liquid treatment train (in the form, for example, of decreased activated sludge settleability resulting from increased organic loadings, hydrogen sulfide production, and volatile acid production) than benefit to the solids treatment train. Because the process objective is more appropriately focused on the clarified liquid, this unit operation herein will be referred to as primary clarification and the units themselves as primary clarifiers.

Design of primary clarifiers has historically been done more empirically than rationally. The main reason for this is a lack of understanding of what pollutants primary clarifiers are capable of removing. For example, it is not uncommon to see in many wastewater treatment plant master or facilities plans a statement such as "The primary clarifiers are designed to remove 60% of the total suspended solids". Never is any basis given for such statements. In reality, 60% removal is assumed, not designed for. With an understanding of the development provided in the pages that follow, the more appropriate statement would be "The primary clarifiers are designed to remove all of the settleable total suspended solids during average dry weather flow conditions". As the settleable total suspended solids concentration is a characteristic of the wastewater, good primary clarifier design begins with a characterization of the wastewater.

Subjects discussed in the following sections include primary clarifier performance with an emphasis on the process objective of primary clarifiers and factors affecting performance; chemically enhanced primary treatment; and design concepts and considerations, including wastewater characteristics, primary clarifier configuration and depth, flow splitting, inlet design, sludge collection and withdrawal, scum collection and withdrawal, and effluent discharge. Finally, although this design manual is intended as a guide for designers, the user should understand that the "science" of primary clarification is not completely understood. For this reason, a brief section on research needs is given.

As the title of this chapter suggests, this is not a "recipe" for primary clarifier design. Instead, what the reader will find in these pages is a discussion of the important factors in primary clarifier design. The discussion begins with the identification of a performance goal. Simplistically, a perfectly designed and operated primary clarifier will have an effluent total suspended solids (TSS) concentration equal to the nonsettleable TSS concentration in the influent to the primary clarifier. With increasing surface overflow rate, increasing concentrations of settleable TSS in the effluent occur, with a concomitant increase in the settleable, particulate biochemical oxygen demand (BOD) and chemical oxygen demand (COD). Often, the process capacity of primary clarifiers is defined, even regulated, in terms of the surface overflow rate. In actuality, the capacity of primary clarifiers, in conventional applications, is defined by the oxidative capacity of downstream biological processes. When special uses exist (such as treatment of combined sewer overflows or blending related to ocean dischargers with 301H waivers), careful consideration must be given by the designer and the operator to optimize primary clarifier performance.

## PERFORMANCE

Municipal wastewater treatment agencies have come under steadily increasing pressure to optimize, to get the absolute most capacity out of existing and new facilities to minimize the cost to ratepayers. This "bottom line" has always been the focus in industrial wastewater treatment facilities. These optimization pressures have resulted in renewed interest in primary clarification at many facilities, municipal and industrial, and with very good reason: primary clarifiers can potentially remove more TSS and COD or BOD for less operational cost than any other treatment process in use today. Primary clarification, depending on wastewater characteristics, can have a profound effect on the size, capacity, and performance of downstream treatment processes. Primary clarification also continues to find extensive application in combined sewer overflow (CSO) and stormwater treatment systems as discussed in Chapter 3. Under these circumstances, primary clarifiers are used to treat excess stormwater induced as part of a CSO abatement strategy. Flows and loads are often well above those typical of standard practice. The storm-related CSO flow is often bypassed directly to downstream disinfection systems, which cannot function effectively when floatable or settleable solids are present. Therefore, an optimized primary treatment system is necessary for this CSO abatement strategy to be effective.

**PROCESS OBJECTIVE.** As a unit operation, physical forces predominate in the removal of TSS in primary clarifiers. Perhaps it is on this account that many think of a primary clarifier as a constant-percentage TSS removal process. The process objective of primary clarifiers is to remove settleable TSS, whether these solids already exist in the raw wastewater or if they are precipitated solids generated as a result of chemical addition for enhanced suspended solids, phosphorus, or heavy metal removal (see Chapter 3). Despite the fact that primary clarifiers remove only *settleable* TSS, performance historically has been quantified based on the removal efficiency of *total* suspended solids, calculated using eq 2.1:

$$E_{\mathrm{TSS}} = 1 - (\mathrm{TSS}_{\mathrm{PE}}/\mathrm{TSS}_{\mathrm{PI}}) \tag{2.1}$$

Where

$E_{\mathrm{TSS}}$ = TSS removal efficiency (often reported as a percentage),
$\mathrm{TSS}_{\mathrm{PE}}$ = primary effluent TSS concentration (mg/L), and
$\mathrm{TSS}_{\mathrm{PI}}$ = primary influent TSS concentration (mg/L).

In removing settleable TSS, primary clarifiers fortuitously remove the COD (or BOD) associated with them. Because downstream biological processes are sized based on the amount of biodegradable material there is in the primary effluent, the performance of primary clarifiers also is often quantified based on the COD (or BOD) removal efficiency, calculated using eq 2.2:

$$E_{\mathrm{COD}} = 1 - (\mathrm{COD}_{\mathrm{PE}}/\mathrm{COD}_{\mathrm{PI}}) \tag{2.2}$$

Where
$E_{\mathrm{COD}}$ = COD removal efficiency (often reported as a percentage),
$\mathrm{COD}_{\mathrm{PE}}$ = primary effluent COD concentration (mg/L), and
$\mathrm{COD}_{\mathrm{PI}}$ = primary influent COD concentration (mg/L).

**FACTORS AFFECTING PERFORMANCE.** Since the classic works of Hazen (e.g., 1904) and Camp (e.g., 1946), the design and operational variable believed to have the most effect on primary clarifier performance is the surface overflow rate. In reality, however, this does not seem to be the case. Figure 2.1 (Wahlberg et al., 1997),

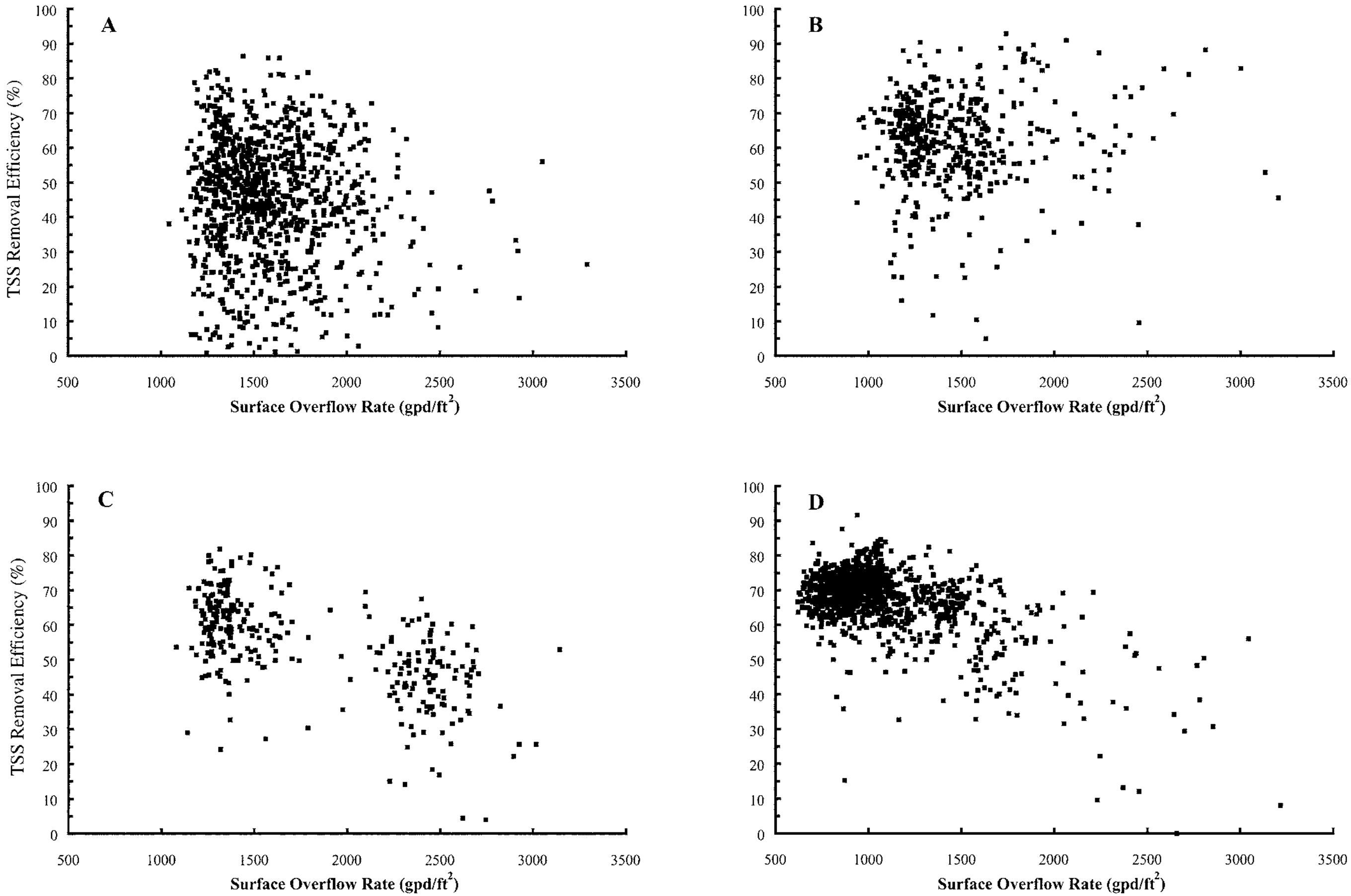


**Figure 2.1** Total suspended solids removal efficiency, $E_{TSS}$, plotted as a function of primary clarifier surface overflow rate (A = Sacramento Regional Wastewater Treatment Plant, B = Dublin San Ramon Services District, C = King County East Section Reclamation Plant, D = King County West Section Reclamation Plant) from Wahlberg et al. (1997) ($m^3/m^2 \cdot d = 0.04075 \times gpd/sq\ ft.$)

typical of historical data at most wastewater treatment plants, shows the TSS removal efficiency as a function of surface overflow rate in primary clarifiers at four wastewater treatment plants. Over a range of surface overflow rates from approximately 24.4 to 134 $m^3/m^2{\cdot}d$ (600 to 3300 gpd/sq ft), TSS removal efficiencies range from essentially 0 to more than 90%. Although there appears to be a downward trend in at least three of these four plots, one cannot conclude from them that there is a strong relationship between TSS removal efficiency and surface overflow rate; that is, it would be unrealistic to provide a "straight-line" correlation between TSS removal efficiency and surface overflow rate from these data. However, two points should be noted. First, a good concentration of these data fall within the range of commonly used assumptions (i.e., 50 to 70% TSS removal at surface overflow rates between 24.4 and 61.1 $m^3/m^2{\cdot}d$ [600 and 1500 gpd/sq ft]). Second, these plots also show that TSS removal efficiencies 50% and greater occur at surface overflow rates on the extreme end of those in practice, 102 to 122 $m^3/m^2{\cdot}d$ (2500 to 3000 gpd/sq ft).

Primary clarifier design fundamentals are grounded in discrete particle (type 1 settling) and flocculent (type 2) settling analyses (see Chapters 3 and 4 for more details). This foundation has been used for most primary clarifier designs in existence today. Additional investigations and research are needed to advance these fundamental theories.

Tebbutt and Christoulas (1975) described the primary effluent TSS concentration in terms of the following equation:

$$TSS_{PE} = TSS_{non} + (TSS_{PI} - TSS_{non})e^{-n\tau} \tag{2.3}$$

Where

$TSS_{non}$ = nonsettleable, influent TSS concentration (mg/L),
$n$ = a constant (1/d), and
$\tau$ = hydraulic residence time (d).

With reference to eq 2.3, it should be noted that the quantity, $TSS_{PI} - TSS_{non}$, is equal to the settleable TSS concentration, $TSS_{set}$. The hydraulic residence time, $\tau$, is equal to the volume of the primary clarifier divided by the influent flow,

$$\tau = V_{PC}/Q_{PI} \tag{2.4}$$

Where

$V_{PC}$ = primary clarifier volume ($m^3$) and
$Q_{PI}$ = primary influent flow ($m^3/d$).

The volume is equal to the surface area times the average depth:

$$V_{PC} = A_{PC} \cdot d \tag{2.5}$$

Where

$A_{PC}$ = primary clarifier surface area ($m^2$) and
$d$ = average primary clarifier depth (m).

Influent flow divided by surface area is equal to the surface overflow rate (SOR, $m^3/m^2 \cdot d$ [gpd/sq ft]):

$$\text{SOR} = Q_{PI}/A_{PC} \tag{2.6}$$

and the remaining product, $n$ times depth, can be replaced with another constant:

$$\lambda = n \cdot d \tag{2.7}$$

Where

$\lambda$ = settling constant (m/d or $m^3/m^2 \cdot d$ [ft/d or gpd/sq ft]).

Therefore, eq 2.3 becomes

$$\text{TSS}_{PE} = \text{TSS}_{non} + (\text{TSS}_{PI} - \text{TSS}_{non})e^{-\lambda/\text{SOR}} \tag{2.8}$$

Dividing both sides of eq 2.8 by $\text{TSS}_{PI}$, subtracting each side from 1, and substituting the result into eq 2.1 yields

$$E_{TSS} = [1 - (\text{TSS}_{non}/\text{TSS}_{PI})] - [1 - (\text{TSS}_{non}/\text{TSS}_{PI})]e^{-\lambda/\text{SOR}} \tag{2.9}$$

Tebbutt and Christoulas (1975) introduced a parameter in their equation development, $E_o$, equal to the TSS removal efficiency under quiescent conditions, although "quiescent conditions" were not defined. In fitting their equation to pilot-scale data, there was an inconsistency in that the estimated value for $E_o$ was greater than 1, a physical impossibility. What they missed in their equation development was the fact that the maximum removal efficiency possible, $E_{TSSmax}$, would be achieved when the primary effluent TSS concentration was equal to the nonsettleable TSS concentration as defined by eq 2.10:

$$E_{TSSmax} = 1 - (\text{TSS}_{non}/\text{TSS}_{PI}) \tag{2.10}$$

Substitution of eq 2.10 into eq 2.9 yields

$$E_{TSS} = E_{TSSmax}(1 - e^{-\lambda/\text{SOR}}) \tag{2.11}$$

Method 2540F of *Standard Methods* (APHA et al., 1998) includes a procedure for measuring the nonsettleable TSS concentration. This procedure calls for settling at least a 1-L sample for 1 hour in a container at least 9 cm (3.5 in.) in diameter and 20 cm (7.9 in.) in depth. This procedure does not address, however, the flocculation potential of whatever sample is used. Solids in primary influents are flocculent to a measurable degree. Figure 2.2 (Parker et al., 2000) shows the supernatant TSS concentration in primary effluent after 30 minutes of settling preceded by different flocculation times (at 50 rpm on a Phipps and Bird, Richmond, Virginia, stirrer). Without chemical addition, the maximum flocculation potential (i.e., the minimum supernatant TSS concentration) occurs after approximately 30 minutes of flocculation for that wastewater; with chemical addition, the minimum supernatant TSS concentration occurs much more rapidly, in fewer than 5 minutes. This example shows that the supernatant TSS concentration was reduced from approximately 110 to 62 mg/L with 32 minutes of flocculation, a significant decrease. Wahlberg et al. (1999) also discussed the flocculation potential of solids in a primary influent. An operational definition of the nonsettleable TSS concentration (i.e., the supernatant TSS concentration after 30 minutes of flocculation at 50 rpm and 30 minutes of settling) was used by Wahlberg et al. (1998), and Wahlberg (1999) noted the need for a standardized test, which includes flocculation, for measuring the nonsettleable TSS concentration.

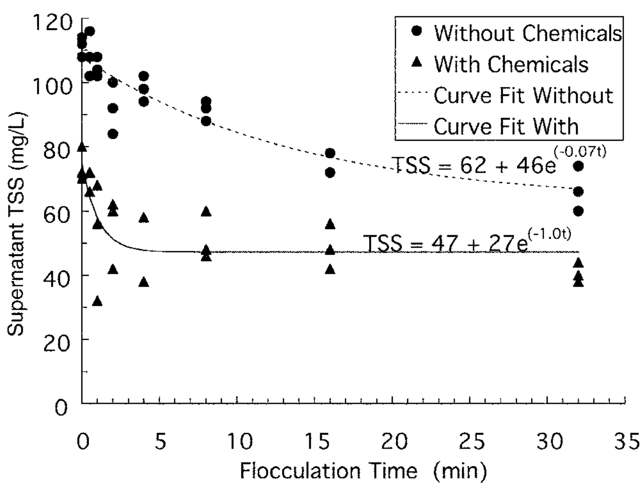


**FIGURE 2.2** Supernatant TSS concentration (after 30 min settling) as a function of flocculation time with and without chemical addition (from Parker et al., 2000).

As indicated above, of more importance to downstream biological processes than the primary effluent TSS concentration is the concentration of organic material, quantified, for purposes of this discussion, in terms of the COD concentration. Primary effluent COD is composed of soluble and particulate fractions:

$$COD_{PE} = sCOD_{PE} + pCOD_{PE} \tag{2.12}$$

Where

$COD_{PE}$ = primary effluent COD concentration (mg/L),
$sCOD_{PE}$ = primary effluent soluble COD concentration (mg/L), and
$pCOD_{PE}$ = primary effluent particulate COD concentration (mg/L).

The particulate component includes the COD associated with nonsettleable TSS and escaping settleable TSS. Defining $\Psi$ as the ratio of $pCOD_{PE}$ to $TSS_{PE}$,

$$\Psi = pCOD_{PE}/TSS_{PE} \tag{2.13}$$

Equation 2.12 becomes

$$COD_{PE} = sCOD_{PE} + \Psi TSS_{PE} \tag{2.14}$$

Substituting eq 2.8 into eq 2.14 yields

$$COD_{PE} = sCOD_{PE} + \Psi[TSS_{non} + (TSS_{PI} - TSS_{non})\, e^{-\lambda/SOR}] \tag{2.15}$$

Under most operational conditions, the primary effluent soluble COD concentration ($sCOD_{PE}$) is equal to the primary influent soluble COD concentration ($sCOD_{PI}$). Recognizing that the primary influent nonsettleable COD ($COD_{non}$) is composed of the soluble COD ($sCOD_{PI}$) plus the particulate COD associated with nonsettleable TSS ($\Psi \cdot TSS_{non}$ or $pCOD_{non}$) and that the particulate COD associated with settleable TSS ($pCOD_{set}$) is equal to the difference between the primary influent COD and the primary influent nonsettleable COD ($COD_{PI} - COD_{non}$), eq 2.15 can be rewritten as

$$COD_{PE} = COD_{non} + (COD_{PI} - COD_{non})\, e^{-\lambda/SOR} \tag{2.16}$$

Equation 2.16 shows that the total primary effluent COD concentration is composed of a nonsettleable fraction ($COD_{non}$) and the particulate fraction associated with escaping settleable TSS [i.e., $(COD_{PI} - COD_{non})e^{-\lambda/SOR}$].

Substitution of eq 2.16 into eq 2.2 yields

$$E_{COD} = 1 - \{[COD_{non} + (COD_{PI} - COD_{non})\, e^{-\lambda/SOR}]/COD_{PI}\} \tag{2.17}$$

Defining the maximum COD removal efficiency, $E_{CODmax}$, similar to the maximum TSS removal efficiency (eq 2.10),

$$E_{CODmax} = 1 - (COD_{non}/COD_{PI}) \tag{2.18}$$

Equation 2.17 simplifies to

$$E_{COD} = E_{CODmax}(1 - e^{-\lambda/SOR}) \tag{2.19}$$

In summary, the important factors affecting primary clarifier performance can be seen directly from eqs 2.9 and 2.17:

1. The nonsettleable TSS concentration,
2. The influent TSS concentration,
3. The settling characteristics of the settleable solids (indirectly quantified by $\lambda$),
4. The surface overflow rate,
5. The soluble COD concentration (should be the same in the primary influent and effluent), and
6. The ratio of pCOD (or $BOD_5$) to TSS in the primary effluent (i.e., $\Psi$).

Interestingly, all of these factors are characteristics of the wastewater. Good primary clarifier design, therefore, begins with a careful study of the wastewater characteristics under all anticipated flow conditions. Moreover, in identifying detailed design elements (e.g., configuration, inlet and outlet design, size, depth, sludge-withdrawal mechanism, and scum collection/withdrawal mechanism), the challenge for the designer is to consistently produce a primary clarifier effluent with a TSS concentration equal to the nonsettleable TSS concentration and a COD concentration equal to the nonsettleable COD concentration (i.e., the soluble COD concentration plus the particulate COD concentration associated with the nonsettleable TSS). This performance goal takes empiricism out of primary clarifier design.

**CASE STUDIES.** With the extensive equation development just presented, it may be difficult to appreciate the usefulness of this approach. As stated previously, process capacity of primary clarifiers upstream of biological treatment depends on the oxidative capacity of downstream secondary facilities. The reader will appreciate that the performance of primary clarifiers is not fixed, as it depends on many variables. In whatever way those variables are affecting performance, the process capacity of primary clarifiers is exceeded when the biological process can no longer fully oxidize the COD load discharged from the primary clarifiers. To be able to predict that load, then, is the key to quantifying capacity. As an illustration of the use of this equation development, two case studies are given. The first uses the COD performance equation (eq 2.16), calibrated using results from a Water Environment Research Foundation (WERF) primary clarifier study (Wahlberg, 2004), to show the effect of additional primary clarifiers on the COD concentration in the influent to a downstream activated sludge plant in Oregon. The second uses historical data to estimate the nonsettleable TSS concentration, $TSS_{non}$, and the settling parameter, $\lambda$. The "calibrated" performance equation was then used in a facility planning effort for the expansion of a 33 690-$m^3/d$ (8.9-mgd) wastewater treatment plant in Northern California.

The plant in Oregon has a design average dry weather flow capacity of 185 465 $m^3/d$ (49 mgd). There are four 40-m-diam (135-ft-diam) primary clarifiers. The plant experiences significant peak flows. At issue was the reduction in COD concentration to the downstream activated sludge plant that would occur with additional primary clarifiers. The plant participated in the WERF primary clarifier study. For 1 full year, approximately every sixth day, plant staff measured $TSS_{PI}$, $TSS_{non}$, $TSS_{PE}$, $COD_{PI}$, $COD_{non}$, and $COD_{PE}$ around one of the primary clarifiers. Equation 2.16 was fit to the COD and SOR data to obtain an estimate of $\lambda$, 106 $m^3/m^2 \cdot d$ (2593 gpd/sq ft). This estimate, in turn, was used to predict the primary effluent COD concentration at increasing flows for four, six, and eight primary clarifiers. In this comparison, the $COD_{PI}$ and $COD_{non}$ concentrations were set equal to 441 and 244 mg/L, respectively. The results are shown in Figure 2.3. As can be seen in that figure, the expense of building four additional primary clarifiers lowers the COD concentration to the activated sludge system by only approximately 40 mg/L.

In the original design documents for the 33 690-$m^3/d$ (8.9-mgd) plant in Northern California, the performance of the two 29-m-diam (95-ft-diam) primary clarifiers was stated as "it is assumed the primary clarifiers will remove 60 and 25 percent of the incoming TSS and $BOD_5$, respectively". The operations staff runs both primary clarifiers in the winter and one in the summer. "Winter" is November through May; "summer" is May through November. May and November can either be wet or dry so were included in both seasons for this analysis. Data were analyzed for the period of January 1995 through December 2002.

Figures 2.4 and 2.5 show TSS removal efficiency for the summer and winter periods, respectively, as a function of SOR. As can be seen in both of these figures, surface overflow rate cannot be used, by itself, to predict primary clarifier perfor-

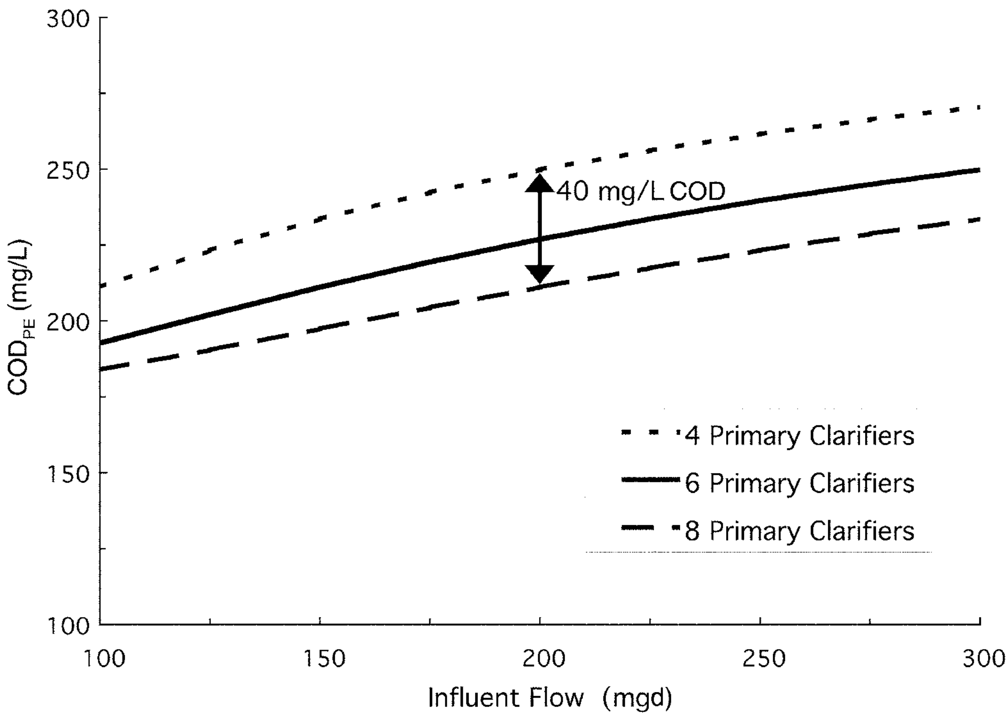


**FIGURE 2.3** The effect of additional primary clarifiers and flow on the primary effluent COD concentration ($COD_{PE}$) at a plant in Oregon ($m^3/h = 158 \times$ mgd).

mance. Figures 2.6 and 2.7, in contrast, show TSS removal efficiency for the summer and winter periods, respectively, as a function of the influent TSS concentration ($TSS_{PI}$). As can be seen from these figures, although there is still scatter in the data, a clearer relationship is seen than is apparent in Figures 2.4 and 2.5. This is because $TSS_{PI}$ is more prominent than SOR in the TSS performance equation given above, eq 2.9.

The raw operational data collected at the plant during this period were used to estimate the magnitude of $TSS_{non}$ and $\lambda$. This was accomplished by calculating $E_{TSS}$ using a number of different combinations of $TSS_{non}$ and $\lambda$ ($TSS_{non}$ was varied between 10 and 100 mg/L; $\lambda$ was varied between 41 and 122 $m^3/m^2 \cdot d$ [1000 and 3000 gpd/sq ft]) and the $TSS_{PI}$ and SOR recorded for each day, calculating the squared error between this estimate of $E_{TSS}$ and the observed $E_{TSS}$ for each day, and generating a three-dimensional surface by plotting $TSS_{non}$ as a function of $\lambda$ as a function of the

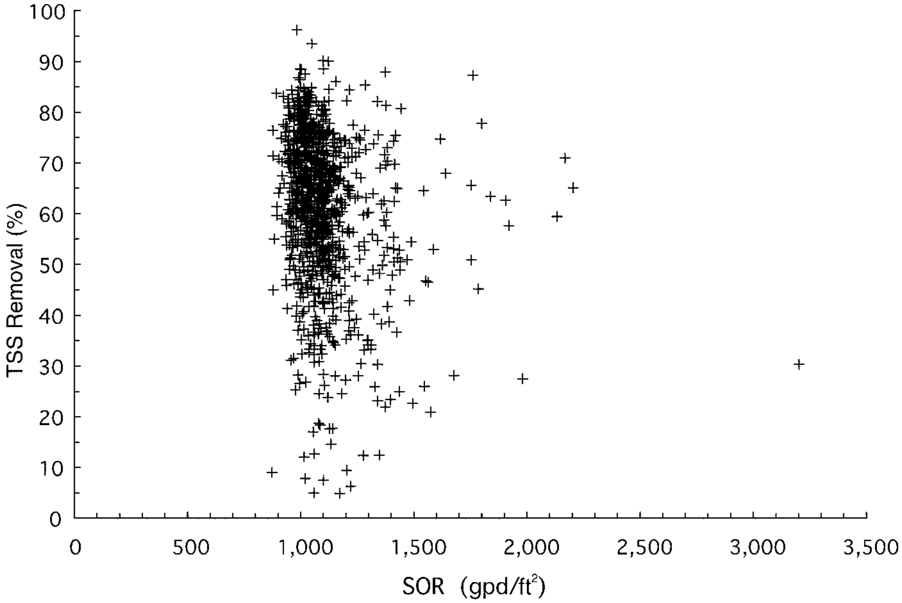


**FIGURE 2.4** Full-scale summer plant operating data: TSS removal efficiency as a function of the surface overflow rate (SOR), 1995-2002; typically, one primary clarifier in service ($m^3/m^2{\cdot}d = 0.04075 \times$ gpd/sq ft.)

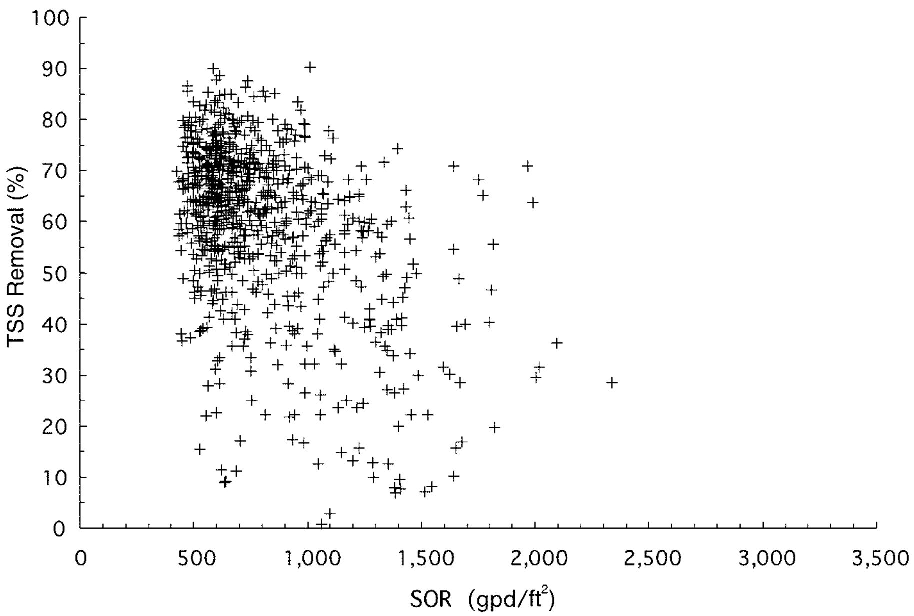


**FIGURE 2.5** Full-scale winter plant operating data: TSS removal efficiency as a function of the surface overflow rate (SOR), 1995-2002; typically, two primary clarifiers in service ($m^3/m^2{\cdot}d = 0.04075 \times$ gpd/sq ft).

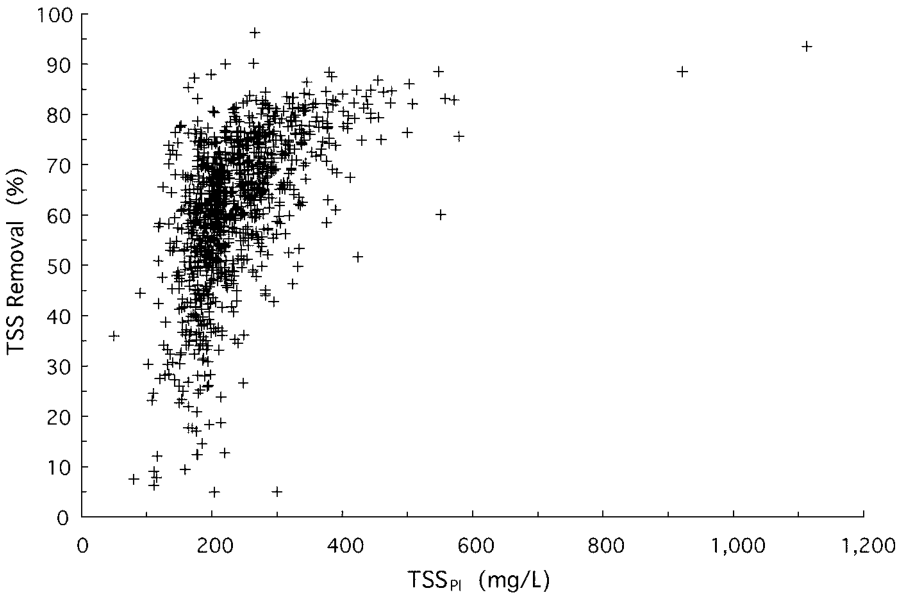


**FIGURE 2.6** Full-scale summer plant operating subdata: TSS removal efficiency as a function of the influent TSS concentration ($TSS_{PI}$), 1995-2002. Surface overflow rate varied between 873 and 3203 gpd/sq ft (35.6 and 130 $m^3/m^2 \cdot d$).

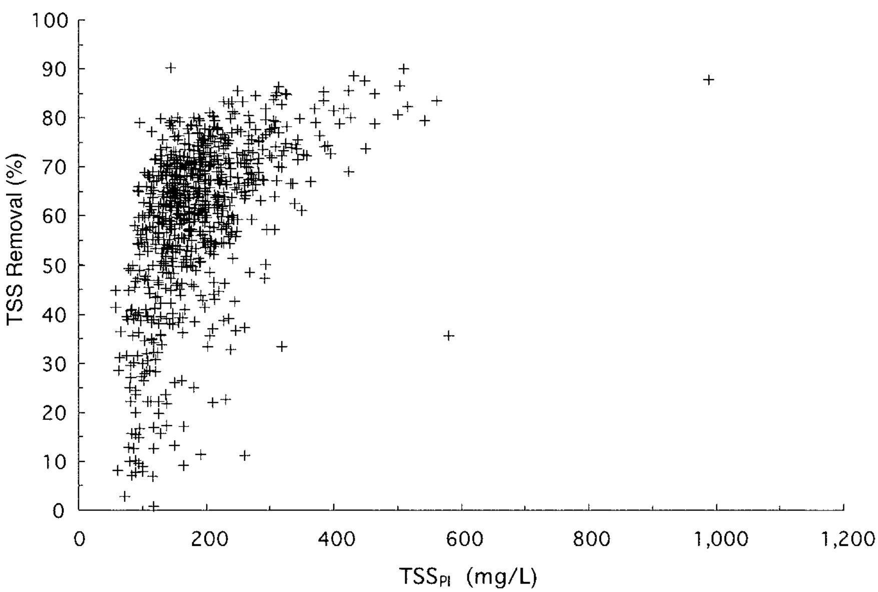


**FIGURE 2.7** Full-scale winter plant operating data: TSS removal efficiency as a function of the influent TSS concentration ($TSS_{PI}$), 1995-2002. Surface overflow rate varied between 427 and 2338 gpd/sq ft (17.4 and 95.3 $m^3/m^2 \cdot d$).

sum of squared errors (not shown). From this surface, the combination of $TSS_{non}$ and $\lambda$ that gave the minimum sum of squared errors was identified. Using the daily data for $TSS_{PI}$ and SOR and these selected values for $TSS_{non}$ and $\lambda$, the removal efficiency was calculated and is plotted over the raw data given in Figures 2.6 and 2.7 as Figures 2.8 and 2.9.

As can be seen in Figures 2.8 and 2.9, the observed performance data are well described by eq 2.10 and the estimates for $TSS_{non}$ and $\lambda$ of 70 mg/L and 102 $m^3/m^2{\cdot}d$ (2500 gpd/sq ft) and 60 mg/L and 122 $m^3/m^2{\cdot}d$ (3000 gpd/sq ft) for the summer and winter conditions, respectively. Anaerobic activity occurring in sewers, elevated by hot summer temperatures, is likely the reason for the difference between the summer and winter estimates.

Taking it one step further, the summer activated sludge maximum COD loading is 15 422 kg/d (34 000 lb/d). With a design $COD_{PI}$ of 505 mg/L and a

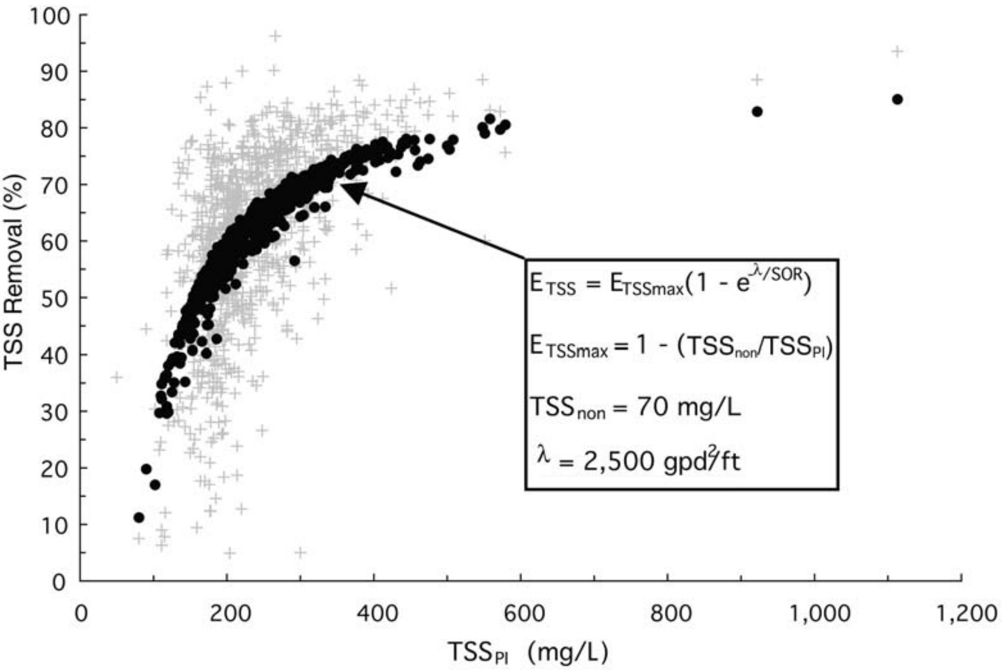


**FIGURE 2.8** Data calculated using Equation 2.10, daily summer $TSS_{PI}$ and SOR measurements, overlain on data from Figure 2.6 ($m^3/m^2{\cdot}d = 0.04075 \times$ gpd/sq ft).

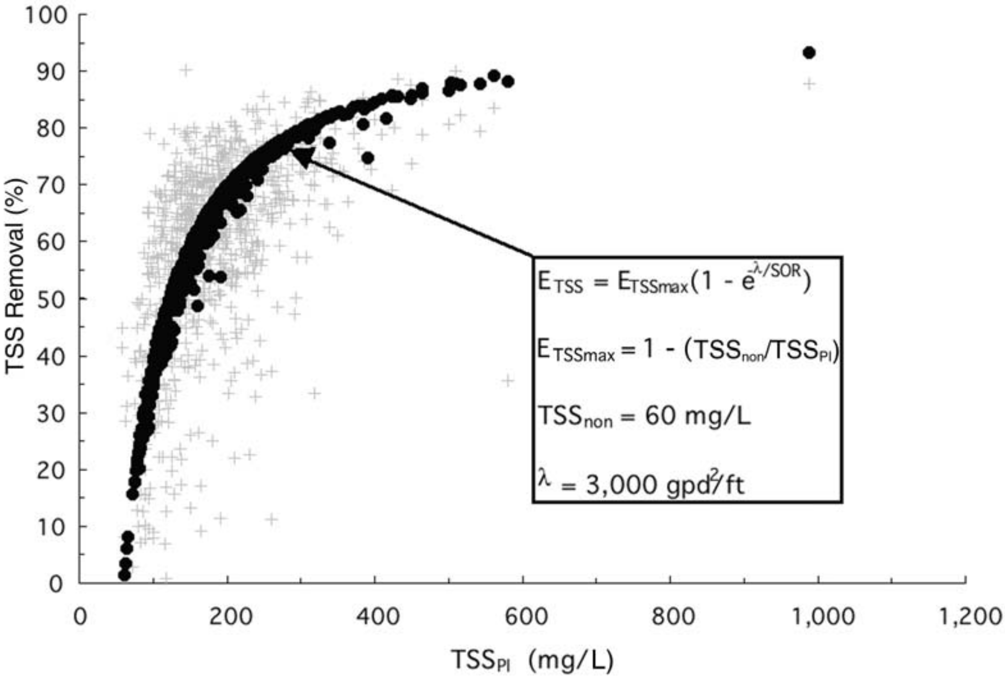


**FIGURE 2.9** Data calculated using Equation 2.10, daily winter $TSS_{PI}$ and SOR measurements, overlain on data from Figure 2.7 ($m^3/m^2 \cdot d = 0.04075 \times$ gpd/sq ft).

$COD_{non}$ concentration 330 mg/L (determined from the $TSS_{non}$ concentration [70 mg/L] and the soluble $COD_{PI}$ concentration of 246 mg/L), the λ estimate (102 $m^3/m^2 \cdot d$ [2500 gpd/sq ft]) was used to predict the $COD_{PE}$ concentration (using eq 2.16), which, in turn, was used to calculate the COD loading to the activated sludge plant. Figure 2.10 shows the results. At a limiting loading to the activated sludge system of 15 422 kg COD/d (34 000 lb COD/d), the capacity of the primary clarifiers is 42 203 $m^3$/d (11.15 mgd). Being able to actually predict $TSS_{PE}$ and $COD_{PE}$ (or $BOD_{PE}$) is a huge step forward from having to assume TSS and BOD removal efficiencies in primary clarifiers.

## CHEMICALLY ENHANCED PRIMARY TREATMENT

Removal of solids from raw wastewaters in primary clarifiers depends on gravity separation. Because solids in raw wastewaters vary substantially in size, shape, and density, gravity separation should theoretically consider the settling velocity distribution of all of the different solids. Within a practical time scale, however, raw wastewater TSS can be considered as either settleable or nonsettleable. Similarly, the total COD (or BOD) in primary influents is either soluble, particulate associated with settleable TSS, or particulate associated with nonsettleable TSS; nonsettleable COD is composed of the soluble COD and the particulate COD associated with nonsettleable TSS. The process objective of chemically enhanced primary treatment is to produce an effluent, with the addition of chemicals, lower in TSS and COD than the nonsettleable TSS and COD, respectively, measured without the addition of chemicals. The history of chemically enhanced primary treatment recently has been discussed in a

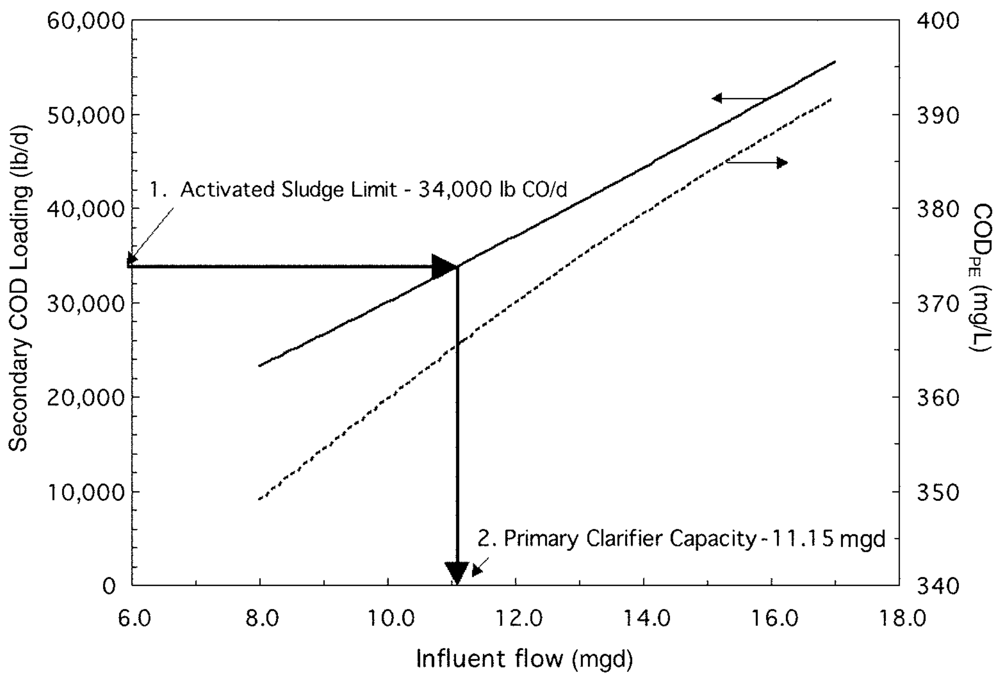


**FIGURE 2.10** Capacity determination of primary clarifiers. $COD_{PE}$ curve developed using Equation 2.16 ($COD_{PI}$ = 505 mg/L, $COD_{non}$ = 330 mg/L, $\lambda$ = 2,500 gpd/sq ft). COD loading curve developed by multiplying $COD_{PE}$ by flow by 8.34 lb/gal. Primary clarifier capacity is defined by the flow corresponding to where the activated sludge COD limitation (1) intersects the COD loading curve (2). ($m^3/m^2 \cdot d$ = 0.04075 $\times$ gpd/sq ft, $m^3/h$ = 158 $\times$ mgd, kg/d = 0.454 $\times$ lb/d).

series of articles in which numerous references are given (Harleman and Murcott, 2001a, 2001b; Parker et al., 2001). Chemically enhanced primary treatment is discussed in more detail in Chapter 3. Although not specifically "enhanced primary treatment", chemical addition to primary clarifiers also is done to remove phosphorus for nutrient control, heavy metals to meet toxicity requirements, and hydrogen sulfide to lower odor emissions. Chemical addition also can be used for CSO abatement and in conjunction with effluent blending approaches related to ocean dischargers with 301H waivers.

Typically, iron or aluminum salts (e.g., ferric chloride [$FeCl_3$], or alum [$Al_2(SO_4)_3$] are added in conjunction with a polymer to improve TSS removal. The fact that the two curves in Figure 2.2 (Parker et al., 2000) are asymptotic to different values (62 mg/L in the case with no chemicals added; 47 mg/L in the case with chemicals added) demonstrates that the addition of chemicals decreases the nonsettleable TSS concentration. Moreover, although all of the samples in Figure 2.2 were settled for the same amount of time (i.e., 30 minutes), the fact that the curve with chemicals approaches the asymptotic nonsettleable TSS concentration substantially more rapidly than the curve with no chemicals demonstrates that the addition of chemicals increases the settling velocity of the settleable TSS. While it is easily understood why reducing the nonsettleable TSS concentration would enhance the performance of primary clarification with chemical addition, enhanced performance by increasing the settling velocity of settleable TSS may not be as intuitive. Data from the WERF study suggest that chemical addition increases the settling constant in eq 2.8 (i.e., $\lambda$) with the net result that fewer settleable TSS would be lost in the effluent at a given surface overflow rate than without chemical addition, thereby improving performance. The effect of chemical addition on $\lambda$ is shown in Figures 2.11 and 2.12: $\lambda$ was increased dramatically (from 44 to 192 $m^3/m^2{\cdot}d$ [1078 to 4714 gpd/sq ft]) on the same primary influent sample with chemical addition. Unfortunately, this one test was the only side-by-side comparison of the effect of chemical addition on $\lambda$ performed during the study.

Often, designers provide for the use of chemically enhanced primary treatment during high flow events. Chemical addition affords designers and operators the ability to manipulate $\lambda$ . By increasing $\lambda$ using chemical addition, as suggested in the previous paragraph, the same performance can be achieved at higher flows. Figure 2.13 shows the effect $\lambda$ has on primary clarifier performance (i.e., $TSS_{PE}$) at increasing surface overflow rates (i.e., higher flows) for the hypothetical case in which the $TSS_{PI}$ and $TSS_{non}$ concentrations are 280 and 60 mg/L, respectively. As can be seen, the larger $\lambda$ is, the less effect surface overflow rate has on primary clarifier performance.

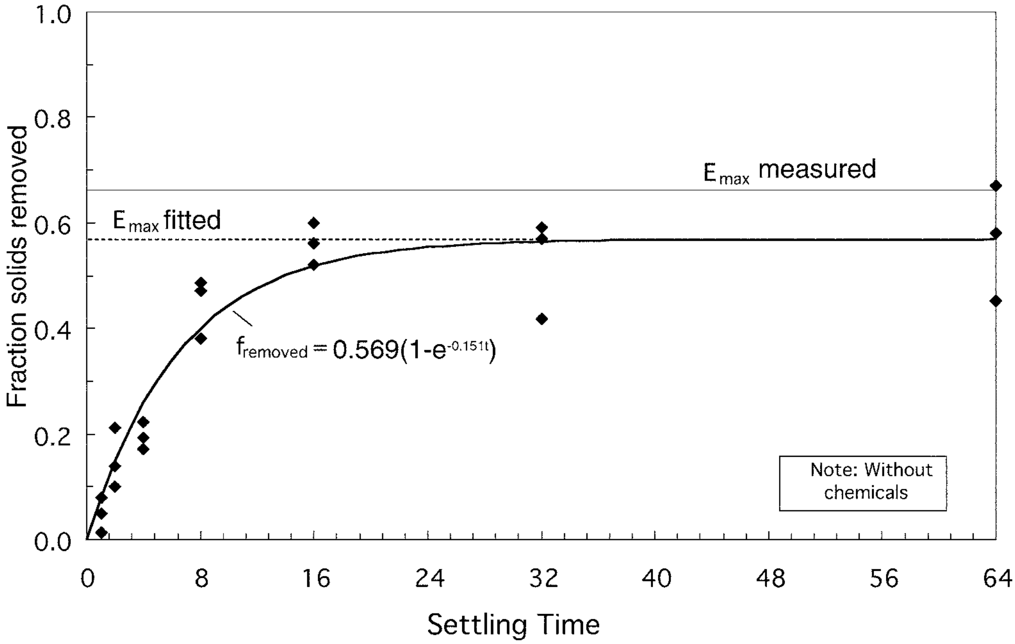


**FIGURE 2.11** Results from Kemmerer (Wildlife Supply Company, Buffalo, New York) settling tests from WERF study. Same primary influent sample as Figure 2.12, without chemical addition; λ calculated to be 1,078 gpd/sq ft (43.9 $m^3/m^2 \cdot d$). (Reprinted with permission from Water Environment Research Foundation (2004) *Determine the Affect of Individual Wastewater Characteristics and Variances on Primary Clarifier Performance)*.

As shown, a primary effluent TSS concentration of 100 mg/L is achieved at 24.0, 47.9, 71.3, or 95.8 $m^3/m^2 \cdot d$ (590, 1175, 1750, or 2350 gpd/sq ft) depending on if λ is equal to 40.7, 81.5, 122, or 163 $m^3/m^2 \cdot d$ (1000, 2000, 3000 , or 4000 gpd/sq ft), respectively.

There are many chemicals on the market used for chemically enhanced primary treatment. Often, chemicals are used in concert, ferric chloride and anionic polymer, for example. Because every wastewater is different and cost is always a consideration, the identification of the best chemical or chemicals to use requires careful analysis, typically beginning with jar testing and sound experimental design. In the past, design engineers and operators have focused primarily on the extent of the flocculation reaction achievable through chemical addition by measuring the clarity of the supernatant in jar tests. While extent is certainly important, the rate of the flocculation reaction also should be considered, especially when

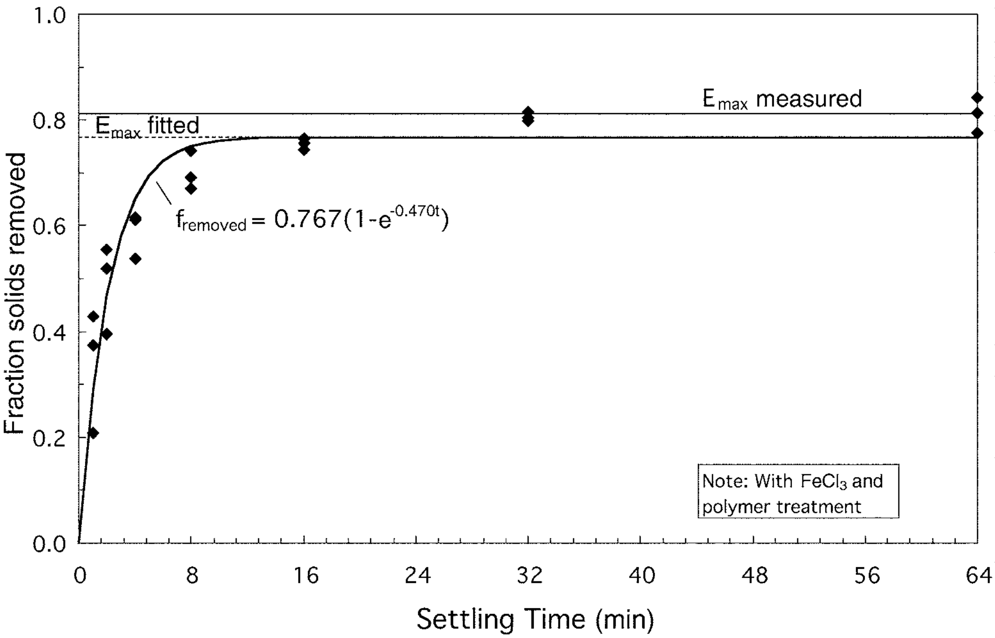


**Figure 2.12** Results from Kemmerer (Wildlife Supply Company, Buffalo, New York) settling tests from WERF study. Same primary influent sample as Figure 2.11, with chemical addition; $\lambda$ calculated to be 4,714 gpd/sq ft (192 $m^3/m^2 \cdot d$). (Reprinted with permission from Water Environment Research Foundation (2004) *Determine the Affect of Individual Wastewater Characteristics and Variances on Primary Clarifier Performance*).

chemically enhanced primary treatment is being considered as a retrofit in an existing facility, constrained by the pipe, channel, and tank sizes available.

When evaluating the potential of chemically enhanced primary treatment, the approach of Wahlberg et al. (1994) should be used to measure both the extent and rate of solids removal by flocculation. In this approach, the flocculation time is varied after the chemicals are injected. After the prescribed flocculation time, the sample is allowed to settle for 30 minutes and the supernatant is tested for TSS. Supernatant TSS is plotted as a function of flocculation time and a decreasing exponential curve is fit to the data. Despite its simplicity, the decreasing exponential function, based on the work of Parker et al. (1970), is grounded in floc aggregation and breakup theory. The equation is:

$$TSS_{super} = TSS_{non} + (TSS_o - TSS_{non})e^{-\phi t} \tag{2.20}$$

Where

$TSS_{super}$ = supernatant TSS concentration (mg/L),
$TSS_{non}$ = curve-fitting parameter corresponding to the nonsettleable TSS concentration (mg/L),
$TSS_o$ = curve-fitting parameter corresponding to the initial supernatant TSS concentration with no flocculation (mg/L),
$\varphi$ = flocculation rate parameter (1/min), and
$t$ = flocculation time (min).

Wahlberg et al. (1999) performed a series of jar tests on raw wastewater with and without chemical flocculation aids in which the flocculation time was varied. Equation 2.20 was fit to the supernatant TSS data using a nonlinear curve-fitting

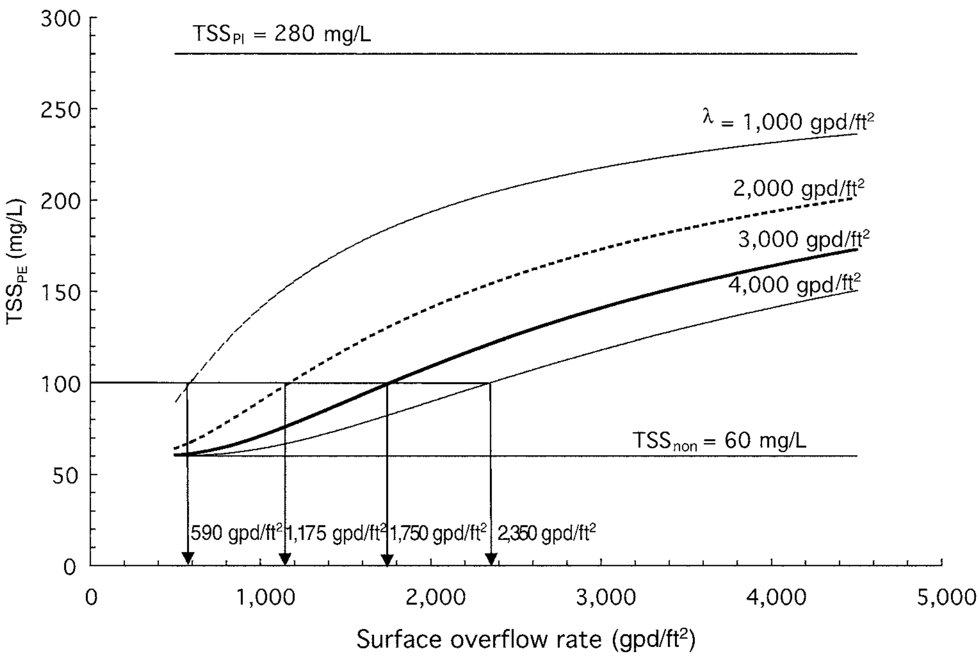


**Figure 2.13** The effect of $\lambda$ on primary clarifier performance at increasing flows for the hypothetical case in which the $TSS_{PI}$ and $TSS_{non}$ concentrations are 280 and 60 mg/L, respectively.

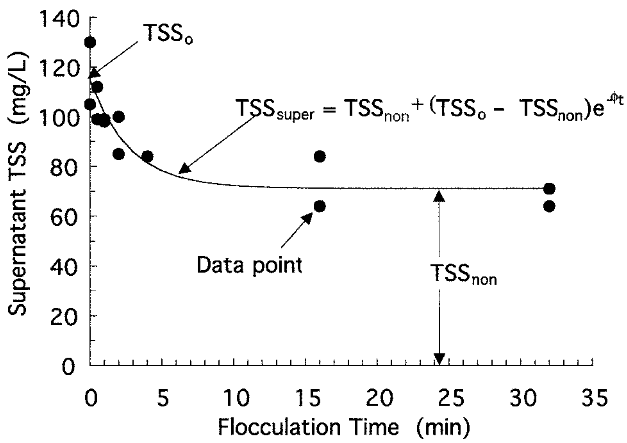


**Figure 2.14** Fit of Equation 20 to jar test data in which flocculation time was varied [modified from Wahlberg et al. (1999)].

technique. An example of the data collected for one chemical treatment is shown in Figure 2.14. Shown with the data is the fit of eq 2.20. Also shown in this figure are arrows indicating the physical meaning of the parameters $TSS_{non}$ and $TSS_o$ (71 and 115 mg/L, respectively). The rate of change of the slope of the curve, equal to the rate of the flocculation reaction, is defined by $\phi$ (0.36 1/min). It is important to note about eq 2.20 that $TSS_{non}$ and $\phi$ reflect the extent and rate of the flocculation reaction, respectively.

Jar tests as they are typically performed provide only an estimate of the extent of the flocculation reaction. This approach (as exemplified in Figure 2.14) also affords an estimate of the rate of the flocculation reaction as quantified by the $\phi$ parameter.

## DESIGN CONCEPTS AND CONSIDERATIONS

As discussed in the previous sections, it is the relative amounts of settleable versus nonsettleable TSS and soluble versus particulate COD in the influent to a primary clarifier that dictate the potential maximum performance that may be achieved. Primary clarifier design, therefore, has to begin with a characterization of the wastewater that is to be treated. Once the maximum performance is identified, ensuring that the primary clarifier performs to that level requires the design engineer to focus on maximizing the flocculation potential of raw wastewater solids and providing as quiescent conditions as possible. From an operational perspective, under most situations, the design engineer should also minimize the possibility of any biological activity occurring in the primary clarifier. All decisions, then, having to do with configuration, depth, inlet design, sludge collection and withdrawal, scum collection and withdrawal, and effluent discharge are made to maximize flocculation, minimize unwanted hydraulic currents (i.e., achieve ideal flow as much as is practical), and minimize biological activity.

Minimal biological activity is not always desired. In biological phosphorus removal plants, primary sludge fermentation, performed in the primary clarifier, is sometimes used to purposefully generate volatile fatty acids. This is, by far, the exception rather than the rule. Designing and operating primary clarifiers as sludge fermentation units is challenging and requires advanced design concepts and considerations that are beyond the scope of this document.

**WASTEWATER CHARACTERIZATION.** To ensure the performance of any primary clarifier, the design engineer must know the characteristics of the wastewater that is to be treated. In most instances, the wastewater already exists so it can be sampled. The TSS and total COD should be measured on samples that are tested for nonsettleable TSS and COD. One method (Larsson. 1986) for characterizing settling properties has been mentioned in Chapter 1 and detailed in Appendix A. This can be used with the approach described in Chapter 1. Alternatively, Wahlberg et al. (1998) suggested that the supernatant TSS and COD concentrations after 30 minutes of flocculation (i.e., 50 rpm on a Phipps and Bird stirrer using a 2-L square flocculation jar) and 30 minutes of settling be operationally defined as nonsettleable TSS and COD, respectively. These data can be used with eqs 2.10 and 2.18 to calculate maximum TSS and COD removal efficiencies, respectively. Because these tests are rarely, if ever, performed by plant personnel, this will require additional sampling. Table 2.1 presents characterization data from the WERF study on primary clarifier performance (Wahlberg, 2004). The results reported in the table are averages collected over 1 year at eight of the ten municipal wastewater treatment plants participating in the study.

As can be seen in the equation development above, the surface overflow rate does affect performance, but only in the removal of settleable TSS and COD. Because maximum flows occur during wet weather events, wastewater characterization sampling should include some storm events. Little is known about the relative amounts of settleable versus nonsettleable TSS and soluble versus particulate COD during

**TABLE 2.1** Typical wastewater characterization from municipal plants participating in a WERF primary clarifier study. Reprinted with permission from Water Environment Research Foundation (2004) *Determine the Affect of Individual Wastewater Characteristics and Variances on Primary Clarifier Performance.*

| Plant | $TSS_{PI}$ (mg/L) | $COD_{PI}$ (mg/L) | Fraction $TSS_{non}$[a] | Fraction $sCOD_{PI}$ | Fraction $COD_{non}$[b] |
|---|---|---|---|---|---|
| 1 | 186 | 399 | 0.33 | 0.39 | 0.59 |
| 2 | 184 | 423 | 0.26 | 0.42 | 0.64 |
| 3 | 210 | 463 | 0.25 | 0.32 | 0.53 |
| 4 | 267 | 555 | 0.24 | 0.36 | 0.52 |
| 6 | 508 | 864 | 0.24 | 0.30 | 0.45 |
| 7 | 287 | 612 | 0.27 | 0.33 | 0.50 |
| 8 | 242 | 452 | 0.23 | 0.39 | 0.55 |
| 10 | 337 | 686 | 0.20 | 0.27 | 0.43 |

[a] $E_{TSSmax}$ = 100%(1 – fraction $TSS_{non}$).
[b] $E_{CODmax}$ = 100%(1 – fraction $COD_{non}$).

storm events, but it is the flows during these events that typically fix the size of the primary clarifiers based on a surface overflow rate calculation. It behooves the design engineer (and the regulator) to understand what kinds of removals are possible at these high flows given the nature of the solids in the wastewater and not on assumptions based on surface overflow rates. Moreover, little is known about the variability of the $\lambda$ parameter, which quantifies the effect of surface overflow rate on settleable TSS and COD removals. There is currently no standardized test to measure $\lambda$, and this is a research need. Wahlberg et al. (1998) used results from a series of settling tests performed in a 4.1-L Kemmerer sampler (Wildlife Supply Company, Buffalo, New York) to estimate $\lambda$, but there are some scale-up effects that need to be considered.

**CONFIGURATION AND DEPTH.** Circular and rectangular primary clarifiers are the most common. "Squircle" primary clarifiers—square tanks with circular sludge collection mechanisms—have been used. Because of unwanted currents and sludge buildup in the corners, however, this configuration can lead to poor hydraulics and biological activity so should be avoided. Other exotic configurations have been and likely will continue to be proposed. Whatever the configuration, it should be the responsibility of the design engineer to evaluate it in terms of maximizing the flocculation potential, providing ideal hydraulics, and minimizing biological activity. Design considerations for primary clarifiers and secondary (or tertiary) clarifiers are much different because of settling properties of the solids and overall performance objectives, so they should be evaluated separately.

Several choices related to performance must be made in the design of circular primary clarifiers, including

- Inlet stilling well size and configuration,
- Floor slope,
- Effluent launder positioning (inboard, outboard, or an intermediate location away from the sidewall),
- Scraper arrangement (conventional versus spiral rake, single rake versus dual rake),
- Scum withdrawal (localized, partial radius, or full radius),
- Amount of freeboard for wind protection,
- Covered versus uncovered, and
- Constant-speed versus variable-speed sludge collectors.

Other non-performance-related choices include materials of construction, bridge arrangement (half bridge versus full bridge), type of drive (electric motor versus hydraulic), and coating of equipment and structures. Instrumentation and controls also must be selected.

Rectangular primary clarifier units offer similar choices for the designer to consider. These include

- Inlet configuration (unbaffled, baffled, target box, or inlet tee),
- Sludge hopper size and arrangement (cross-collector or stationary sludge header or multiple withdrawal pipes),
- Effluent launder requirements (end wall weir trough, multiple intermediate weir troughs, submerged pipe, etc.),
- Scum withdrawal (conventional rotating pipe, downward opening weir gates, or other specialized vendor furnished packages),
- Collector drive arrangement (multiple collectors connected to a single drive versus a single drive dedicated to each collector),

- Covered versus uncovered,
- Type of collector (chain and flight, traveling bridge with plows, and traveling bridge with hydraulic suction can be used),
- Flight depth and spacing (more flights may improve performance but will result in increased operations and maintenance requirements),
- Flight speed (faster speeds may improve performance but will wear out faster), and
- Constant-speed versus variable-speed sludge collectors.

Other non-performance-related choices include materials of construction, type of drive, and coating of equipment and structures. Instrumentation and controls also must be selected.

Whatever the configuration, it is the responsibility of the design engineer to evaluate it in terms of maximizing the flocculation potential, providing ideal hydraulics, and minimizing biological activity. The design engineer must consider life-cycle costs, site layout/space availability, interchangeability with existing units, the presence (or absence) of upstream preliminary treatment systems, overall treatment objectives, odor control requirements, reliability, and efficiency of existing units during the selection of a design arrangement. For example, it may be acceptable to leave existing inefficient units "as is" if downstream unit processes have the capacity to provide proper treatment. Conversely, it may be appropriate to upgrade existing designs for improved performance to alleviate the need for expansion of downstream unit processes.

Side-by-side performance comparisons of circular versus rectangular primary clarifiers have not been published, so a recommendation of one over the other cannot be made. Typical average surface overflow rates of 24.4 to 48.9 $m^3/m^2 \cdot d$ (600 to 1200 gpd/sq ft), with peak surface overflow rates of 102 to 122 $m^3/m^2 \cdot d$ (2500 to 3000 gpd/sq ft) have been successfully used for both circular and rectangular primary clarifier designs (see Table 3.29 in Chapter 3). There are other considerations of each application that would dictate circular over rectangular or vice versa, such as space and length of primary sludge pump suction lines. In the predesign of a primary clarifier expansion project, a large plant in northern California with existing rectangular primary clarifiers recently opted for circular units (Brown and Caldwell, 2003). It was the belief of the operations and maintenance staff at the plant and other interviewed owners and operators of rectangular and circular units that the operations and maintenance requirements of rectangular units exceed those of circular units. At the subject plant, one failed link in the existing rectangular primary clarifiers was enough to take out one-quarter of the primary clarifier capacity.

In the past, a minimum side water depth of 2 m (7 ft) was cited for both circular and rectangular primary clarifier designs. Currently, it is more common to use minimum side water depths of 3 to 3.6 m (10 to 12 ft). The choice of depth must consider climate and wastewater temperature. The improved performance that could result from increased depth may be lost if the extra hydraulic detention time provided by the additional depth creates septic conditions (i.e., biological activity), which, in turn, results in floating sludge. This is more of a consideration in warm climates with warm wastewater. Prolonged detention times are to be avoided as well with industrial wastewaters that are prone to septicity because of their rapid biodegradability.

Because sludge storage during storm events is not as much of an issue with primary clarifiers as it is with secondary clarifiers, deciding on a depth also must consider how depth affects the hydraulics of a primary clarifier, particularly at high flows. While there has been considerable material published evaluating secondary clarifier hydraulics with computational fluid dynamic models, relatively little has been published regarding how depth affects primary clarifier hydraulics (an exception is Gerges et al., 1999). While many operators will argue that poor removal efficiencies occur during high flows, there are no data to differentiate between hydraulic problems and changing wastewater characteristics. More research is needed.

**FLOW SPLITTING.** Historically, the effect of momentum and turbulence on flow splitting was not always recognized. Many primary clarifiers were installed using effluent weirs to balance the flow into multiple tanks. Momentum and turbulence can cause more flow (and solids) to be "forced" into one tank than another adjacent identical unit, even though weir elevations are identical.

Upflow distribution structures with fixed weirs can be used to provide precise, identical flow to multiple units. If units have different surface areas, weir lengths are adjusted so that they are proportional to surface area. If clarifiers have different side water depths, then weir lengths are adjusted so that they are proportional to volume. In general, the upflow velocity in the flow splitting box should be less than 0.3 m/s (1.0 ft/sec) at peak flow to maintain nonturbulent water surface conditions. Also, adequate submergence must be provided in the structure between the top of the inlet conduit and the weir to dissipate energy. A depth of two to three times the diameter (or height) of the inlet conduit will be suitable if peak flow velocity entering the structure is less than 1 m/s (3.5 ft/sec). Higher inlet velocities could require larger, deeper distribution structures.

Using gates or valves to control flow into clarifiers can also be used. It is essential to provide automated actuators on the gates or valves and an accurate flow meter with a feedback signal to modulate valve position for this method to function properly.

Finally, it may be possible to keep nonoptimal flow splitting if downstream treatment units can tolerate less than optimal primary clarifier performance. Invariably, the primary clarifiers receiving lower flows and loads will produce better effluent, offsetting to some degree the poorer quality effluent exiting the units receiving higher flows and loads. This may be a better solution than introducing extra head loss into a facility, which could reduce the hydraulic capacity or produce other detrimental effects on upstream unit processes.

Computational fluid dynamics (CFD) modeling is often used to optimize modern flow splitting designs. Refer to Chapter 6 for more details regarding the capabilities of CFD models.

**INLET DESIGN.** Energy dissipation is the main objective in designing a primary clarifier inlet. Typically, this is accomplished using a rotating circular feed well with circular primary clarifier designs and several different types of baffle configuration in rectangular primary clarifier designs. These devices are generally intended to break up high-velocity currents and prevent flow jets from traveling toward the effluent withdrawal area.

Because the rate and extent of the flocculation reaction is dependent on, among other variables, the concentration of particles to be flocculated, the design engineer must ensure optimum conditions for flocculation at the inlet where the concentration of solids is highest. While much has been done to improve inlet design in secondary clarifiers to promote flocculation, little has been done with regards to primary clarifiers. In essence it is up to the design engineer to provide an inlet environment that gives a mixing intensity that promotes flocculation and maximizes the solids concentration for the right amount of time. In activated sludge secondary clarifier design, the "right" mixing intensity is a root-mean-square velocity gradient on the order of 30 to 70 $s^{-1}$ and the "right" detention time is 20 minutes.

Flocculation has historically not been given much consideration in primary clarifier design. Primary clarifiers typically have influent TSS concentrations ranging from 200 to 400 mg/L, with approximately 20% of those solids being inert. Still, as demonstrated in Figures 2.2 and 2.14, flocculation of primary influent solids is a real phenomenon of which the profession needs to take advantage (in both cases, interestingly, flocculation resulted in 48 mg/L of additional TSS removal). Moreover, because chemically enhanced primary treatment is becoming more widely used, it is becoming more important to enhance the formation of floc, improving TSS removal. Historically, preaeration tanks and aerated grit chambers have been successfully used for flocculation in chemically enhanced primary treatment applications. More research is needed to determine the optimum variables for primary clarifier inlet design as it pertains, specifically, to optimizing the flocculation potential of the influent solids.

**SLUDGE COLLECTION AND WITHDRAWAL.** Discussion of sludge collection and withdrawal in primary clarifiers must begin with a discussion of thickening primary sludge. Many of the operational problems with the activated sludge process documented over the years are the direct result of thickening in primary and secondary clarifiers. While it is without question that discharging a thicker primary sludge to the solids treatment train has numerous benefits, it comes at considerable cost to the liquid treatment train. Thickening primary sludge outside the primary clarifier is a paradigm shift for the profession because thickening in the primary clarifier has been done for so long. The fact of the matter is that sludge blankets in primary clarifiers can, and often do, violate all three of the design engineer's goals. High sludge blankets can cause unwanted currents because there is less water mass into which the momentum of the incoming flow can be dissipated. Anaerobic biological activity will commence quickly in primary sludge blankets. This activity will solubilize particulate COD into readily biodegradable volatile fatty acids, generate hydrogen sulfide, decrease the pH of the sludge, and generate gas bubbles that resuspend particles into the water column. Because these particles are resuspended into an environment of low mixing intensity and low solids concentration, the chances of reflocculating these particles is remote. Albertson and Walz (1997) found that increasing sludge blanket retention deteriorated primary effluent quality more so than increasing surface overflow rate. On these accounts, sludge collection and withdrawal should be conducted as quickly as possible. If need be, thickening should be conducted externally to the primary clarifier. Because of a lack of computational or physical modeling results, a specific mechanism cannot be recommended. It should be stated, however, that sludge hoppers should always be located at the influent end of rectangular primary clarifiers, as this location will provide the quickest sludge withdrawal time.

If continuous sludge withdrawal is designed for with primary sludge thickening accomplished external to the primary clarifier—as recommended for optimal performance—centrifugal sludge pumps can be used. In contrast, if some thickening is desired, proper sludge-pumping equipment and controls must be used. The key to this approach is to use a positive-displacement type pump. These units are able to withdraw concentrated sludge at a relatively constant flow without clogging. Pump capacity should be coordinated with the sludge hopper size, whether the primary clarifier is circular or rectangular. The pump should evacuate one hopper volume each pumping cycle. The control sequence should use automatic timers and the timing cycle should be adjustable so that the pump is not started until the hopper has had a chance to "refill" with sludge. With this type of design, the operator will need to periodically monitor the sludge blanket in the clarifier and primary sludge concentration. The pumping time can be increased in response to a buildup of the sludge blanket or decreased in response to a decrease in the sludge solids concentration.

**SCUM COLLECTION AND WITHDRAWAL.** Primary scum consists primarily of fats, oils, grease, and debris. The better the screening operation in preliminary treatment, the less debris there will be in the primary scum. Scum collection and withdrawal have relatively little effect on how a primary clarifier performs relative to flocculation, hydraulics, and biological activity but if these processes are not designed properly they can cause problems for the operators.

Scum is removed in primary clarifiers using automated skimmers connected to sludge collection equipment. Fixed collection troughs with beach plates arranged for localized, partial-radius, or full-radius coverage are used in circular primary clarifiers. Rectangular primary clarifiers use slotted pipe, weir gate, or other special, vendor-furnished devices for primary scum withdrawal

Typically, scum is drained by gravity into a wet well, which is customarily located directly adjacent to the clarifier. From there, the scum is typically pumped to another location for processing and disposal. The following pumps are commonly used to convey this material:

- Progressing cavity,
- Double disc diaphragm,
- Chopper,
- Hose,

- Plunger, and
- Rotary lobe.

With all primary clarifiers, the most difficult aspect of scum removal is to find a suitable disposal outlet. If it is disposed in anaerobic digesters it must be fed as continuously and well mixed with other feed stock as possible. If this strategy is not followed, the material accumulates, undigested, over time. This requires the digesters to be cleaned to maintain optimum active digestion volume. It is not desirable to mix the scum with the sludge because it can inhibit the performance of dewatering devices. Primary scum typically contains significant quantities of plastic and other nonbiodegradable solids, which can adversely affect beneficial reuse of biosolids. Some larger-scale facilities have successfully used scum concentrators. The cost and effectiveness of the scum concentrator must be given careful consideration. These units require heat tracing to keep the material fluidized until it is removed into the disposal container. Careful operation is required to minimize the amount of water mixed with the concentrated scum. If the concentrated scum is too wet, it cannot be disposed in landfills, which commonly place restrictions on material containing less than 20% solids.

One novel approach used at the Brunswick Sewer District in Brunswick, Maine, is to use a static wedge-wire filter screen to dewater the scum, remove solid particles, and return the filtrate to the headworks. The screened solids are placed into the screenings or grit container and codisposed. The screened scum solids concentration equals or exceeds the solids concentration of the grit and screenings material. This method removes much of the undesirable plastic and debris from biosolids at the facility.

**EFFLUENT DISCHARGE.** While primary clarifiers would not be expected to have the density currents caused by concentration differences between incoming flow and the clarifier like secondary clarifiers would, density currents caused by differences in temperature are possible. Positioning of effluent weirs would depend on whether a buoyant current or sinking current formed. Not enough data have been reported, however, to give direction in this regard. For the design engineer, therefore, it boils down to positioning the effluent weirs to provide as ideal hydraulics as possible.

Circular primary clarifiers typically have effluent launders consisting of troughs outfitted with weir plates. Substantially less debris accumulates on the weir plates if rectangular notches, rather than v-notches, are used. The launders may be positioned inboard along the outside wall, outboard along the outside wall, or at an intermediate location away from the outside wall. Rectangular clarifiers have used flat broad-crested weirs located along the back wall of the tank; multiple weir troughs with flat, v-notch, or rectangular-notch weir plates; or submerged pipes with orifices for effluent withdrawal. Again, substantially less debris accumulates on the weir plates if rectangular notches, rather than v-notches, are used.

Some regulatory agencies have had published design guidelines in place for many years restricting average weir loading rates between 124 and 186 $m^3/m \cdot d$ (10 000 and 15 000 gpd/lin ft). There is essentially no published reason for these limitations.

Careful consideration should be given to positioning the scum removal trough at high weir loading rates. Some manufacturers' standard designs are submerged at peak flows, and the skimmer arms do not have sufficient adjustment to allow the beach plate to be raised to overcome flooding.

## RESEARCH NEEDS

Understandably, a list of "research needs" is not often given in a design document. The user of this document, however, must understand where the "science" of primary clarifiers is not well established. It is on this account that this list of research needs, identified throughout the text, is given:

- Testing to determine the settling parameter, $\lambda$;
- How the characteristics of wastewaters change at high flows caused by storm events;
- The flocculation kinetics of raw wastewater solids; and
- Primary clarifier inlet design

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# Chapter 3

# High-Rate and Wet Weather Clarifier Design Concepts and Considerations


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# INTRODUCTION

**BACKGROUND.** Since the passage of the Clean Water Act in 1972, nearly all municipal facilities in the United States have implemented a minimum of secondary treatment. With most dry weather pollution from sanitary sewer systems under control, the attention of the regulatory establishments has shifted to the capture and treatment of wet weather induced overflows and bypass flows that can significantly affect receiving water quality. The water quality effects of wet weather wastewater flows vary depending on their frequency, magnitude, and water quality of the wet weather discharge relative to the flow and quality of the receiving water. Wet weather overflows adversely affect receiving water by impairing aquatic habitat, degrading receiving water aesthetic quality, and potentially affecting human health by contaminating beaches and shellfish (Sherbin and Weatherbee, 1993; U.S. EPA, 2001). Wet weather flows are rainfall induced so that both their magnitude and frequency are variable and the time of occurrence difficult to predict.

Wastewater system flows can increase significantly during rainfall events, even in systems with separate wet weather flow collection. In northern climates and some mountain regions peak flows can result from spring snowmelt. Rainfall-derived wastewater flows that arrive at treatment plants can severely tax both the hydraulic and process capacity. Common practice for older communities in the Northeast, Midwest, and coastal areas of the Pacific Northwest was to design combined sewer systems with combined sewer overflows (CSOs), which allow peak wet weather flows to discharge directly to surface waters, thereby limiting the need to size the downstream transmission and treatment systems for peak flows. To reduce the number and frequency of CSO events and avoid remote wet weather treatment facilities, some communities have chosen to transmit all flows to the wastewater treatment plant by increasing the capacity of the sewer collection/transmission system. While this practice is effective in reducing the discharge of untreated wastewater from the system, it typically only exacerbates the effect of wet weather flows on the treatment facility.

Treatment of wet weather flows resulting from inflow and infiltration to the wastewater collection and transmission system differs significantly from the treatment of the base dry weather wastewater flows in a number of aspects. Wet weather flows are typically short-duration events with flow rates greater than normal diurnal peaks. With knowledge of historical precipitation and the physical characteristics of the service area and the collection/ transmission system, the magnitude and duration of wet weather flows can be predicted; however, the time of occurrence cannot.

Perhaps most importantly, clarification processes treating wet weather flows operate intermittently under continually varying influent flow and pollutant concentrations, with potentially long periods of inactivity between storm events (Field et al., 1997). Treatment efficiency will likely vary temporally depending on the influent flows and loads and the condition of the clarification unit at startup (wet or dry). Similarly the storage volume available in the process tank(s) relative to the duration and intensity of the wet weather flow events will affect the mass of flow and pollutants captured. Numerical simulation and statistics offer rational procedures for evaluating the performance of wet weather treatment alternatives under dynamic conditions (Averill and Gall, 2000).

**CURRENT PRACTICE.** Current practice depends on the type of collection system (separate or combined), location, and state requirements. Before the advent of national CSO regulations, most communities with combined systems continued routine use of CSO facilities or bypassed peak flow around parts of their wastewater treatment facilities. Perhaps the most common practice at treatment plants has been the provision of preliminary and primary treatment for all flows, with bypass of peak flows around secondary treatment, blending of the biological effluent with the bypassed flow, disinfection, and discharge. A variety of other wet weather treatment strategies are in use. Principal alternatives to clarification for wet weather flow management include construction of additional treatment plant capacity, use of inline and offline wet weather storage, decreasing peak flows through reduction of rainfall-derived infiltration and inflow, sewer separation, and rerouting flow to a different treatment plant.

**REGULATORY CONSIDERATIONS.** Wet weather issues came to the forefront in the later part of the 1980s and the early years of the 1990s (U.S. EPA, 1995). The U.S. Environmental Protection Agency (U.S. EPA) issued a National Combined Sewer Overflow Control Strategy in 1989 (54 *Federal Register* 37370, August 10, 1989) and a CSO Control Policy in 1994 (59 *Federal Register* 18688, April 19, 1994). More recently, U.S. EPA issued a proposed policy on blending (68 *Federal Register* 63042, November 7, 2003). However, in May 2005, U.S. EPA announced that the blending policy will not be finalized. January 1, 1997, was the deadline set by the 1994 policy for implementing minimum technology-based controls, known collectively as the "nine minimum controls". One of these requires that communities maximize flow to the local publicly owned treatment works (POTW). National Pollutant Discharge Elimination

System (NPDES) permit holders are required by the 1994 policy to develop long-term plans for controlling CSOs. Long-term control plans must either demonstrate that the plans are adequate to meet water quality requirements or implement a minimum level of treatment (U.S. EPA, 1995). Water-quality standards are presumed to be met if the technology-based approach is used. Some states have implemented laws or regulations that go beyond the U.S. EPA CSO policies. Georgia, for example, requires that all flow entering a POTW receive secondary treatment.

Regulatory requirements pertaining to CSOs and sanitary sewer overflows (SSOs) continue to change at a fairly rapid pace. Many utilities have been required to reduce the frequency of SSOs and CSOs as a result of evolving regulations. To accomplish this goal, utilities often rely on a combination of increased trunk sewer capacity, wet weather storage, remote CSO treatment facilities, infiltration/inflow removal, sewer separation, and potentially the routing of peak storm flows to new CSO treatment plants. Combined sewer overflow programs other than increased transmission system capacity will reduce the peak flows received at wastewater treatment facilities, thereby allowing these utilities to maximize the capacity of these existing treatment facilities to treat municipal wastewater.

**ROLE OF CLARIFICATION.** As a result of current regulations, most municipal wastewater treatment plants are expected to provide some degree of treatment to all of the flow received at their facilities regardless of the magnitude and duration. Clarification is often a key component of wet weather treatment strategies. Wet weather treatment strategies may include measures to minimize the investment in treatment facilities for peak wet weather flows that occur infrequently, while still providing adequate protection for receiving water. Examples of wet weather treatment strategies incorporating clarification range from increasing the rated capacity of existing conventional primary clarifiers to construction of dedicated wet weather clarifiers. Alternatively process modifications can be implemented to protect secondary settling tanks from the effect of periodic high flows.

Wet weather clarifiers can be storage basins operated as flowthrough clarifiers once the storage volume is full, conventional clarifiers operated at traditional loading rates, or clarifiers enhanced by one or more modifications designed to increase the allowable hydraulic loading or improve pollutant removals. Many names are used to describe enhanced clarification processes, including high-rate clarification (HRC), enhanced high-rate clarification, high-rate flocculated settling, dense sludge, high-rate sedimentation, microcarrier weighted coagulation, ballasted flocculation, chemically enhanced high-rate separation, and microcarrier coagulation–sedimentation. High-rate clarification will be used in this chapter to describe advanced clarification processes that use some combination of chemical coagulation, ballast, and plates or tubes to improve clarifier performance. More detail is provided on HRC processes later in this chapter.

Chemically enhanced primary treatment (CEPT), whereby wastewater is chemically coagulated before clarification, is the simplest enhancement to conventional primary clarification used to treat wet weather flows. The use of chemicals (typically metal salts and polymer; see Chapter 2) allows a higher peak overflow rate during peak flow events, while still maintaining primary clarifier performance. This minimizes the clarifier surface area that must be provided for peak flows. Polymer used alone in high doses can also provide consistently high performance, while enhancing the ability to disinfect with UV light (Averill et al., 1999). Chemically enhanced primary treatment can be a full-time treatment method; however, when used for control of wet weather flows its use is limited to peak wet weather periods.

Inclined plates or tubes significantly increase the allowable upflow velocity in a clarifier (based on horizontal tank area) by increasing the settling area by a factor of approximately 8 to 10, thereby allowing a higher peak flow to be treated in a given tank surface area. While the classic location for plates and tubes is in primary clarifiers, they have been used in secondary clarifiers, and researchers in Germany have investigated their use at the end of the aeration tanks or at the entrance to secondary settling tanks (Buer et al., 2000; Plaß and Sekoulov, 1995). Plates or tubes in these locations reduce the mixed liquor suspended solids (MLSS) concentration entering secondary settling tanks, thereby increasing the peak flow capacity of the secondary settling tanks.

Chemical coagulation can be combined with the use of recycled sludge (dense sludge process) or floc-weighting agents (ballasted flocculation) and tubes or plates (Lamella®) to achieve additional increases in overflow rate and performance. Two forms of HRC are in current use—dense sludge and ballasted flocculation. Dense sludge refers to a HRC process that combines chemical coagulation, sludge recirculation, and Lamella settling, whereby solids inherent to the influent water are recycled to increase particle density and settling velocities. Ballasted flocculation refers to clarification processes that increase particle size and density, and hence settling velocity, by binding solids to a weighting agent or "ballast" with metal hydroxide floc and polymer. Very small sand particles (microsand) are the most common ballast.

One common wet weather flow control strategy is to provide primary treatment for all flows followed by biological treatment for a base flow (some factor times dry weather flow). Flows above the base flow can be disinfected and discharged or blended with biological effluent, disinfected, and discharged. Similarly, flows greater than the capacity of existing facilities can be treated in dedicated wet weather clarifiers followed by disinfection.

Rather than attempting to increase primary treatment capacity, an alternative wet weather strategy is to implement enhancements to the biological process that increase the capacity of secondary settling tanks during wet weather flows. Common techniques are to switch the aeration tank feed pattern to a step-feed or contact stabilization activated sludge configuration or to provide additional "wet weather" secondary settling tanks. Wet weather secondary clarifiers can be constructed that serve the dual purposes of wet weather flow storage and secondary settling; however, storage at this location in the process provides few benefits (Carrette et al., 2000). Using step-feed operation allows the plant to maintain a relatively high degree of treatment, while treating a significantly higher flow rate. Another wet weather treatment method that relies on the same basic mechanism as step-feed is aeration tank settling (ATS). Turning the air off in all or just the later parts of an aeration tank during peak flow periods allows the MLSS to begin to settle in the aeration tank and reduces the MLSS concentration entering the final clarifiers.

Wet weather clarifiers can be located directly at a municipal plant or remotely at a satellite facility. At remote locations, peak flows are diverted to wet weather flow clarifiers, with overflow to receiving water during events exceeding the design capacity (Schindewolf et al., 1995). The clarifier contents, including sludge and floating materials, are returned to the sewer system after wet weather flows end. Intermittently operated clarifiers provide removal of soluble and particulate solids captured by storage during small storms and additional removal of particulate solids by sedimentation and flotation during larger events that result in overflows (Schraa et al., 2004).

While vortex separators, also known as swirl concentrators, are commonly used to treat wet weather flows and CSOs, they can also be used to treat peak wet weather flows at wastewater treatment facilities. Vortex separators can be used with and without chemical flocculation in a manner analogous to conventional primary clarifiers. Vortex separator configurations with specially designed flow modifying internal components are resilient to shock hydraulic and solids loadings. Because of their hydrodynamic flow regime, vortex separators have been found to be suitable for use as "plug-flow mixing devices" for chemical disinfection. Recent developments in the technology have resulted in configurations that combine a number of unit processes (sedimentation, disinfection, and screening) in a single vessel.

**ROLE OF STORAGE.** Storage can be used alone or in combination with clarification to increase wet weather treatment capacity. Diversion of peak flows to storage tanks, tunnels, or lagoons either before or after primary treatment reduces the magnitude of peak flows, thereby enabling plants to adequately treat wet weather flows that would otherwise be beyond the plant capacity. Storage tanks and lagoons can be designed to provide some primary settling by providing two or more cells in series. The first cell functions as a primary settling tank and minimizes the poststorm cleanup required in subsequent cells. Another approach that combines elements of storage and treatment is to construct deep secondary settling tanks to provide short-term storage for wet weather induced solids loads that exceed the maximum solids flux. Although this technique will only allow a plant to maintain effluent quality during relatively short-duration peak flows, this may be adequate in many situations.

The overall cost of managing wet weather flows can be reduced by combining treatment techniques with some form of wet weather storage. This concept is illustrated in Figure 3.1, which shows the relative capital costs for combinations of treatment and storage to provide facilities for design wet weather flow as a function of the fraction of the wet weather capacity provided by storage. The storage volume required is the volume of wastewater generated beyond the treatment capacity provided as shown in Figure 3.2. As the wet weather treatment capacity provided

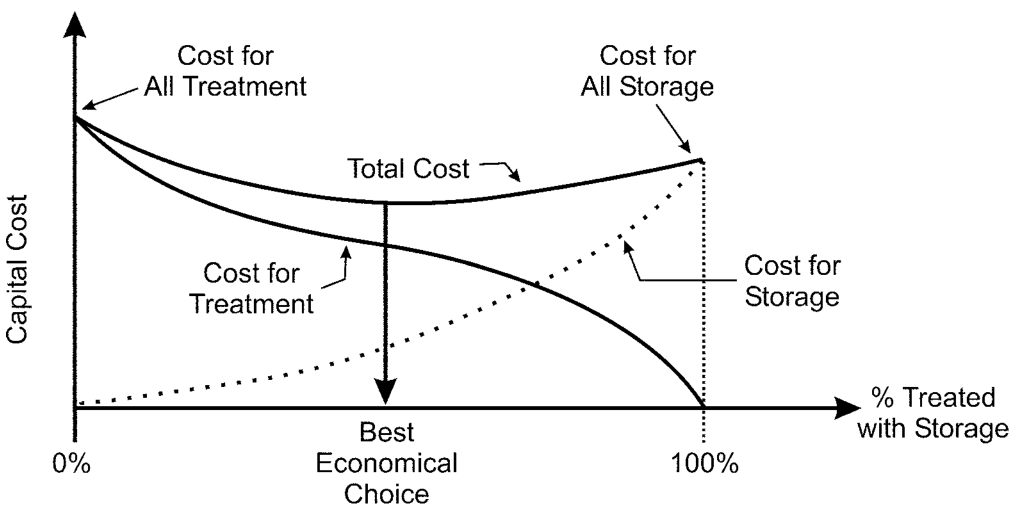


**FIGURE 3.1** Relative cost of combination of treatment and storage for wet weather flows.

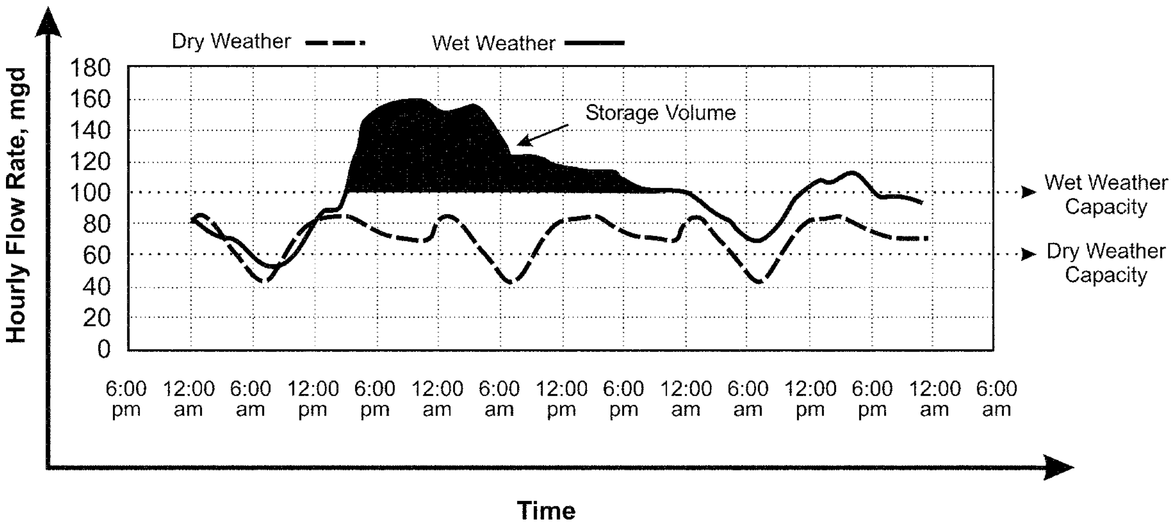


**FIGURE 3.2** Comparison of daily flow pattern—typical dry weather versus wet weather flows (mgd $\times$ 3785 = $m^3/d$).

increases, the cost of storage decreases while the cost of treatment increases. For the example in Figure 3.2, the use of treatment and storage in combination is less expensive than either treatment or storage alone. Consideration must be given to the difference in pollutant-removal efficiency provided by the main treatment facility versus that obtained from a wet weather clarifier. Because all of the wet weather flow from a storage facility is returned to the main stream, more pollutant removal is provided by storage than wet weather treatment (Averill and Gall, 2000). Combinations of treatment and storage should be investigated as part of preliminary planning for most wet weather treatment projects to establish the potential to optimize cost and pollutant-removal efficiency.

**METHODOLOGY.** Traditional methods used for sizing and predicting performance of clarifiers treating dry weather flow are inadequate for evaluating the relative cost and efficiency of different types of intermittently operated wet weather clarifiers receiving dynamic flows and loads with the resulting variations in treatment efficiency. Two general approaches have been identified. One is to create a design storm that is representative of average annual conditions and the second is to perform long-term, dynamic simulations (Schraa et al., 2004). Either approach requires that representative relationships be established between flow, pollutant concentrations, and treatment efficiency. Whereas new approaches are required for predicting performance of wet weather clarifiers, even these methods must be based on design particle settling velocities. Hence, the general clarifier design approach proposed in Chapter 1 can be integrated to wet weather clarifier designs.

Design wet weather flows can be established through the use of many common hydrological methods. Included in this group are (1) the rational method, (2) the U.S. Department of Agriculture (USDA)/National Resources Conservation Service (NRCS) Technical Release 55 (TR-55) method, (3) USDA–NRCS TR-20 Model, and (4) the U.S. Army Corps of Engineers Hydrologic Engineering Center (HEC)-1 Model. Use of these methods is beyond the scope of this document, and the reader is referred to other references that provide detailed information (NJDA et al., 2000; USDA and NRCS, 1992, 1997, 2002; USDA and SCS, 1986). Useful models for the purposes of estimating wet weather storage volumes are the HEC Storage, Treatment, Overflow, and Runoff Model (STORM) (U.S. Army Corps, 1976) and the TRTSTORM model (Kluitenberg and Cantrell, 1994). TRTSTORM is a modified version of STORM that uses a statistical approach based on historical rainfall and evaporation rates and an infiltration coefficient to estimate the relationship between treatment capacity, storage volume, and number of treatment capacity exceedances or overflows. TRTSTORM tracks the number, duration, and volume of flow exceedances, allowing the user to optimize the combination of storage and treatment facilities for design. Figure 3.3 contains a graph showing an example of the estimated plant site storage volume requirements versus the annual number of times the peak flow exceeds treatment

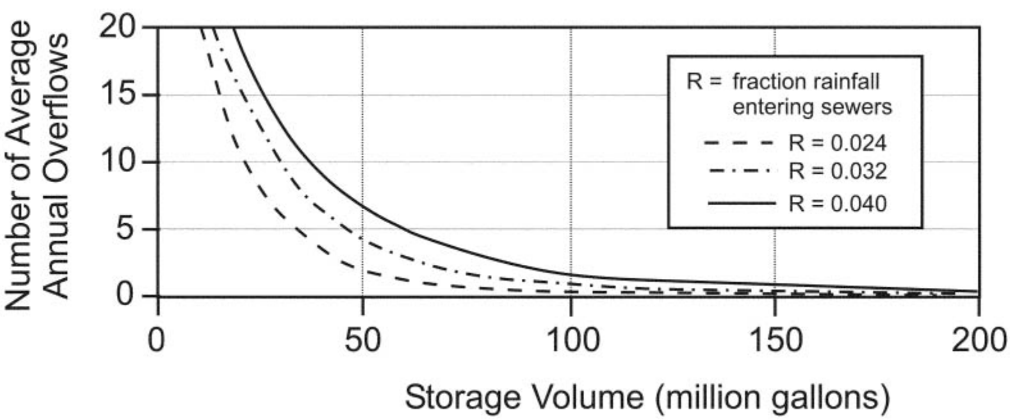


**FIGURE 3.3** Example curves showing wet weather storage volume versus estimated number of overflows per year.

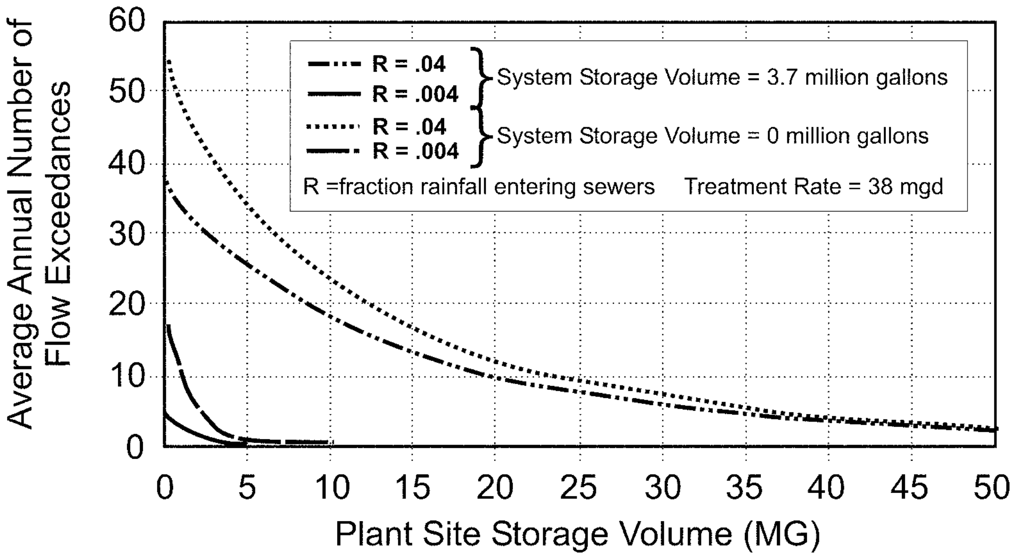


**FIGURE 3.4** Example curves showing effect of system storage on number of times per year plant flows exceed treatment capacity (MG = million gallons; mgd × 3785 = $m^3/d$).

capacity. Figure 3.4 contains similar information for simulations with and without system storage in addition to plant site storage. Both Figures 3.3 and 3.4 were developed for a specific collection system and are not applicable to other locations. Use of the STORM models is also beyond the scope of this manual.

Development of design suspended solids time series data that correspond with a design hydrograph is more challenging, in part, because of the lack of a significant historical database and the cost involved in developing one. Research conducted as part of Canadian efforts to reduce pollution in the Great Lakes from CSOs created a method to develop the required flow, pollutant, and treatment relationships and apply them to dynamic simulations of the efficiency of wet weather treatment processes (Averill and Gall, 2000; Gall et al., 1997; Schraa et al., 2004).

Treatment efficiency for all clarification processes is a function of the hydraulic loading rate relative to the settling velocities of suspended solids in the wastewater. Regardless of the design method selected, sufficient evaluations should be conducted to make rational estimates of wet weather hydrographs, particle concentrations and settling velocities of the wet weather wastewater, and treatment process efficiency under variable loads.

# BASICS—THE SCIENCE OF DESIGN

Performance of all clarification devices is determined, in large part, by the settling characteristics of suspended particles, especially the settling velocity. Clarifiers used for wet weather treatment conform to the same theories as primary and secondary clarifiers in traditional applications. Settling in primary clarifiers is flocculent or type 2 settling, whether it is used for dry or wet weather wastewater. Settling in secondary sedimentation tanks is hindered, or type 3, settling. Primary sedimentation with the addition of waste sludge to the primary influent (cosettling) is still type 2 or flocculent settling under nearly all conditions. With a strong wastewater (biochemical oxygen demand [$BOD_5$] greater than 300 mg/L) and a short solids retention time (SRT) activated sludge process, the suspended solids concentration in the influent to primary clarification could increase approximately 500 mg/L or more with the addition of all waste sludge to the primary clarifiers, and type 3 settling might result. See Chapters 2 and 4 for detailed discussions of the science of primary and secondary clarifiers.

**WASTEWATER CHARACTERISTICS.** Water quality and resulting mass loads imposed on the treatment process by wet weather flows differ from the base dry weather flow. Wet weather flows can have significantly lower concentrations of some pollutants and higher concentrations of others depending on antecedent conditions, the magnitude of the flows, and the time since the start of a storm event. As discussed later, the "first flush" of wet weather flow often results in a transient increase in the mass load of pollutants received at a treatment plant. Exceptionally high and prolonged wet weather flows can resuspend sediments deposited in the collection system or scour biomass from pipe walls and transport both to the treatment plant. Depending on the season and location, wet weather flows can be colder or saltier than normal flows. In addition, it is reasonable to expect that wet weather flows will have different amounts of organic matter and different frequency distributions of particle sizes. For instance, the proportion of soluble $BOD_5$ and the fraction of particles in wet weather flows that can be removed by gravity settling may be different from dry weather wastewater. Thus, evaluation of wet weather treatment should be based as much as possible on characterization of real wet weather flows generated in the collection system.

Significant work has been done to characterize urban water, including wastewater; wet weather flows; and, to some extent, peak wet weather wastewater quality.

Research and practical experience show that both dry weather and wet weather wastewater contain a complex mixture of solids. Suspended solids present in wet weather wastewater originate from three main sources—surface runoff that enters the collection system, biofilms or slimes that erode from conduit walls, and native particulate matter from sanitary wastes (Michelbach, 1995). The composition, size, and settling characteristics of these solids are likewise a complex function of many parameters, including pipe sizes; materials; slopes; range of water velocities experienced; type of collection system (separate or combined); size and physical characteristics of the service area; duration and intensity of rainfall; and, to some extent, historical changes in these parameters.

Particulate inorganic and organic materials in typical wastewater are reported to range in size from smaller than 0.001 μm to larger than 100 μm (Levine et al., 1991a, 1991b; Odegaard, 1979). Functional definitions for particle sizes are given in Table 3.1 (Levine et al., 1991a).

Table 3.2 provides a summary of reported size distributions for organic particles in wastewater (Levine et al., 1991a). In a typical wastewater, approximately 30% of the carbonaceous oxygen demand (COD) may be associated with settleable particles, approximately 25% with supracolloidal particles, and 15% with colloidal particles (Levine et al., 1991b). As with settling velocity data, however, few data have been published on particle sizes and densities in wet weather flows. Reported values for the unsettleable fraction of wastewater solids and the distribution of particle sizes cover a wide range. Collection of site-specific data for wastewater characteristics is desirable.

**FIRST FLUSH.** Research has been conducted to understand the effect of hydrodynamic mechanisms on the "first flush" of pollutants received at a wastewater

**Table 3.1** Functional definitions of particle size (Levine et al., 1991a).

| Category | Particle size (μm) |
|---|---|
| Dissolved | <0.001 |
| Colloidal | 0.001–1 |
| Supracolloidal | 1–100 |
| Settleable | >100 μm |

**Table 3.2** Distribution of organic matter in untreated municipal wastewater (Levine et al., 1991a)

| Percent of organic matter contained in indicated size range (μm) | | | | |
|---|---|---|---|---|
| **< 0.001** | **0.001–1** | **1–100** | **> 100** | **Reference** |
| 41 | 16 | 28 | 15 | (Balmat, 1957) |
| 31 | 14 | 24 | 31 | (Heukelekian and Balmat, 1959) |
| 38 | 13 | 19 | 30 | (Painter and Viney, 1959) |
| 29 | 13 | 31 | 27 | (Walter, 1961a) |
| 29 | 15 | 22 | 34 | (Walter, 1961b) |
| 25 | 14 | 27 | 34 | (Hunter and Heukelekian, 1961) |
| 18 | 15 | 25 | 42 | (Hunter and Heukelekian, 1961) |
| 25 | 14 | 27 | 34 | (Hunter and Heukelekian, 1965) |
| 23 | 14 | 23 | 40 | (Hunter and Heukelekian, 1965) |
| 30 | 19 | 10 | 41 | (Hunter and Heukelekian, 1965) |
| 50 | 9 | 18 | 23 | (Rickert and Hunter, 1967) |
| 47 | 9 | 19 | 25 | (Rickert and Hunter, 1967) |
| 40 | 10 | 21 | 29 | (Rickert and Hunter, 1971) |
| 12 | 15 | 30 | 43 | (Munch et al., 1980) |

treatment plant in a combined system (Krebs, Holzer, et al., 1999; Krebs, Merkel, and Kuhn, 1999). From wave theory, it can be shown that the wave velocity is greater than the flow velocity and that the addition of rainwater to a sewer can result in the formation of a wave that travels downstream faster than diluted wastewater. In a combined sewer, this means that the "first flush" of pollutants received at a wastewater plant during a storm event is often composed of undiluted wastewater. The normal concentrations combined with the higher flow rate can result in a significant increase in load. Depending on the time of day when the storm event begins, the load on the wastewater plant can be doubled. This wave effect is most pronounced in systems with mild slopes, long residence times, and long reaches before the wastewater plant where no further inflow is added to the sewer and when the rain event is intense.

**SETTLING VELOCITIES.** Settling-velocity distributions for dry and wet weather wastewater have been reported by a number of researchers over the past decade (Krebs, Merkel, Kuhn, 1999; Michelbach and Wohrle, 1992, 1993; Pisano et al., 1990; Shin et al., 2001; Tyack et al., 1996). Typically, the size, density, and settling velocity of suspended solids cover a wide range. Certain generalizations, however, can be made about the relative settling velocity of suspended solids in combined wastewater. Particles in wet weather flows tend to be heavier, denser, and faster settling than solids in either dry weather wastewater or street runoff (James, 2002; Michelbach and Wohrle, 1993; Shin et al., 2001). Two mechanisms are responsible for the increase in settling velocities observed in wet weather flows. First, higher flows increase the shear stress at the pipe walls and increase the sediment-transport capacity of the collection system, allowing coarser, faster settling solids that have accumulated in the collection system to be resuspended. Second, the higher velocities erode biofilms from the pipe walls.

Settling-velocity data from a number of sources are summarized in Table 3.3. Figure 3.5 combines data from Pisano et al. (1990) and Michelbach and Wohrle (1993) and highlights the range and variability of measured settling velocities in wastewater. The Pisano et al. and Michelbach and Wohrle curves show that approximately 80 to 90% of the solids from the German studies settle with velocities greater than 10 m/h, whereas the settling velocities reported for many American cities are significantly lower, with mean settling velocities in the range of 0.70 to 4.0 m/h. Figure 3.6 shows the range of settling velocities reported for suspended solids in combined wastewater and dry weather flow at one location (Michelbach and Wohrle, 1993).

Settling-velocity data from various types of collection systems were categorized according to criteria presented in Table 3.4 and conclusions were made about the effect of the collection system on the expected settling velocity. In large collection systems, organic material begins to degrade, which decreases settling velocities and increases turbidity and the fraction of unsettleable solids (Smisson, 1990).

In doing this analysis, it was reasoned that the limits to settling velocity are between 0.36 and 100 m/h. On the low end, Stokes law predicts settling velocities of 0.32 to 0.43 m/h for particles with diameters of 8 to 12 μm and specific gravities of

**TABLE 3.3** Wastewater settling velocities (Smisson, 1990).

| Catchment | Unsettleable solids (%) | Median settling velocity (m/h) | General catchment size |
|---|---|---|---|
| APWA[a] grit | — | 101.5 | Not wastewater |
| APWA[a] organics | — | 40.3 | Not wastewater |
| Rural[b] | 1.5 | 32.1 | Very small |
| Primary[b] | 8.5 | 12.8 | Not wastewater |
| Raw wastewater[b] | 9.5 | 25.5 | Small |
| Raw wastewater[c] | 15.0 | 25.5 | Small |
| Primary[c] | 9.5 | 12.8 | Not wastewater |
| Exeter low limit | 42.5 | 6.4 | Large |
| Exeter upper limit | 32.5 | 10.2 | Large |
| Boston commerce | 14.0 | 10.2 | Not typical |
| Saginaw, Michigan | 31.0 | 4.0 | Large |
| Burlington, Vermont | 32.5 | 6.4 | Large |
| Boston residential | 46.0 | 8.1 | Large |
| Philadelphia, Pennsylvania | 30.0 | 4.0 | Large |
| San Francisco, California | 43.0 | 4.0 | Large |
| City of New York, average | 45.0 | 1.6 | Large |
| James Bridge, Walsall 1 | 18.0 | 8.1 | Medium |
| James Bridge, Walsall 2 | 16.2 | 16.1 | Medium |
| James Bridge, Walsall 3 | 38.0 | 10.1 | Medium |
| Bexhill | 10.0 | 32.1 | Small |

[a]APWA = American Public Works Association.
[b]Bulk sampling technique.
[c]Small tube sampling technique.

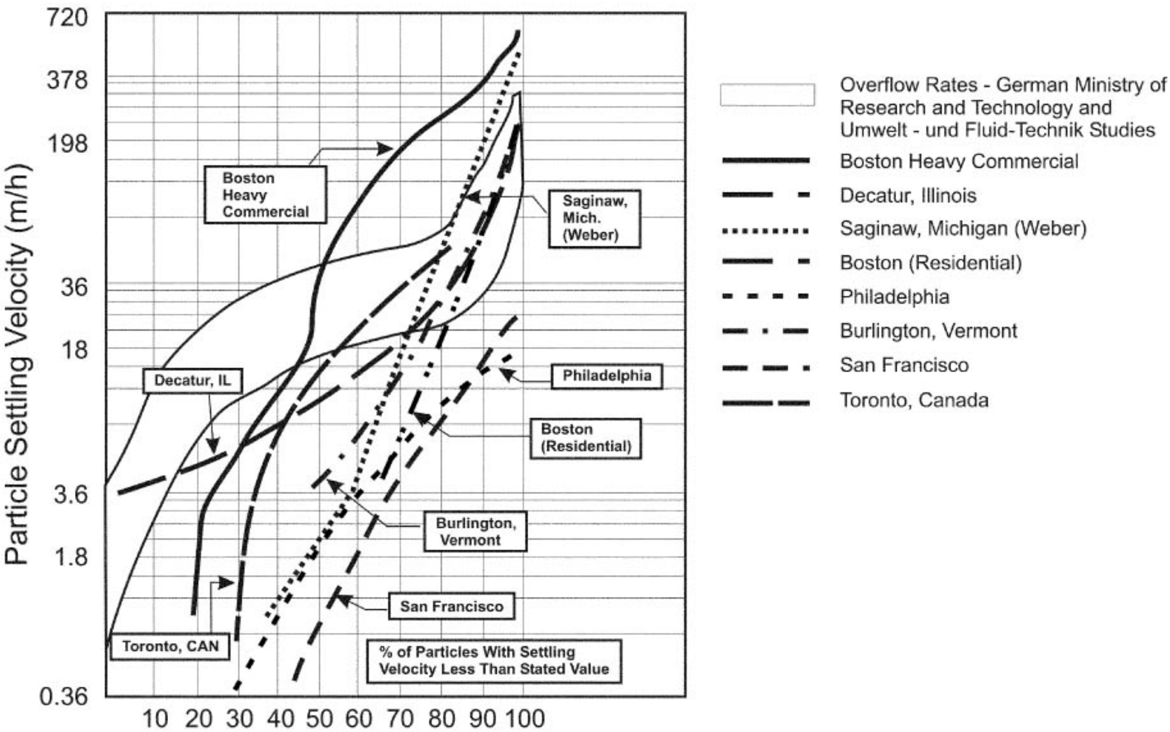


**FIGURE 3.5** Reported settling velocities for wet weather flow solids (Michelbach and Wohrle, 1993; Pisano, 1990).

2.6. For the upper end, it was assumed that particles with settling velocities greater than 100 m/h would be excluded from automatic samplers.

**MEASUREMENT OF SETTLING VELOCITY.** Probably the major reason for the lack of more information on particle size and settling velocities in municipal wastewater is the difficulty in collecting accurate and representative data. All velocity-measurement methods are based on the use of a settling column or series of settling columns. The methods differ in the column size, configuration, and test procedure. Each method has advantages and disadvantages. Detailed information on settling test apparatus and procedures can be found in several references (Aiguier et al., 1996, 1998; Camp, 1945; Metcalf and Eddy, 1991; O'Connor et al., 1999; Tyack et al., 1992).

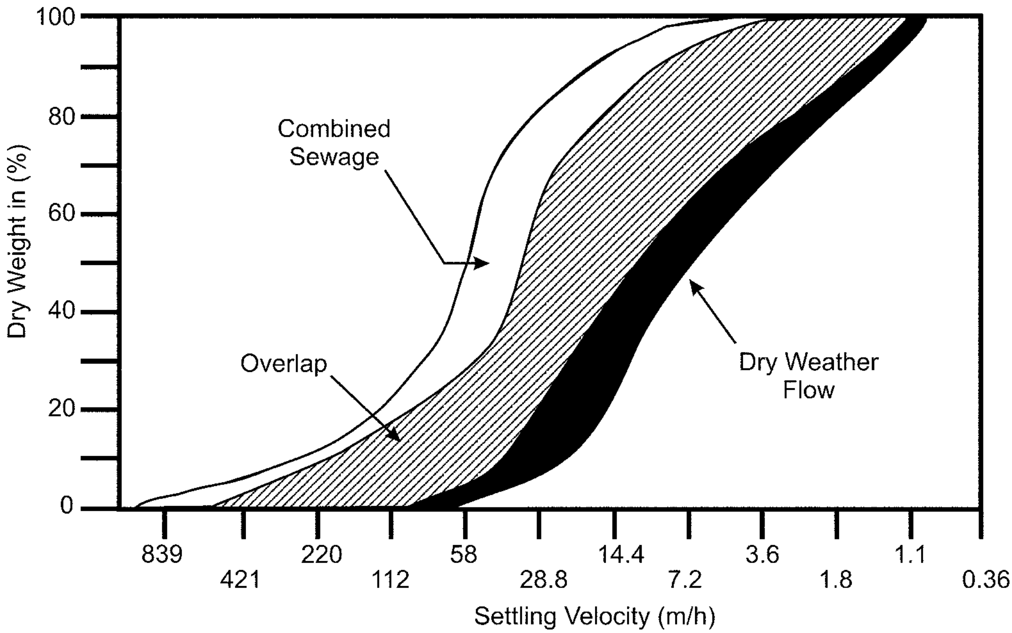


**FIGURE 3.6** Range of particle-settling velocities reported for dry and wet weather flows (adapted from *Water Sci. Technol,* **27,** 153–164, with permission from the copyight holder, IWA).

Settling velocities for type 2 flocculent settling have been traditionally measured using a long settling column with sample ports at regular depth intervals (Metcalf and Eddy, 1991). A typical long column has an internal diameter of approximately 190 mm, a height of 1.8 to 2.5 m, and a volume of approximately 70 L. Sample ports are located approximately every 0.3 m (1 ft) on opposite sides of the column (16

**TABLE 3.4** Settling velocity categories (Smisson, 1990).

| Category | Time of concentration (min) | Septic wastewater | Settling velocity (m/h) |
|---|---|---|---|
| Very large | Very long | Yes | < 30 |
| Large | Very long | No | Some > 300 |
| Medium | 20–30 | No | — |
| Small | < 20 | No | 50% > 30 |

total). Long columns are filled by pumping the sample into the top of the column or through a valve in the bottom. Traditional long column tests have several shortcomings, especially in the measurement of fast settling particles (O'Connor et al., 1999). One significant problem with traditional long columns is the difficulty in obtaining a uniform mixture at the beginning of the test (because of height and volume of the column). As a result, the overflow rates for fast settling particles tend to be overstated. Other problems include the inability to obtain simultaneous measurements at $t_o$, the large testing volumes required, and the large number of solids analyses required.

U.S. EPA research to develop a better method for measuring particle settling velocities has focused on a method developed by Centre d'Enseignement et de Recherche pour la Gestion des Ressources Naturelles et de l'Environnement (CERGENE) of France (O'Connor et al., 1999). In the CERGENE method, suspended solids removal and settling velocities are measured in a series of four or more columns. Each column is 1 m tall with an inner diameter of 65 mm with a volume of 2.2 L. A vacuum pump is used to fill each column in a time-delayed sequence. Each column has three valves—one at the top, one at the bottom, and a 65-mm full port ball valve 40% of the length from the bottom. By closing the ball valve in each column at a specified sampling time, each column provides a discrete measurement of suspended solids removal at one settling velocity. Average settling velocity is computed by dividing 0.5 the length of the upper portion of the column by the sample time. Suspended solids removal is computed by comparing the suspended solids concentration of the top portion by the assumed or measured initial suspended solids concentration. While the sample volume required by this method is reduced, it suffers from its own problems, including a lack of repeatable results, difficulty with suspended solids analysis because of the large volume of analyte, and a loss of sand mass (sand recoveries are often significantly less than 100%).

**ESTIMATION OF SETTLING THEORY.** In the absence of actual settling-velocity measurements, settling velocities must be estimated from particle size and density using Stokes law. See Chapter 4 for a discussion of Stokes law.

**COAGULATION/FLOCCULATION.** The maximum total suspended solids (TSS) and $BOD_5$- or COD-removal efficiency that can be obtained by any sedimentation process can be no better than the percentage of settleable TSS in the wastewater and the fraction of the $BOD_5$ or COD that is associated with the settleable solids.

Coagulation improves TSS and $BOD_5$ removal by increasing particle size and settling velocity by associating or aggregating the colloidal particles with particles of "settleable" size. Coagulation and flocculation are typically associated with the use of chemicals. However, the energy input associated with rapid mix and flocculation facilities can result in larger particle sizes and enhance the performance of sedimentation tanks even without the use of chemicals (Wahlberg et al., 1999). Conventional primary clarifiers with typical TSS- and BOD-removal efficiencies of approximately 50% and 30%, respectively, are reasonably efficient at removing settleable particles (Odegaard, 1998; Wahlberg et al., 1997). The efficiency of primary sedimentation, however, can be increased significantly—to 40 to 80% for organic carbon and to 60 to 90% for suspended solids—by increasing the fraction of particles of settleable size. With chemical treatment, particles as small as to 0.1 μm can be removed from wastewater by sedimentation (Levine et al., 1985).

Coagulation kinetics for conventional and high-rate clarification processes can be predicted by the same equations (Argaman and Kaufman, 1970; Letterman et al., 1999; Stumm and Morgan, 1996; Young and Edwards, 2000). Use of traditional coagulation rate equations for detailed design is hindered by the lack of information about values of rate coefficients and the sometimes difficult mathematical manipulation of equations. Despite the difficulty in quantitative use of rate equations, they are valuable in understanding the basic mechanisms involved and the qualitative effect of different design concepts. For example, the rate of coagulation in ballasted flocculation can be predicted according to the following equation (Argaman and Kaufman, 1970; Young and Edwards, 2000):

$$\frac{dN}{dt} = -K_A NG + K_B N_o G^p \tag{3.1}$$

Where

$K_A$ = aggregation constant (see Table 3.5),
$G$ = mean velocity gradient ($s^{-1}$),
$K_B$ = breakup constant (see Table 3.5),
$N$ = number concentration of primary (original) particles remaining at any time ($m^{-3}$),
$N_o$ = initial number concentration of primary particles entering flocculation($m^{-3}$), and
p = exponent (typically 2).

**TABLE 3.5** Agglomeration and breakup coefficients for high-rate clarification (Young and Edwards, 2000).

| Water treated | $K_A$ | $K_B$ | $G$ ($s^{-1}$) |
|---|---|---|---|
| Conventional—surface water | $1.2 \times 10^{-5}$ | $0.8 \times 10^{-7}$ | 15–120 |
| Conventional—wastewater | $2.3 \times 10^{-4}$ | $(12 \ln G + 9.1) \times 10^{-7}$ | 12–150 |
| Ballasted flocculation—wastewater | $1.1 \times 10^{-5}$ | $(-5.7 \ln G + 41) \times 10^{-9}$ | 400–1200 |

Whereas the rate of floc aggregation ($dN/dt = K_B N_o G^p$) in ballasted flocculation is similar to that in conventional systems, the rate of particle breakup ($dN/dt = -K_A NG$) is much less (Young and Edwards, 2000). Implications for the design of flocculation facilities for ballasted flocculation are that higher $G$ values and shorter detention times can be used to achieve the same degree of particle aggregation.

The time required for effective flocculation decreases as floc volume increases, which is why the time to flocculate dense sludge and ballasted floc is lower than for chemically enhanced settling processes without recycled sludge or ballast. Calculations (Young and Edwards, 2000) show that there is an optimum combination of ballasting agent and chemical precipitate for a given settling time and surface overflow rate. An overview of coagulation theory and available coagulants is provided in Chapter 5.

All CEPT processes will increase the amount of primary sludge produced. Both the precipitation of chemical solids and increased removal of solids contribute to the increased sludge. A subsequent reduction in biological sludge production resulting from the decreased BOD load sent to biological treatment will partially offset the increased primary sludge. Consideration must be given to both the increased mass and volume of sludge. Several methods have been published for estimating the mass of additional sludge produced. Some were developed specifically for CEPT (Morrissey and Harleman, 1992; Odegaard, 1998), whereas others are intended to estimate the production of chemical sludge associated with chemical phosphorus removal but are also applicable to the use of chemicals for enhanced solids removal (Jenkins and Hermanowicz, 1991; U.S. EPA, 1976, 1987; WEF, 1998a) (also, see Chapter 5). Estimates of the additional sludge volume must be based on industry guidelines, empirical experience, or testing.

One method for estimating the quantity of sludge from a CEPT process using ferric chloride is given by eq 3.2 (Morrissey and Harleman, 1992).

$$RS = 1.0\,TSS_{rem} + 1.42\,P_{rem} + 0.66\,FeCl_{3in} \tag{3.2}$$

Where

RS = raw sludge concentration (mg/L),
$TSS_{rem}$ = influent TSS minus effluent TSS (mg/L),
$P_{rem}$ = influent phosphorus minus effluent phosphorus (mg/L), and
$FeCl_{3in}$ = concentration of metal salt ($FeCl_3$) added (mg/L).

Changes in numerical coefficients are required for other primary coagulants. Two simplifying assumptions are inherent in eq 3.2. Ferric chloride is assumed to precipitate only as ferric hydroxide and ferric phosphate, and the added ferric results in formation of $FePO_4$ first and that any excess forms $Fe(OH)_3$. More recent work on phosphorus removal suggests other precipitates also form (Takács et al., 2004; WEF, 1998a). A similar equation was published based on Scandinavian experience (Odegaard and Karlsson, 1994).

$$SP = SS_{in} - SS_{out} + K_{prec}\left(D\right) \tag{3.3}$$

Where

SP = sludge production (mg/L),
$SS_{in}$ = influent TSS (mg/L),
$SS_{out}$ = effluent TSS (mg/L),
$K_{prec}$ = sludge production coefficient (mg TSS/mg metal ion [Me] [4 to 5 for iron and 6 to 7 for aluminum]), and
$D$ = dose of metal salt (mg/L).

**PLATES AND TUBES (LAMELLA®).** Early papers published by Hazen et al. (1904) and Camp (1945) developed the theory for sedimentation tank design. They were the first to establish that suspended solids removal in gravity clarifiers depends only on the surface area and not the tank depth. Plates or tubes installed at an angle in a clarifier will significantly increase the settling area available within a given footprint. This can be readily demonstrated by filling a long, thin glass tube with fine sand and water and observing the time for the sand to settle when the tube is vertical compared with the time when the tube is held at an angle. The term Lamella

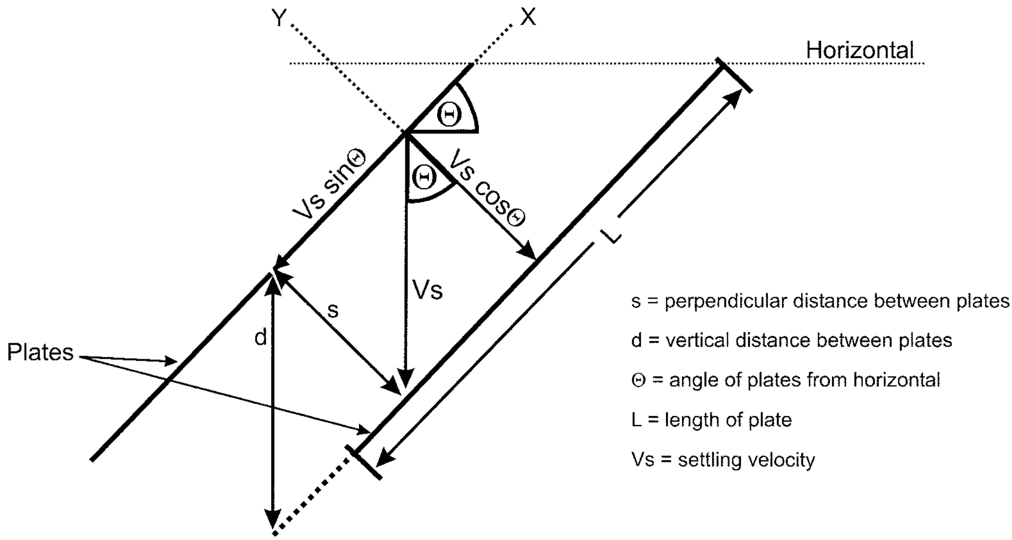


**FIGURE 3.7** Lamella settling definitions ($s$, $d$, and $L$ are in meters; θ is in degrees; and $V$s is in meters per hour).

(Parkson Corporation, Fort Lauderdale, Florida) is a registered trademark of one product; however, it is commonly used interchangeably with "plates" and "tubes".

Figure 3.7 defines the basic geometry for calculating the additional area provided by Lamella. The basic equations are provided in Table 3.6.

Particle velocity vectors for Lamella settlers are given by the following equations (Andersen, 1996):

$$u_{sx} = u - u_s \sin \Theta$$
$$u_{sy} = -u_s \cos \Theta$$
$$u = \frac{v}{\sin \Theta} \qquad \text{(3.4) through (3.7)}$$
$$-u_{sy} = u_s \cos \Theta \geq \frac{Q_L}{A_L}$$

Where

$Q_L$ = flow through Lamella ($m^3$/h),
$A_L$ = total Lamella area ($m^2$),

$\Theta$ = angle of Lamella from the horizontal (deg),
$v$ = vertical fluid velocity through the Lamella (m/h),
$u$ = fluid velocity (m/h),
$u_s$ = settling velocity of free-falling particles (m/h),
$u_{sx}$ = settling velocity component in the $x$ direction (m/h), and
$u_{sy}$ = settling velocity component in the $y$ direction (m/h).

**TABLE 3.6** Lamella equations.

| Parameter | Equation |
|---|---|
| Lamella (Hazen) velocity | $V_l = \dfrac{Q}{A_{tp}}$ |
| Projected area one plate | $A_p = a \bullet b \bullet \cos\Theta$ |
| Total projected area | $A_{tp} = n \bullet a \bullet b \bullet \cos\Theta$ |
| Specific surface area (separation area per unit area) | $A_{sp} = \dfrac{\cos\Theta}{d}$ |

Where

$V_l$ = Lamella (Hazen) velocity (m/h),

$Q$ = influent flow ($m^3/h$),

$A_p$ = projected surface area ($m^2$),

$A_{tp}$ = total projected surface area ($m^2$),

$A_{sp}$ = specific surface area, plate area per unit plan area (unitless or $m^2/m^2$) ,

$\Theta$ = angle of inclination of the plate from the horizontal plane (deg),

$n$ = number of inclined plates,

$a$ = length of single plate (m),

$b$ = width of single plate (m), and

d = vertical separation distance between Lamella (m).

In working with Lamella, the calculated hydraulic overflow rate based on the tank water area must be distinguished from the velocity based on the projected area of the Lamella. Equations 3.8 and 3.9 define the Lamella velocity and overflow rate, respectively. The Lamella overflow rate will be used to report hydraulic loading rates in this chapter unless otherwise stated.

$$\text{Lamella or Hazen velocity} = \frac{Q}{A_{tp}} = \frac{\text{flow}}{\text{total projected surface area}} \tag{3.8}$$

$$\text{Lamella overflow rate} = \frac{Q}{A_s} = \frac{\text{flow}}{\text{tank surface area}} \tag{3.9}$$

For an installation with Lamella with a 55-deg angle of inclination and a Lamella separation of 10 cm, the ratio of the projected area to the water area is approximately 10: 1, and the Hazen velocity will be approximately 1/10 the overflow rate based on the Lamella footprint.

**EXAMPLE 3.1.** A wastewater treatment plant is evaluating the potential increase in settling surface area and clarification for its existing primary clarifiers by retrofitting Lamella into the existing tanks. Estimate the approximate Lamella area that might be placed in each tank and the potential capacity resulting from the retrofit. Assume that each primary clarifier is 15 m wide by 40 m by 4 m side water depth and the sludge-settling velocity is 1.7 m/h. Consider a distance between Lamella (normal to the plane of the Lamella) of 5 and 10 cm. Assuming a distance between Lamella of 10 cm, the ratio of the total Lamella area to the existing water surface area is approximately 8.2: 1 as calculated below. For a distance of 5 cm between Lamella, the ratio is twice as great, approximately 16.4 :1.

| | |
|---|---|
| *a*, length of single Lamella m | 2.1 |
| *b*, width of single Lamella m | 1.3 |
| Θ | 55.0 deg |
| sin (Θ) | 0.819 |
| cos (Θ) | 0.574 |

| | |
|---|---|
| $s$, assigned variable for the perpendicular distance between Lamella | 10.0 cm |
| $h = d/\sin(\Theta)$ (and $h$ = horizontal projection of the Lamella spacing, $s$, and $d$ = vertical separation distance between Lamella) | 12.2 cm |
| no. Lamella/m | 8.2 |
| $A_s$ | 1.5 m$^2$ |
| $A_{tp}$ | 12.5 m$^2$ |
| $A_{tp}/A_s$ | 8.19 |

Although the nominal increase in unit surface area is approximately 8: 1 and 16: 1, in reality the effective increase in settling area can be as much as 70% less.

On a gross basis, the tank settling area could be increased from approximately 600 m$^2$ (15 ft × 40 ft) to approximately 4900 or 9800 m$^2$. However, modifications required to the hydraulics of the settling tank to provide proper feed distribution to the Lamella will reduce the tank area available for Lamella. Using one manufacturer's standard Lamella modules, four longitudinal rows of Lamella approximately 1.25 m wide and 30 m long can be installed the length of the tank with approximately 1.3 m between rows. With these assumptions, Lamella will cover a total of approximately 150 m$^2$ of the tank. This results in total effective Lamella areas of approximately 840 m$^2$ and 1680 m$^2$ for Lamella distances of 5 cm and 10 cm, respectively. At a design overflow rate of 1.4 m/h, the capacity of each original clarifier was approximately 20 200 m$^3$/d. Maintaining the Lamella velocity at 1.4 m/h, the new capacity will be approximately 28 200 or 56 500 m$^3$/d, depending on the Lamella spacing. Checking the settling velocity condition of eq 3.7 shows that the component of the settling velocity in the $y$ direction is greater than the overflow velocity based on the total Lamella area.

$$
\begin{aligned}
&-u_{sy} = u_s \cos\Theta \geq \frac{Q_L}{A_L} \\
&-u_{sy} = 1.7\cos(55) = 0.98\ m/h \\
&\frac{Q_L}{A_L} = \frac{(56,500/24)}{(4\times 16.4\times 30)(1.3)(2.1)} = 0.44\ m/h \qquad (3.10) \\
&0.98 \geq 0.44
\end{aligned}
$$

**CHANGES IN SUSPENDED SOLIDS CONCENTRATION.** Secondary clarifier capacity is a function of clarifier surface area, sludge quality (settling velocity), and the MLSS concentration. Figure 3.8 illustrates the changes in settling velocity that occur as the suspended solids concentration increases in flocculent suspensions (Patry and Takacs, 1992). Reducing the TSS concentration of the aeration tank effluent has a significant effect on the flow capacity of the secondary settling tank. Because the maximum capacity of a secondary clarifier occurs when the clarifier is hydraulically limited, the clarifier flow capacity is inversely proportional to the aeration tank effluent suspended solids concentration. Use of the following formula based on the Vesilind (1968) settling-velocity equation allows the clarifier capacity to be calculated.

$$Q = N \cdot A_S \cdot V_O \cdot \exp\left[-\left(a' + b' \cdot SVI\right) \cdot \left(\frac{X_T}{1000}\right)\right] \cdot \frac{1}{SF} \qquad (3.11)$$

Where

$A_S$ = clarifier surface area ($m^2$),
$V_O$ = Vesilind settling coefficient (m/h),

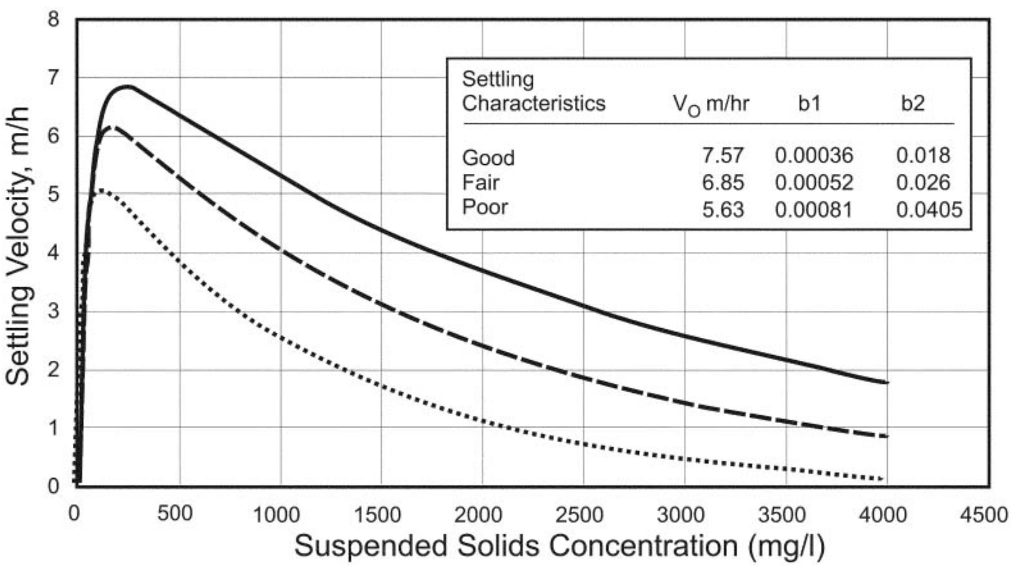


**FIGURE 3.8** Settling velocity model for flocculent suspensions (Takács et al., 1991).

$a'$, $b'$ = Vesilind settling coefficients,
$N$ = number of clarifiers,
$Q$ = clarifier effluent flow ($m^3/d$),
SVI = sludge volume index (mL/g or L/kg),
$X_T$ = MLSS concentration (mg/L),
SF = safety factor on settling velocity.

For any specific plant operating in a specific configuration with a given sludge quality, the clarifier capacity is inversely proportional to the aeration tank suspended solids concentration.

$$Q_T \propto \frac{K}{X_T} \tag{3.12}$$

where $K$ = specific proportionality coefficient.

**CHANGES IN TEMPERATURE.** A change in influent suspended solids concentration is not the only parameter that can affect clarification. Temperature has been shown to have an effect on the hydrodynamics of clarification (McCorquodale, 2001). Variations in incoming salinity can also affect clarifier performance.

The effect of temperature on the hydrodynamics of a clarifier is entirely dependent on the variability, not the actual value, of the incoming wastewater (McCorquodale, 2001). If the wastewater temperature is constant with time, there should be minimal effect on the hydrodynamics of the clarifier. However, if there is significant temporal variation in temperature, for example, wet weather flows during cold weather, then the effect of temperature variation on clarifier hydrodynamics can be dramatic. When the influent temperature is less than the ambient tank temperature, there will be a tendency for a bottom density current to develop. This will be followed by stratification. When the influent temperature is greater than the tank temperature, a buoyant plume and a surface density current may occur.

In municipal secondary clarifiers, the effect of suspended solids typically dominates. There is some evidence that diurnal temperature and heat-transfer effects are important in some systems, especially in cold climates.

The formation of density currents caused by changes in temperature are an unsteady phenomena, and with time the tank temperature will approach that of the influent. This transition has been observed in scale models. The following list details what occurs when the influent temperature to a clarifier is lower than the temperature found within the clarifier (McCorquodale, 2001; McCorquodale et al., 1995; Van Marle and Kranenburg, 1994; Zhou et al., 1994).

- A strong bottom density current (with possible scouring of the blanket) is formed. Also there may be short-circuiting to the sludge hopper.
- A "splash" or runup will occur at the launder, wall with possible washout of solids.
- An internal hydraulic jump occurs that travels from the wall towards the inflow.
- A reflection of this jump occurs back towards the wall.
- Density stratification and displacement of the ambient liquid above the stratification occur. This stage seems to give good solids removal.
- With time, the temperature within the clarifier approaches that of the influent, resulting in the disappearance of the bottom density current.

The following list details what occurs when influent temperature to a clarifier is higher than that found within the clarifier.

- If the inflow temperature rises above the tank temperature, there can be violent turnover in the tank. The influent becomes a rising plume with high turbulence. This is accompanied by a surface density current that short-circuits to the launder. This stage may result in poor solids removal.
- The surface density current affects the clarifier wall and initiates an inverted traveling hydraulic jump moving toward the inflow.
- Stratification is again achieved after a few passages of the internal jumps (waves).

Eventually a uniform temperature is achieved because the lower temperature bottom water is removed by entrainment or by the return activated sludge (RAS).

# TYPES

**CONVENTIONAL PRIMARY TREATMENT.** In conventional, or classic, primary sedimentation, the design and performance of the unit process is based on the natural tendency of the particles in wastewater to agglomerate into larger particles (type 2 settling) and settle from the water under quiescent conditions (Droste, 1997).

Primary sedimentation has long been a staple of municipal wastewater treatment because of its simplicity and proven ability to remove a large percentage of TSS and $BOD_5$ in raw wastewater at a low unit cost.

These same advantages will ensure that classic primary sedimentation will play a role in many wet weather treatment strategies. At the same time, however, there are inherent limitations in classical primary sedimentation, which make it economically unattractive for occasional use. The primary disadvantage is the relatively low settling velocity of many wastewater particles, which translates into relatively large sedimentation tank surface areas and high capital cost if they are used only for occasional extreme flow events.

**RERATED CONVENTIONAL PRIMARY CLARIFICATION.** Studies show that primary clarifiers typically remove a significant fraction of settable solids in raw wastewater and that performance is only weakly related to tank hydraulic overflow rate (Wahlberg et al., 1997). During storm events, particle-settling velocities in wastewater may increase and, depending on the magnitude and duration of a storm, the suspended solids concentration may decrease because of dilution by infiltration and inflow. This implies that higher flow rates can be tolerated through primary clarifiers during storm events without a significant increase in effluent suspended solids unless there is a concurrent increase in nonsettleable solids concentration. Standards for peak overflow rates for primary clarifiers in most traditional design guidelines range from approximately 2.0 to 5.0 m/h (GLUMB, 1997; Metcalf and Eddy, 2003; U.S. EPA, 1975b; WEF, 1998b) . Demonstrating that a clarifier operates satisfactorily at a velocity of 5.0 m/h or higher during intermittent peak flows as opposed to 2.0 m/h means a substantial difference in wet weather treatment capacity. This highlights the importance of quantifying the expected performance of primary clarifiers based on settling velocity distributions of real wet weather suspended solids or by full-scale testing during actual storm events.

**CHEMICALLY ENHANCED PRIMARY TREATMENT.** Chemically enhanced primary treatment, whereby wastewater is chemically coagulated before clarification is the simplest enhancement that can be made to conventional primary clarification to increase treatment capacity. Chemical coagulants such as ferric chloride and alum (typically <60 mg/L) provide cations that destabilize colloidal particles in wastewater while flocculent aids such as polymer (typically <2 mg/L), recycled sludge, and microsand function to accelerate the growth of floc, enlarge the floc, improve floc shape, strengthen floc structure, and increase particle specific gravity. The use of chemicals allows a higher peak overflow rate during peak flow events while maintaining or increasing primary clarifier performance, thus minimizing the clarifier surface area that must be provided for peak flows. Chemically enhanced primary treatment can be a full-time treatment method; however, when used for control of wet weather flows its use is limited to peak wet weather periods. Figure 3.9 shows typical ranges of TSS removal for conventional primary sedimentation and CEPT versus overflow rate (CDM, Inc., and Montgomery Watson, 1995).

Chemically enhanced primary treatment has evolved over time. Early applications typically consisted of simply adding ferric, alum, or lime to a conventionally designed primary settling tank. Current practice uses smaller metal salt doses (20 to 40 mg/L) in combination with polymer addition (<1 mg/L) and includes the use of rapid mix and flocculation before the settling tank. Use of iron salts can decrease the efficiency of downstream disinfection with UV light. As a result, Canadian researchers investigated

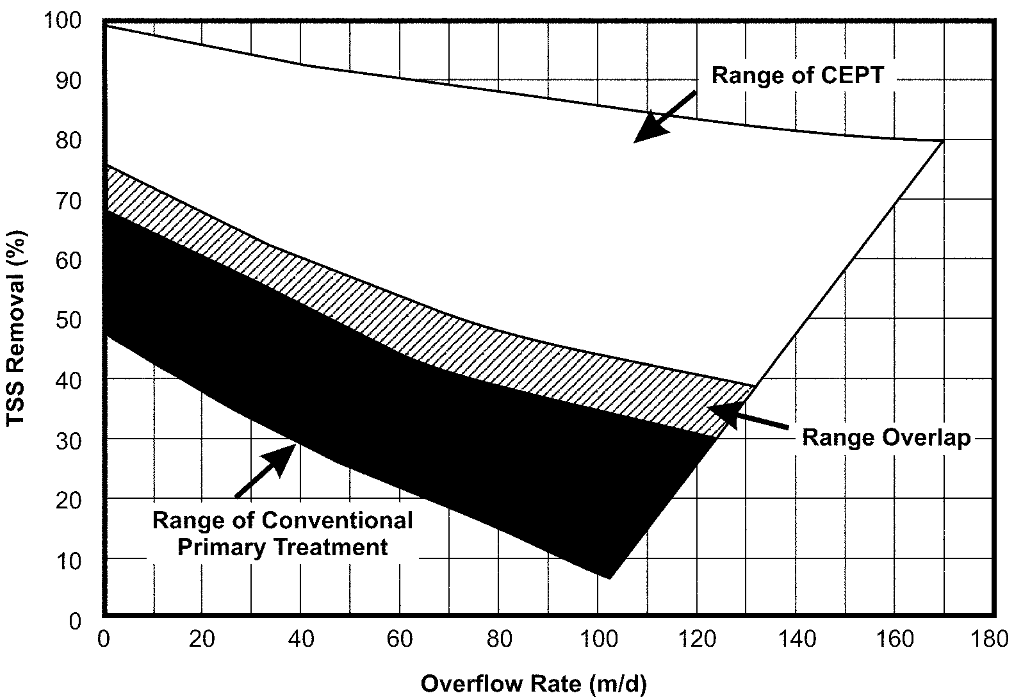


**Figure 3.9** Range of TSS removal with conventional and CEPT (CDM/Montgomery Watson, 1995).

the use of high polymer doses (~8 mg/L) and discovered that polymer-only coagulation resulted in improved removal of suspended solids at higher overflow rates than coagulation with ferric chloride and polymer (Averill et al., 1999). The effect of basin geometry on solids-removal efficiency was also reduced. It remains to be determined if these findings are applicable in other locations. While CEPT can be practiced by simply adding chemicals to grit tanks and primary clarifier influent channels, optimum performance depends on adequate coagulation before sedimentation. Jar testing is essential for determining design chemicals, doses, and rapid mix and flocculation times (Hudson and Wagner, 1981; Wagner, 2004; Wahlberg et al., 1999; Yu, 2000).

Published hydraulic loading rates are rare for CEPT. U.S. EPA (1975a) suggests 2 m/h at average flow to 4 m/h at peak flows for CEPT with lime addition. More recent work for Deer Island and Hong Kong (CDM, Inc., and Montgomery Watson, 1995) suggests that an annual average loading rate approximately 3 m/h is possible.

Some advantages and disadvantages to CEPT are summarized in Table 3.7. These advantages and disadvantages apply to all advanced or high-rate clarification processes used for primary treatment.

**RETENTION TREATMENT BASINS.** Wet weather flow storage tanks can be designed to also operate as clarifiers once the storage volume fills and overflow from the tank occurs. Such units have been called retention treatment basins (RTBs) and are defined as any vessels that provide some storage and treatment when operating with wastewater flowing through the unit (Schraa et al., 2004). Both sludge and floating solids are typically returned to the main stream flow to be treated by the main treatment facility. Excess flow from small storm events will be completely retained while overflows from large storms will be treated and discharged or blended. As with other types of clarification, the performance of RTBs can be enhanced with chemical coagulation. Work in Canada has focused on the use of vortex separators and rectangular tanks with high polymer doses; however, many applications operate with other tank configurations and without chemical addition, although at reduced efficiency. Pilot studies performed using a vortex separator with ferric and polymer and then just polymer were followed with a full-scale test in a rectangular basin (Averill et al., 1999). With traditional ferric and polymer coagulation, removal efficiencies in the pilot unit decreased rapidly with increasing overflow velocity from approximately 70 to 80% at 5 m/h to less than 20% at 20 m/h. In contrast, the use of a high polymer dose resulted in removals that ranged between 60 and 80% at 10 m/h to more than 50% at 30 m/h. Results from the full-scale test with a high polymer dose in a rectangular tank were similar to the pilot-scale results.

**Table 3.7** Advantages and disadvantages for advanced primary treatment (Morrisey and Harleman, 1992; Murcott and Harleman, 1992; Reardon, 1995).

| Advantages | Disadvantages |
|---|---|
| Increased removal of BOD, TSS, phosphorus, and metals | Requires chemical addition |
| Reduces size/cost or increases capacity of biological secondary | Produces increased quantities of sludge |
| Enhanced biological treatment kinetics; performance | Reduces alkalinity |
| Allows primary tanks to be designed for higher overflow rates | Safety concerns with respect to chemical handling |
| Smaller treatment plant footprint | May decrease sludge settleability in secondary system |
| Ability to absorb shock loadings/wet weather flows | Sludge is not as easy to dewater as conventional primary |
| Can provide odor and corrosion control | Chemicals may have adverse effect on plant aesthetics (staining) |
| Decreases carbon-to-nitrogen ratio in primary effluent, increasing the fraction of nitrifiers in the MLSS and enhancing ammonia removal | Metal salts may remove too much phosphorus, making the primary effluent nutrient deficient |

Dynamic simulations conducted as part of the same project showed that a RTB with a high polymer dose required less tank volume than either unaided setting or storage alone to meet the Ontario CSO control guidelines (Schraa et al., 2004).

**LAMELLA (PLATE OR TUBE) CLARIFIERS.** Plates or tubes may be used to improve clarification with or without chemicals. Total suspended solids and $BOD_5$-removal efficiency in Lamella clarifiers is reported to be similar to that obtainable with conventional primary clarifiers operating at the same overflow rate based on projected area (35 to 40% for $BOD_5$ and 50 to 60% for TSS) (Dudley et al., 1994). Limited data are available on TSS- and $BOD_5$-removal efficiency for Lamella preceded by chemical coagulation; however, it is reasonable to expect that this too will be similar to conventional CEPT at comparable overflow rates.

Lamella systems typically consist of inclined parallel metal plates or bundles of hexagonal plastic tubes installed at the surface of the settling tank to a vertical depth of approximately 2 m. Inclined plates or tubes installed at an angle of 45 to 60 deg and spaced at intervals of 40 to 120 mm increase the effective settling surface area by a factor of 6 to 12, thereby allowing a higher peak flow to be treated in a given tank surface area. Decreasing the Lamella angle increases the total settling area; however, when the Lamella angle is too shallow, the settled solids do not slide down the Lamella surface, and periodic cessation of flow (possibly with back flushing) is necessary to remove sludge (Ross et al., 1999). Space must also be provided between the Lamella for the movement of both water and sludge. Decreasing the Lamella spacing also increases the total settling area, but a minimum spacing is established by the critical velocity above which turbulence and the risk of solids resuspension increase (Dudley et al., 1994). It has been recommended that the flow in the Lamella have a Reynolds number less than 2000, a Froude number greater than $10^{-5}$, and detention times longer than 3 to 5 minutes to obtain good settling conditions (Fischerstrom, 1955). In calculating the Reynolds number ($N_{\mathrm{Re}}$), the hydraulic radius ($R$, cross-sectional area/wetted area of the Lamella) should be used in the equation $N_{\mathrm{Re}} = VR/\nu$, where $V$ is the velocity of water and $\nu$ is the kinematic viscosity. With this definition, laminar flow occurs at Reynolds numbers less than 500. Within the context of flow in a sedimentation basin, the Froude number ($N_{\mathrm{Fr}}$)has been defined as $N_{Fr} = V^2/(R \cdot g)$, where $R$ is again the hydraulic radius and $g$ is the gravitational constant, and has been used to indicate the stability of flow (Fischerstrom, 1955).

Countercurrent designs are the most common flow pattern in use and are reported to be less expensive to install and operate (Dudley et al., 1994). In a countercurrent flow pattern, the influent is fed under the plates or tubes and flow is upwards in the channels formed by adjacent Lamella. Solids settle onto the top surface of the lower Lamella of each channel and slide down the Lamella surface. To provide the widest and most economical Lamella width while still maintaining good flow distribution, flow must be fed from both sides of the Lamella. For wastewater applications, influent is typically fed longitudinally through inlet ports located below the Lamella to provide better flow distribution. Other possible flow patterns are shown in Figure 3.10 (Buer et al., 2000). In cocurrent configurations, the flow is fed on top of the Lamella, and both water and solids flow downwards. For cross-flow patterns, the water moves horizontally between the Lamella whereas the sludge again flows downward. Cocurrent designs require particles with high settling velocities to avoid sludge reentrainment whereas cross-patterns may be used when floating and settling material must be removed (Dudley et al., 1994).

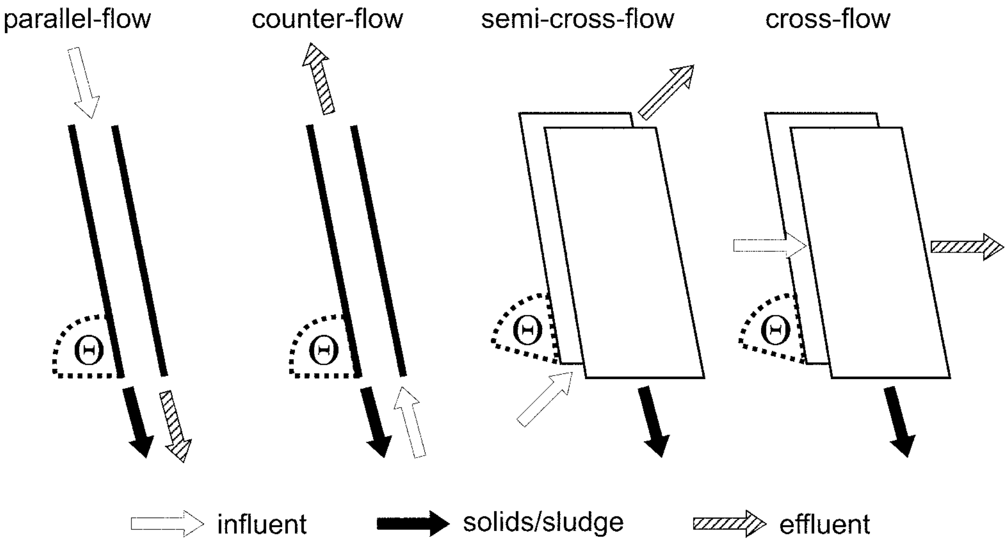


**FIGURE 3.10** Plate settler flow patterns (Buer et al., 2000).

Lamella have the ability to increase the capacity of an existing clarifier or reduce the land area required for new ones. In wet weather applications, the use of Lamella settlers reduces the cost and space requirements to construct clarifiers for peak wet weather flows. Reduced contact area between wastewater and the atmosphere facilitates odor control, whereas the reduced footprint may allow the facility to be enclosed to increase aesthetic effects for the community. For primary treatment with Lamella but without chemical addition, suggested design hydraulic loading rates are 10 to 15 m/h at peak flow. The use of Lamella requires fine screening and satisfactory grit and grease removal before the Lamella tanks. Although it has not been explicitly stated, inspection of existing facilities suggests that Lamella function best when enclosed to eliminate clogging from blowing debris and algae growth. Besides the potential for clogging of the Lamella, other concerns associated with the use of Lamella include the increased need for reasonably uniform water distribution to each channel, low (laminar) flow velocities and uniform flow distribution within each channel, and collection of the sludge while preventing resuspension (Ross et al., 1999). Maintenance requirements are expected to be higher for Lamella clarifiers because of the need for regular cleaning of the Lamella. Provision of Lamella that are independently supported, easy access to the Lamella for cleaning, and Lamella that can be individually removed have been reported to facilitate maintenance (Dudley et al., 1994). Another reported disadvantage to Lamella settlers is the production of a more dilute sludge that may increase the cost of sludge handling.

Lamella clarifiers have not been commonly used in wastewater applications in the United States. This is not the case for Europe, and especially France, where they have been used more frequently. Approximately 130 full-scale wastewater facilities with plate or Lamella settlers were identified from the reference lists of three manufacturers of plate equipment. A summary of these installations is provided in Table 3.8. Plates or tubes are used to enhance primary treatment in most of these

**TABLE 3.8** Summary of wastewater facilities with plate and tube settlers identified from manufacturer reference lists.

| Country | Number of installations | | Design flow* ($m^3/d$) | Selected facility locations |
|---|---|---|---|---|
| | Primary | Secondary | | |
| Australia | 1 | 0 | NA | Woodmans Point |
| Austria | 0 | 1 | NA | Ara Naunders |
| Belgium | 1 | 0 | 70 000 | Malmedy |
| Canada | 10 | 0 | 70 000–1 360 000 | Longueil–Montréal Québec East |
| China | 0 | 1 | NA | Hong Kong Stanley |
| France | 88 | 0 | 5000–1 700 000 | Marseille, Bordeaux Toulon Est, Paris Columbes |
| Germany | 2 | 1 | 275 000 | Hereford |
| Greece | 1 | 0 | NA | Rhodos |
| Italy | 1 | 2 | 110 000 | Roma Sud |
| Japan | 3 | 0 | 4800 | Chiba (stacked) |
| Mexico | 0 | 1 | NA | Toluca City Brewery |
| Portugal | 2 | 0 | 20 000–255 000 | Lisbone |
| Spain | 4 | 0 | 18 000 | Saragosse |
| Sweden | 4 | 0 | 3840–30 000 | Karlstad |
| Switzerland | 6 | 0 | 2160–100 000 | Vevey–Montreux |
| United Kingdom | 7 | 0 | 100 000–216 000 | Brighton |
| Total | 130 | 6 | 2160–1 700 000 | |

* Design flows provided in manufacturer reference lists.

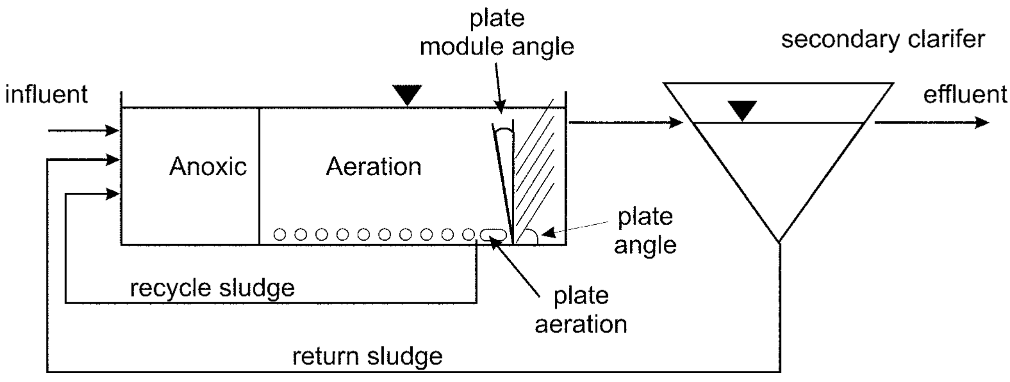


**FIGURE 3.11** Use of plates in the aeration tank to presettle MLSS (from Buer et al., 2000).

facilities and a large majority is located in Western Europe with more than one-half of these in France. Design flows ranging from 3100 to 1 700 000 $m^3/d$ have been reported. Approximately 90% of the reported installations have a design flow of less than 200 000 $m^3/d$.

While the classic location for Lamella is in primary clarifiers, researchers in Germany have investigated their use at the end of the aeration tanks or at the entrance to secondary settling tanks. Lamella in either of these locations reduce the MLSS concentration entering secondary settling tanks, thereby increasing the peak flow capacity of the secondary settling tanks (Buer, 2002). This is illustrated in Figures 3.11 and 3.12, respectively.

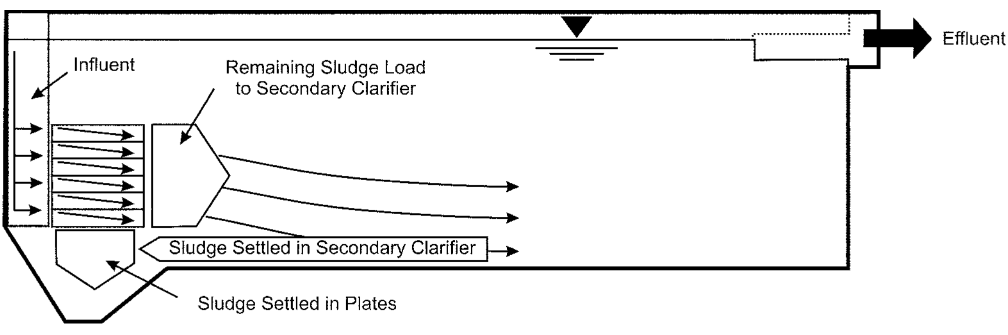


**FIGURE 3.12** Use of plates to presettle MLSS (from Buer et al., 2000).

**HIGH-RATE CLARIFICATION PROCESS.** Figure 3.13 illustrates the use of high-rate clarification processes (e.g., dense sludge and ballasted flocculation) to treat peak wet weather flows. High-rate clarification processes are well suited for wet weather clarification applications because of reduced space requirements; rapid startup and response times; relative insensitivity to fluctuations in raw water quality; and improved removal of TSS, BOD, total Kjeldahl nitrogen, total phosphorus, and metals. Because high-rate clarification facilities for wet weather flow may only be used several times per year, several plants have located high-rate clarification after the biological treatment process as shown in Figure 3.14, where it can also be used for tertiary suspended solids or phosphorus removal during dry weather periods.

The primary disadvantage of high-rate clarification is the increased doses of metal salts and polymer required to operate the process. This increases annual operating costs; however, if the process is only used to treat peak wet weather flows, the total operating time during a year is relatively small and the additional chemical costs are acceptable. Another disadvantage associated with high-rate clarification processes is the use of hydrocyclones and plates or tubes, which require fine screens

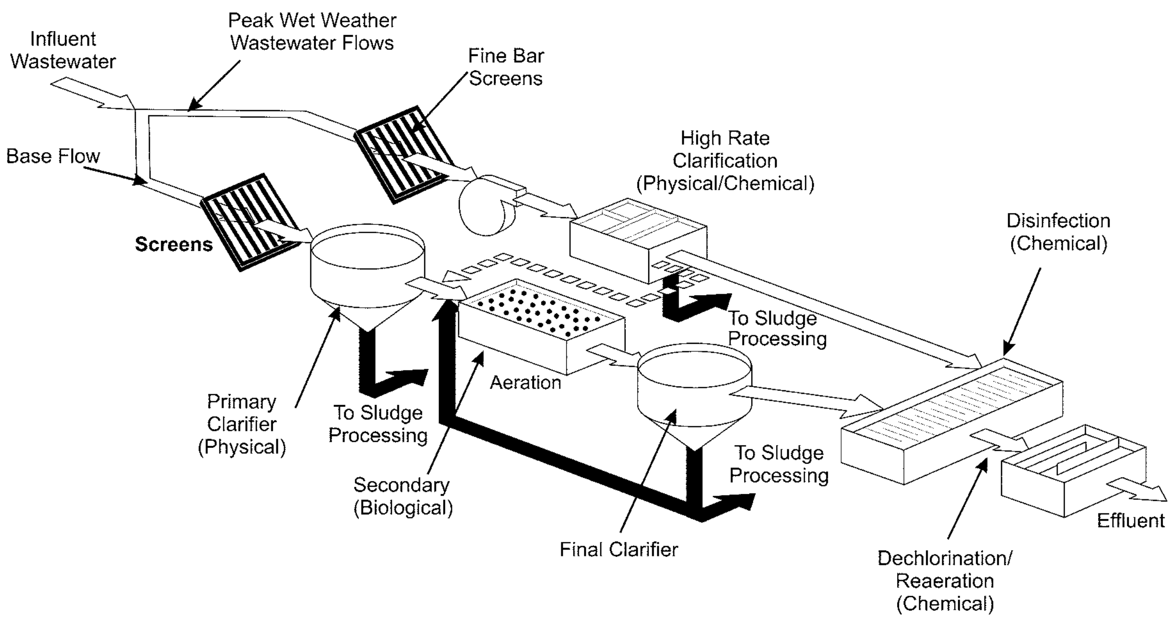


FIGURE 3.13 Use of high-rate clarification to treat peak wet weather flows.

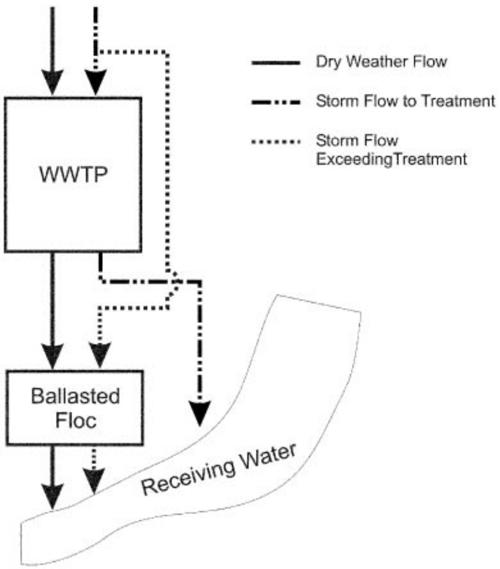


**FIGURE 3.14** Dual use of high-rate clarification.

before the process. High-rate clarification processes using sand may also experience higher wear rates for pumps and piping moving sludge and sand.

Two different types of high-rate clarification processes, sometimes referred to as the dense sludge process and the ballasted flocculation process, are in common use (Metcalf and Eddy, 2003). Dense sludge is a high-rate clarification process that combines chemical coagulation, sludge recirculation, tube settling, thickening, and sludge recycle. Ballasted flocculation refers to high-rate clarification processes that increase particle size and density, hence settling velocity, by binding solids to a weighting agent or "ballast" with metal hydroxide floc and polymer. Very small sand particles (microsand) are the most common ballast although other high-density materials (sp gr ~2.65 or higher) with fine particle sizes have been used. Coagulation for dense sludge and ballasted flocculation processes is accomplished in a similar manner as with conventional processes. Both processes are currently proprietary. Infilco Degrémont (Richmond, Virginia) markets the dense sludge process in the United States under the trade name DensaDeg® and Krüger, Inc.—A Veolia Water Systems Company (Cary, North Carolina) markets the ballasted flocculation process as the Actiflo® process.

Rapid mixing design procedures for dense sludge and ballasted flocculation are similar to that used for conventional sedimentation with $G$ values of 200 to 300 $sec^{-1}$ and hydraulic detention times of 30 to 60 seconds. Flocculation times for both the dense sludge process and ballasted flocculation are significantly less than for conventional designs (~12 minutes). Hydraulic retention times for ballasted flocculation are in the range of 1 to 3 minutes, resulting in $Gt$ ($G$ = average velocity gradient [L/s] and $t$ = hydraulic detention time [s]) values of 6000 to 20 000. Dense sludge flocculation times are approximately 4 minutes in the draft tube reactor and 1.5 minutes in the transition zone. There is an optimum mixing intensity in the flocculation zone (or maturation tank) that keeps the floc in suspension but does not shear newly formed floc. Rapid mix and flocculation design values for a number of ballasted flocculation projects are summarized in Table 3.9.

Designs for rapid mix and flocculation basins for dense sludge and ballasted flocculation processes have several variations, including the use of two and three zones for rapid mix and flocculation. The use of three reactors follows the conclusions of Desbos et al. (1990), who found that the use of plug-flow, a reduced coagulant dose, and a higher energy input can reduce the overall coagulation/flocculation reactor

**TABLE 3.9** Ballasted flocculation design criteria.

| | **Coagulation tank** | | **Injection tank** | | **Maturation tank** | |
|---|---|---|---|---|---|---|
| **Project** | **HDT[a] (min)** | **$G$ ($s^{-1}$)** | **HDT (min)** | **$G$ ($s^{-1}$)** | **HDT (min)** | **$G$ ($s^{-1}$)** |
| Bremerton Pine Road | — | — | 1.0 | 290 | 3.0 | 210 |
| Bremerton Westside | — | — | 1.0 | 290 | 3.0 | 210 |
| Ft. Worth | 0.7 | — | 1.7 | 455 | 4.8 | 210 |
| Lawrence | 0.8 | 265 | 1.0 | 265 | 3.0 | 190 |
| New York | 1.0 | 300 | — | — | 3.0 | 260 |
| CCWS[b] | 1.0 | 300 | — | — | 3.0 | 250 |

[a]HDT = hydraulic detention time.
[b]CCWS = Cobb County Water System.

size and operating cost. As with conventional systems, rapid mix and flocculation for dense sludge and ballasted flocculation is best based on jar and pilot testing.

Coagulant doses for high-rate clarification in wastewater applications typically range from 40 to 125 mg/L with ferric chloride, 80 to 85 mg/L with ferric sulfate, 60 to 70 mg/L with alum, and 45 to 100 mg/L with polyaluminum chloride. Doses of 0.9 to 1.2 mg/L of a high-molecular-weight anionic polymer are common. Several studies (CDM, 1999; Keller et al., 2001; Moffa et al., 2000) report that the performance of ballasted flocculation improves with increasing coagulant addition up to a point with no incremental increases in performance for subsequent increases in dose. Coagulant and polymer doses from several pilot plant studies are summarized in Table 3.10.

Reported sludge concentrations from ballasted flocculation vary. Guibelin et al. (1994) reported TSS concentrations of 2 to 8 g/L. Sawey et al. (1999) reported that the

**Table 3.10** Coagulant and polymer concentrations reported for ballasted flocculation tests (modified from Young and Edwards, 2000).

| Project | Coagulant dose (mg/L) | Polymer dose (mg/L) | Reference |
|---|---|---|---|
| Galveston, Texas | 75–125 | 1.0–1.15 | Chang et al., 1998 |
| Cincinnati, Ohio | 45–100 | 1.0–1.30 | Chang et al., 1998 |
| Tucson, Arizona | 110 | 0.5 | Chang et al., 1998 |
| Mexico Valley, Mexico | 180 | 1.0 | USFilter Co., 1997–1998 (Sullivan, 2002) |
| Jefferson County, Alabama | 40, 70 | 1.1 | USFilter Co., 1997–1998 (Sullivan, 2002) |
| Fort Worth, Texas | 70–125 | 0.75–1.0 | CDM, 1999 |
| Fort Worth, Texas | 150 | 10 | USFilter Co., 1997–1998 (Sullivan, 2002) |
| Fort Smith, Arkansas | 100 | 1.0 | USFilter Co., 1997–1998 (Sullivan, 2002) |
| New Park, Kentucky | 180 | 0.4 | USFilter Co., 1997–1998 (Sullivan, 2002) |

sludge volume was typically approximately 1% of the forward flow, with TSS concentrations between 3 and 5 g/L. Testing at San Francisco (Jolis and Ahmad, 2001) found that sludge concentrations were between 5 and 7 g/L, with the volatiles fraction at approximately 60% and the sludge volume at approximately 4 to 5% of the flow. Scruggs and Wallis-Lage (2001) found that sludge concentrations ranged from approximately 2.5 to 2.8 g/L and that up to one-third of the sludge mass was a result of chemical addition. Pilot testing at the Fort Worth, Texas, Village Creek plant (CDM, 1999) included significant evaluations of sludge quantity and quality. This is summarized in Table 3.11.

Reported solids concentrations from a number of dense sludge pilot studies are summarized in Table 3.12. Concentrations have ranged from 0.5 to 7%, with most values between approximately 2 and 5%.

Startup and shutdown of high-rate clarification in wet weather applications requires special attention because of their intermittent operation, the use of chemicals, and the presence of sludge and sand in the reactor basins (Keller et al., 2001). Polymers, in particular, often must be aged or activated before use and then may only remain active for a limited period of time. Because wet weather events cannot be anticipated, polymer-feed solutions must be made up and replaced on a regular basis, whether or not they are used. Ballast or dense sludge is an integral component of high-rate clarification, and a substantial inventory (tonnes) exists in the reactors while the process is in operation. Design and operating decisions are required on

**TABLE 3.11** Average clarifier sludge concentrations (CDM, 1999).

| Process | Sludge concentration (%) | VSS/TSS ratio |
|---|---|---|
| Infilco-Degrémont (Richmond, Virginia) DensaDeg 4D® | 2.98 | 0.71 |
| USFilter/Kruger (Cary, North Carolina) Actiflo® | 0.32 | 0.61 |
| Parkson (Ft. Lauderdale, Florida) Lamella® clarifier | 2.91 | 0.61 |
| USFilter (Cary, North Carolina) Microsep® | 0.38 | 0.54 |

**Table 3.12** Sludge characteristics reported during dense sludge CSO/SSO pilot studies.

| | Dry solids (%) | | Volatile solids (%) | |
|---|---|---|---|---|
| **Location** | **Minimum** | **Maximum** | **Minimum** | **Maximum** |
| Compiegne, France | 4.0 | 7.0 | | |
| Douai, France | 4.5 | 5.7 | 43 | 57 |
| Nice, France | 3.5 | 6.0 | 45 | 60 |
| Mexico City, Mexico | 4.5 | 9.0 | 40 | 55 |
| Fort Worth, Texas | 0.6 | 6.6 | 55 | 82 |
| New York City, New York | 0.6 | 3.9 | | |
| San Francisco, California ($FeCl_3$[a]) | 1.3 | 3.3 | 64 | 79 |
| San Francisco, California (PACl[b]) | 0.52 | 1.3 | 56 | 69 |
| Bremerton, Washington | 1.3 | 2.5 | | |
| Little Rock, Arkansas | 1.1 | 2.4 | | |
| Sydney, Australia | 2.8 | 6.1 | 75 | |
| Salem, Oregon | 0.6 | 1.7 | | |

[a]$FeCl_3$ = ferric chloride.
[b]PACl = polyaluminum chloride.

how this inventory is created at startup and maintained or removed at shutdown. Finally, in cold climates, freezing of reactor contents will prevent system operation.

***Dense Sludge Process.*** A typical dense sludge installation consists of influent screening, rapid mix, and flocculation followed by clarification and thickening with external sludge recirculation. Alternate designs include grit removal at the beginning of the process and grease and scum removal after flocculation (see Figure 3.15). Coagulant is added in the rapid mix zone and a polymer is added in the flocculation zone. Fine screens (maximum opening of approximately 10 mm) are needed to remove large solids that might clog the tubes in the settling zone. A portion of the settled sludge (2 to 6% of flow) is recycled to the bottom of the flocculation zone. By increasing the number of particles in the water, sludge recirculation

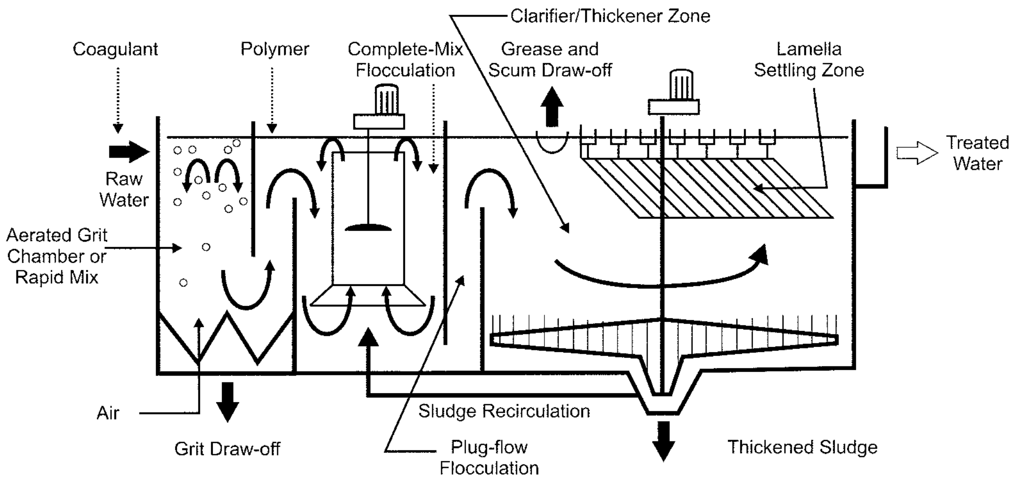


**FIGURE 3.15** Dense sludge process schematic.

increases the rate of flocculation and by increasing the particle densities increases the particle-settling velocities. A unique aspect of the dense sludge process is the use of a draft-tube mixer to create a complete-mix flocculation zone. A plug-flow transition zone follows complete-mix flocculation to further condition the floc. Chemical coagulation combined with sludge recycle forms denser floc particles that settle rapidly in the clarifier/thickener. The thickener provides the mechanism and produces the sludge that is returned to the reactor turbine. Reported values of the sludge solids concentration range between 5 and 90 g/L; however, under optimum conditions, the dense sludge process typically achieves sludge concentrations of 40 to 60 g/L. As a result, the sludge volume produced with the dense sludge process is significantly less than with the ballasted flocculation process. A volatile solids concentration in the sludge of 40 to 60% is typical. Sludge is discharged intermittently from the dense sludge process.

Early dense sludge processes were designed for peak hydraulic loading rates of 25 m/h. Improvements made to the process, including the use of a deeper clarification/thickening zone and two injection points for polymer, allow the dense sludge process to treat significantly higher flow rates. Design hydraulic loading rates for the dense sludge process are now typically 100 m/h under peak conditions (based on the horizontal footprint of the tube section). Suspended solids removals of 85% are expected at design conditions. Tubes are used to improve clarification by removing straggler floc and imposing an additional hydraulic head loss that reduces the formation of turbidity currents and short-circuiting.

When a dense sludge process is started dry, full efficiency is attained within 20 to 30 minutes (Westrelin and Bourdelot, 2001). When started wet, full efficiency is reached almost immediately. After wet weather operations cease, operators can choose to leave a dense sludge unit full of water and sludge for some period of time in case another high flow period occurs, remove the sludge and drain the unit, or remove the sludge and refill it with effluent or potable water. Care must be taken to avoid septic conditions and the resulting increased potential for odors and corrosion. To prevent freezing, the in-tank water temperature can be monitored and the signal used to initiate draining of the unit and refilling with effluent (Keller et al., 2001).

While the dense sludge process is a versatile clarification process that has been successfully used worldwide for water, wastewater and CSO applications, it has seen limited use in wastewater clarification applications for wet weather flows in the United States. However, several dense sludge facilities are under design and construction in North America, including a 636 000-$m^3$/d facility to treat peak flows during wet weather at the Bayview treatment plant in Toledo, Ohio. Existing full-scale dense sludge processes treating wastewater are summarized in Table 3.13.

Two notable full-scale applications are the 40 000-$m^3$/d Clos de Hilde Wastewater Treatment Facility in Bordeaux, France, which began operation in February 1994, and the 240 000-$m^3$/d (1 000 000-$m^3$/d peak flow) Colombes Wastewater Treatment Facility in Paris that has been in operation since 1998. Both facilities use the dense sludge process to provide full-time primary treatment before biological treatment. Overflow rates vary from approximately 7 m/h at average flows to 30 m/h at peak flows in Bordeaux and from approximately 5 m/h at average to 30 m/h at peak in Colombes. Typical $BOD_5$ and TSS removals at the Bordeaux facility are reported to be approximately 70% for both parameters whereas TSS removal at Colombes averages approximately 80%. The sludge concentration at Colombes averages approximately 5 to 6% dry solids. Peak wet weathers flows are treated in the dense sludge process but bypass the biological treatment units at both plants. Both plants use biological aerated filters after the dense sludge process. During dry weather the biological aerated filters at Columbes operate in series to provide carbon and nitrogen removal; however, during wet weather all three sets of filters are operated in parallel.

**TABLE 3.13** Wastewater applications for the dense sludge process.

| Country | Number of installations: Primary | Tertiary | CSO | Design flow ($m^3$/d) | Selected facility locations |
|---|---|---|---|---|---|
| Andorra | 1 | — | — | 7600 | Pas De La Casa |
| Belgium | 1 | — | — | 18 900 | Malmedy |
| Canada | 6 | — | — | 53 000–636 000 | Sherbrooke, Quebec; Laval, Quebec |
| France | 24 | 11 | 7 | 1100–1 050 000 | Aix-En-Provence, Columbes, Metz |
| Germany | 2 | — | — | 760–12 100 | Berlin, Hamburg |
| India | 1 | — | — | — | Bangalore |
| Italy | 1 | 1 | — | 7600–51 100 | Pulsano, Comodepur |
| Mexico | 4 | — | — | 43 500–129 000 | Puebla (D'Atoyac Sur) |
| Spain | 5 | — | — | 1900–5700 | Sarrio Uranga, Tolosa |
| Switzerland | 4 | — | — | 2300–28 800 | Bagnes, Nyon |
| United Kingdom | 2 | — | 1 | 15 100–64 400 | Poole, Edimbourg |
| United States | — | 5 | — | 7600–120 000 | San Rafael, Breckenridge |
| Total | 51 | 17 | 8 | 760–1 050 000 | |

***Ballasted Flocculation.*** Ballasted flocculation is the generic term for a high-rate clarification process that adds fine sand along with metal salts and polymer to wastewater during coagulation and flocculation. Sand provides two significant benefits. First, it is incorporated to floc particles, which dramatically increases the specific gravity and settling velocity of floc particles. Second, the sand increases the number and size of the particles in the water, which has a positive effect on flocculation kinetics. The benefits of the ballasted flocculation process are very similar to those of the dense sludge process, with the principal difference being the higher overflow rates possible with ballasted flocculation. For microsand systems, the design hydraulic loading rates are stated to be 30 to 50 m/h at average flows and as high as 100 to 130 m/h at peak flow.

The ballasted flocculation process typically consists of influent screening, rapid mixing, and flocculation, clarification with Lamella, and sand stripping and recirculation (see Figure 3.16). As with the dense sludge process, the process should be preceded by fine screens to minimize clogging of the plates or cyclones. After screening, a coagulant (typically ferric chloride) is added to destabilize the wastewater, followed by the addition of fine sand and polymer to enlarge and weight the floc, flocculation, and a settling zone with Lamella. The sludge is passed through a hydrocyclone to recover the sand, which is returned to the process while the sludge is directed to further treatment. Slightly different coagulation terminology is used to describe ballasted flocculation than to describe conventional coagulation and sedimentation processes. The origin of this terminology is not clear, but it might result from a combination of translation from the original

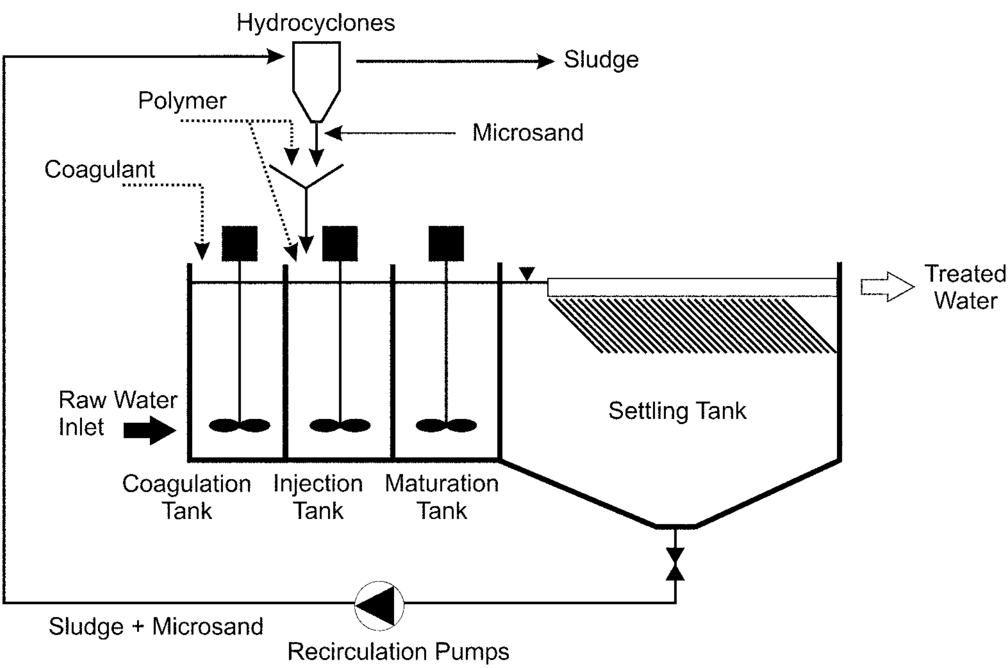


**FIGURE 3.16** Ballasted flocculation process schematic.

French terminology and marketing. The rapid mix tank is typically called the coagulation tank, and the flocculation tank is called the maturation tank. When a second rapid mix zone is used for the addition of sand and polymer, it has frequently been called the injection tank.

Three Hungarian inventors first patented the addition of fine sand, or microsand, to water to enhance sedimentation in 1964. The patent rights were acquired by the French company Omnium de Traitements et de Valorisation (Saint Maurice cedex, France), who further developed and refined the process. The process was originally used for physical–chemical treatment of surface water under the trade name Cyclofloc®, with surface overflow rates up to 8 m/h. The process was improved with the addition of Lamella and a fluidized sand bed, was named the Fluorapid® process, and was used at overflow velocities up to 15 m/h. The process was further improved with the addition of separate tanks for coagulation flocculation and settling, with peak overflow rates between 70 and 130 m/h. Use of the process has expanded from water to wastewater applications and, in 2002, the first CSO installation in the United States began operation in Bremerton, Washington.

Sand used in the ballasted flocculation can vary in size from smaller than 40 μm to 300 μm. Work by Sibony (1981) evaluated five different sand sizes and reported that the best performance (lowest effluent turbidity) was obtained with sand ranging from 40 to 60 μm. Studies by Young and Edwards (2000) evaluated four sand sizes ranging from 44 to 500 μm. The lowest effluent turbidity was obtained with the largest sand size (210 to 300 μm). In both the Sibony (1981) and Young and Edwards (2000) works, the difference in performance between sand sizes was not great. This suggests that, though the sand size could be a variable to be considered in design (a trade-off between cost and performance), the selection of sand should be based primarily on price and availability. Similarly ballasted flocculation performance has not been found to vary significantly with sand dose above a certain value. An upper limit exists to the amount of sand that can be incorporated to the floc, and any additional sand above this dose contributes little to the treatment process. Typical sand doses range from 1 to 12 g/L with a makeup dose of approximately 1to 3 mg/L.

Startup and shutdown sequences for ballasted flocculation are similar to that of the dense sludge process, with the added complication that the reactor contains a significant mass of sand and attached chemicals and sludge. When the process is inactive, the sand will settle to the bottom of the reactors. When the process is restarted the sand must be resuspended without losing any significant quantity in the effluent. After shutdown, one approach is to pump the sludge from the clarification tank through the cyclones to separate the sludge for processing while returning the sand to the rapid mix tanks for storage until the next use (Keller et al., 2001).

**AERATION TANK SETTLING.** Aeration tank settling is a term used for the practice of turning off the air to all parts or just toward the end of the aeration tank during peak flows as illustrated in Figure 3.17. Without aeration, the MLSS begin to settle in the aeration tank, and the solids concentration sent to the secondary settling tanks is reduced. By reducing the suspended solids concentration during peak flow, the sludge-settling velocity is increased and the clarifier capacity increased when it is most needed. Plants reported to use some form of settling in

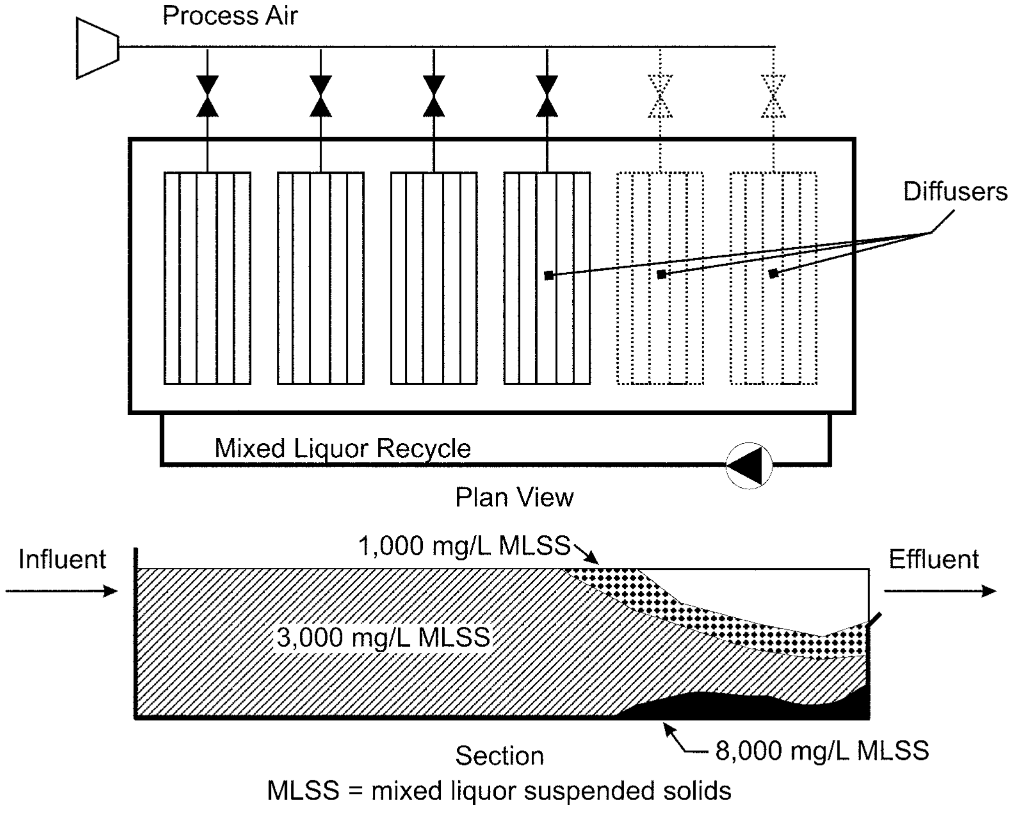


FIGURE 3.17 Aeration tank settling (reprinted from *Water Sci. Technol*, **41** (9), 179–184, with permission from the copyright holder, IWA).

**TABLE 3.14** Existing facilities using aeration tank settling.

| Plant | Population equivalent | In operation |
|---|---|---|
| Marshfield, Wisconsin | | |
| Half Moon Bay, California | | |
| Aalborg West, Denmark | 265 000 | 1992 |
| Aalborg East, Denmark | 75 000 | 1995 |
| Gässlösa (Borås, Sweden) | 130 000 | 1998 |
| Bjergmarken, Denmark | 80 000 | 1999 |
| Hirtshals, Denmark | 53 000 | 2001 |
| Lundtofte, Denmark | 115 000 | 2001 |
| Slagelse, Denmark | 85 000 | 2002 |
| Vedbáek, Denmark | 15 000 | 2002 |

the aeration tanks as a wet weather treatment technique are listed in Table 3.14. Most of the recent literature on this subject has been published by a manufacturer who has patented a version of aeration tank settling called STAR® ATS (Bundgaard et al., 1996; Nielsen and Onnerth, 1995; Nielsen et al., 1996, 2000). This system combines aeration tank settling with an internal mixed liquor recycle stream and a high-level process control system. The recycle stream transfers mixed liquor from the last zone of the aeration tank (without air or mixing) to a preaeration anoxic zone, and extends the period of time for which aeration tank settling can be effective. Published data show that aeration tank settling results in increased denitrification and lower effluent orthophosphate accompanied by a slight increase in effluent turbidity (Bundgaard et al., 1996). No data were reported on changes in final effluent TSS or total phosphorus.

With the Kruger ATS concept, process air is turned off and RAS is reduced to approximately 20% of the influent flow. The combination of reduced mixed liquor concentration and reduced RAS increases the clarifier hydraulic capacity by 50% during storms.

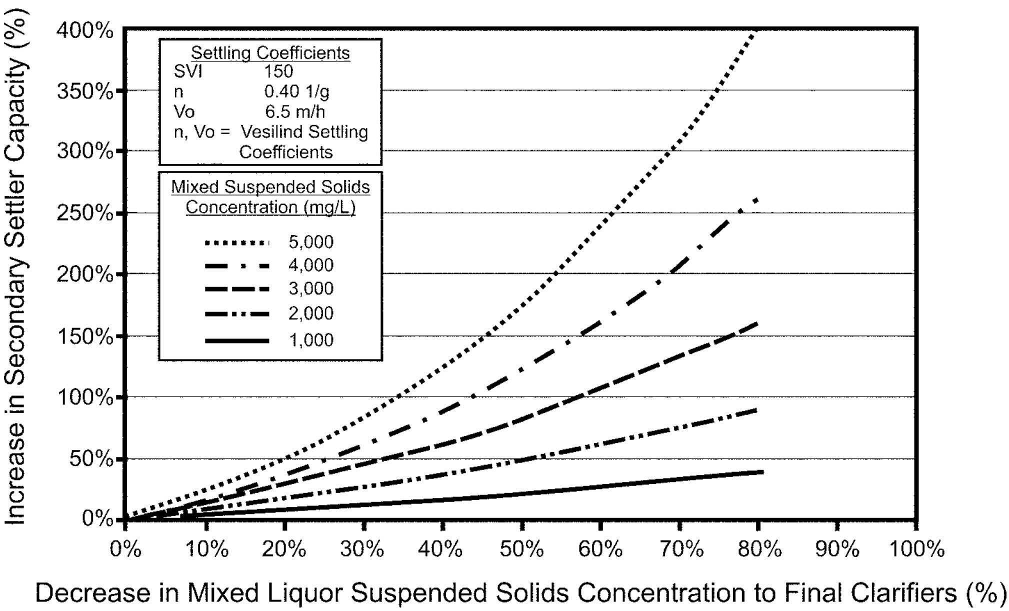


**FIGURE 3.18** Aeration tank settling potential to treat peak flows ($n$ = Vesilind coefficient calculated using the Daigger SVI correlation with an SVI of 150).

An evaluation of the effect of aeration tank settling, based on common U.S. practice using the Vesilind equation with the Daigger (1995) SVI correlation for the settling coefficients, is summarized in Figure 3.18. Figure 3.18 shows the estimated increase in clarifier capacity that results from a decrease in the mixed liquor concentration. Assuming that the secondary settling tanks are clarification limited, the effect of aeration tank settling is most pronounced at higher mixed liquor concentrations. For an SVI of 150 and a mixed liquor concentration of 3000 mg/L, a 50% drop in the mixed liquor concentration increases the clarifier capacity by more than 80%. A drop in the return sludge flow would not affect clarifier capacity under these conditions.

**STEP-FEED.** Switching to a step-feed or contact stabilization mode of operation during peak flows allows a greater mass of MLSS to be stored in the initial portions of the aeration tanks and minimizes the MLSS concentration fed to the secondary settling tanks. Using step-feed operation allows the plant to maintain a relatively high degree of treatment while treating a significantly higher flow rate. By varying the number and location of aeration tank feed points during wet weather flow events, the suspended solids concentration in the aeration tank effluent (secondary settling tank feed) can be reduced and the capacity of the secondary settling tanks can be increased significantly (Buhr et al., 1984; Monteith and Bell, 1998; Thompson et al., 1989). In conventional activated sludge processes, both the aeration tank influent and RAS are added to the beginning of the aeration tank, resulting in a relatively uniform concentration of suspended solids throughout the tank or tanks. A suspended solids gradient can be created in the aeration tank by feeding all or a portion of the influent stream at one or more locations along the length of the aeration tank while continuing to feed all of the RAS to the beginning of the aeration tank. Use of a step-feed pattern creates a high solids concentration at the beginning of the tank and a lower concentration at the end of the aeration tank. Thus, step-feed minimizes the solids loading applied to the final clarifiers for a given SRT and provides a greater mass of biomass and hence a larger SRT for a given tank volume than conventional activated sludge. The step-feed configuration becomes a contact stabilization process when all of the influent flow is added to a small zone at the end of the aeration tank. A balance must be established; however, between the increased clarifier capacity and reduced contact time between the aeration tank influent and the aeration tank biomass, as reduced contact time will, at some point, result in poorer treatment efficiency.

A mass balance on the MLSS solids and flow coupled with the assumption that the feed solids and biological growth are minor compared to the MLSS concentration will result in the following simplified design equations (Buhr et al., 1984). Equal volumes in each pass are also assumed.

$$\frac{X_n}{\overline{\overline{X}}} = \frac{(R+1)}{R+a_{cn}} \frac{X_N}{\overline{\overline{X}}} \tag{3.13}$$

$$\frac{X_N}{\overline{\overline{X}}} = \frac{N}{\frac{R+1}{R+a_{c1}} + \frac{R+1}{R+a_{c2}} + \ldots + \frac{R+1}{R+a_{c(N-1)}} + 1} \tag{3.14}$$

Where

$N$ = number of passes,
$n$ = individual pass number,
$X_N$ = last pass MLSS concentration,

$X_n$ = pass $n$ MLSS concentration,
$a_{cn}$ = cumulative fraction of flow to all passes up to and including pass $n$,
$\overline{X}$ = mean MLSS concentration, and
$R$ = return activated sludge ratio.

Research and full-scale implementation of step-feed for control of wet weather flows has demonstrated that secondary treatment standards can be met while switching between conventional and step-feed modes of operation (Georgousis et al., 1992; Thompson et al., 1989). For nitrifying activated sludge and biological nutrient removal (BNR) processes, maintaining complete nitrification and BNR while switching from conventional to step-feed can be more difficult. The ease and cost of modifying an existing conventional activated sludge process to be able to switch to a step-feed configuration during peak flows depends on the design of each facility. One study estimated the cost to convert several plants to allow step-feed operation at from approximately \$0.11 to \$5.28 $m^3$/d (\$400 to \$20 000/mgd) (Monteith and Bell, 1998). Care must be taken to provide adequate aeration capacity in zones not originally designed to receive influent flow.

**VORTEX SEPARATORS.** Vortex separators are rotary flow solids–liquid separation devices used to separate particulate matter from water. Vortex separators, also known as hydrodynamic vortex separators (HDVS) and swirl concentrators, are characterized by tangential inlets and surface overflows. Solids separated by gravity and inertial forces generally move towards the center of the unit by secondary currents and are removed from the base region of the device as a dilute sludge with a volume of approximately 5 to 10% of the influent flow. Solids removal can be accomplished continuously or on an intermittent basis. When used in CSO applications, solids removal is typically continuous. Because HDVS rely on secondary currents and other forces (e.g., centrifugal forces induced by a rotary flow pattern) to enhance gravity separation, they are unlike conventional clarifiers that rely only on the force of gravity. More than one thousand installations of vortex separation devices exist throughout Europe and North America, primarily in wet weather flow and CSO applications, with hundreds of installation for grit removal at wastewater treatment plants (Andoh et al.,2001, 2002). However, only a few installations are reported in use as high-rate clarifiers at wastewater treatment plants (Andoh and Saul, 2003; Andoh et al., 1996; Field and O'Connor, 1996).

While vortex separators first appeared in the literature in approximately 1949, the first substantial development and application occurred in the United Kingdom in the 1950s and 1960s (Smisson, 1967). In the 1970s, U.S. EPA sponsored additional research and development of vortex separators that resulted in the U.S. EPA Swirl Concentrator (Sullivan et al., 1972, 1982; Walker et al., 1993). Other designs have been developed from continuing research in the United Kingdom (Balmforth et al., 1994), Germany (Brombach et al., 1993), and Japan (Field et al., 1997).

Most vortex separators are relatively low-energy rotary flow devices in which complex secondary and recirculatory flows occur in addition to the main rotary flow pattern. Though the flow regimes in these devices have been described by idealized flow patterns such as rotational flow dynamics (forced vortex) or irrotational flow dynamics (corresponding to a free vortex flow regime), the actual flow patterns differ significantly from the ideal flow regimes with velocity distributions that vary both spatially and temporally. Depending on the configuration and flow regime, head losses in vortex separators can vary from smaller than 0.1 m (6 in.) in devices with a predominantly forced vortex type regime to approximately 0.9 to 3 m (3 to 10 ft) in devices with a predominantly free vortex type regime (Hides, 1999). Computational fluid dynamics (CFD) is increasingly being used for modeling the complex flow patterns in vortex separators and is an effective tool for gaining insights to flow regimes for different configurations of HDVS (Faram and Harwood, 2003). Though there are limitations in the applications of current CFD modeling tools such as difficulties in accurately simulating two-phase (water and solids) and three-phase flow (air, water, and solids), improvements and advancements are continuously being made to CFD codes and techniques.

Despite the similarities in operating principles, each type of HDVS is unique, with different geometries and internal components designed to stabilize the inherently unstable vortices developed by the rotary flow patterns. Three main designs are in common use and described in the literature—the U.S. EPA Swirl Concentrator (nonproprietary), the Storm King® (Hydro International US, Portland, Maine), and the FluidSep®—UFT Umwelt- und Fluid-Technik, Bad Mergentheim, Germany (John Meunier, Inc., Saint-Laurent, Quebec, Canada)—although other designs have been developed. Detailed descriptions of each design have been published (Andoh, 1998; Field and O'Connor, 1996; Field et al., 1997).

Hydrodynamic vortex separators have no moving parts and operate at higher hydraulic loading rates than conventional clarifiers. They are compact and can provide significant removal of settleable solids when properly sized and applied. While reported to be lower in cost than conventional clarifiers, the sludge, or underflow, from HVDS is typically more dilute than conventional primary sludge. As with other clarification systems, HDVS performance can be improved with the addition of chemicals and their performance decreases with increasing surface loading rates. Some configurations of HDVS used on CSOs, particularly those without continuous sludge removal, require cleaning after each use.

Because HDVS often operate on an intermittent basis in wet weather applications, evaluation of their treatment efficiency is more complicated than for conventional, continuous-flow clarifiers at wastewater treatment plants where concentration-based efficiency is typically calculated by assuming negligible underflow. Equations (see Table 3.15) have been developed to better differentiate the solids separation in HDVS obtained simply by splitting the flow versus concentrating the solids into the sludge stream (Field et al., 1997). Three performance measures have been

**Table 3.15** Vortex separator performance equations.

| Performance indicator | Equation* |
|---|---|
| Removal | $= \dfrac{C_i V_i - C_e V_e}{C_i V_i} \times 100 = \dfrac{M_i - M_e}{M_i} \times 100$ |
| Reduction | $= \dfrac{V_i - V_e}{V_i} \times 100$ |
| Net removal | = Removal − Reduction |
| Treatment factor | $= \dfrac{\text{Removal}}{\text{Reduction}} = \dfrac{(C_i V_i - C_e V_e)/C_i V_i}{(V_i - V_e)/V_i} = \dfrac{C_u}{C_i}$ |

*Pollutant mass (M), flow volume (V), and pollutant concentrations (C) are all storm-flow-event flowrate-weighted averages.

Where

$V_i$ = influent volume ($m^3$),
$C_i$ = influent concentration ($g/m^3$),
$V_e$ = overflow or effluent volume ($m^3$),
$C_e$ = overflow concentration ($g/m^3$), and
$C_u$ = underflow concentration ($g/m^3$).

defined for vortex separators—removal, net removal, and treatment factor (Field and O'Connor, 1996). Net removal quantifies the removal of solids beyond that obtained with a simple flow split (reduction), whereas the treatment factor is the ratio of the removal (by separation and concentration) to the reduction (by flow split). Similarly, a treatment factor greater than 1 indicates that solids are being removed from the flow. Short-duration events will result in high removals but low net removals and treatment factors.

Vortex separators are most effective at removing solids with relatively high settling velocities (>3.6 to 5.0 m/h [Pisano et al., 1990; Sullivan et al., 1982]). Successful use of these devices requires that the range of particle-settling velocities in the wastewater to be treated be adequately characterized. In general, settleable particles are considered to be inorganic solids (0.2 to 2 mm) with relatively high specific gravity (~2.65) and relatively large organic solids (0.2 to 5.0 mm), with specific gravities greater than 1.2 (Sullivan et al., 1982). Determination of the fraction of settleable solids can be made by allowing a sample to settle in a settling column or graduated cylinder (> 20 cm high) for 1 hour. Assuming that the sample is siphoned from middepth after 1 hour, the equivalent settling velocity is approximately 0.11 m/h (Field et al., 1997). As with conventional clarifiers, vortex separators are not effective at removing solids with near neutral buoyancy. Recent developments in vortex separation technology include variants that incorporate novel self-cleansing screening systems to capture the neutrally buoyant solids fraction in CSOs (Andoh and Saul, 2000).

Particle-settling velocity, hydraulic surface loading rate, and the ratio of the underflow to the inlet flow are the primary factors affecting the particle-separation efficiency of vortex separators. Dimensionless analysis and model studies show that efficiency decreases rapidly when the ratio of the surface loading rate to the particle-settling velocity increases from 0.1 to 2.0 (Weiß and Michelbach, 1996). Figure 3.19 shows the suspended-solids separation efficiency as a function of hydraulic loading rate relative to the settling velocity ($q_A/v_s$) and the ratio of the separator underflow to the influent flow ($Q_{out}/Q_{in}$) for one vortex separator design. Vortex separator performance improves with decreasing overflow rate and increasing underflow rate. Design hydraulic loads vary depending on the application and performance objectives. For example, the suggested hydraulic loading range for primary treatment equivalency is approximately 5 to 10 m/h, which contrasts with a range of 70 to 140 m/h for grit removal and a range of 10 to 100 m/h for CSO applications (Andoh et al., 2002; Field et al., 1997). Care must be taken in calculating hydraulic loading rates, as different researchers and suppliers may define the surface area differently.

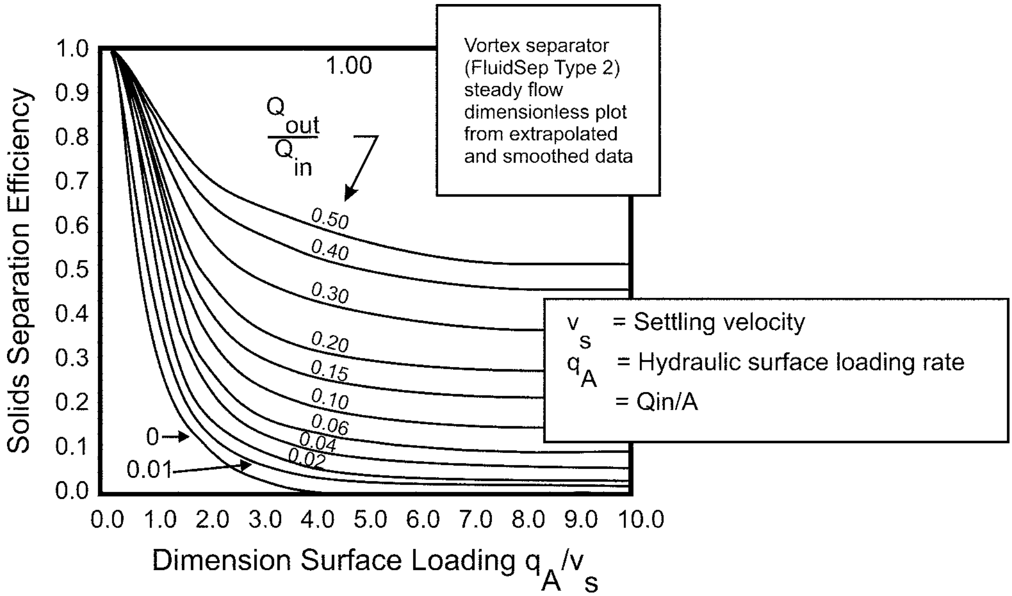


**FIGURE 3.19** Dimensionless steady-flow efficiency curves and dependence on the parameters $q_A$ (hydraulic surface loaing rate, m/h; $A$ = surface area of HVS, $m^2$)/$v_s$ (settling velocity, m/h) and $Q_{out}$ (effluent flow, $m^3$/h)/$Q_{in}$ (influent flow, $m^3$/h) (reprinted from *Water Sci. Technol,* **33,** 277–284, with permission from the copyright holder, IWA).

Comparatively little performance data are available for vortex separators, particularly for use at wastewater treatment plants. Selected performance data on full-scale vortex separators in CSO applications are summarized in Table 3.16. Some data on the performance of vortex separators treating municipal wastewater are available (Andoh et al., 1996; Dudley, 1994; Dudley et al., 1994). Trials at the Chester-le-Street and the Totnes Wastewater Treatment Works (WWTW), United Kingdom, demonstrated that vortex separators can provide primary treatment according to the European Commission Urban Wastewater Directive (20% $BOD_5$ removal and 50% TSS removal). Observed $BOD_5$ removals ranged from approximately 35 to 85%, whereas TSS removals ranged from approximately 55 to 90%. During the Totnes trials, $BOD_5$ and TSS removals averaged 23 and 47%, respectively, whereas with chemical addition the $BOD_5$ and TSS removals averaged 73 and 92%. On this basis, a vortex separator with optimized chemical addition is expected to exceed 70% for TSS and 90% for $BOD_5$ (Andoh et al., 1996).

**TABLE 3.16** Vortex separator performance.

| Location | Device type | Design flow (mgd) | Mass suspended solids removal (%) | Mass suspended solids net removal (%) | Source |
|---|---|---|---|---|---|
| Washington, D.C. | EPA Swirl | 507 000 | 37.6 (21.1–67.5) | 12.4 (0.9–73) | Field and O'Connor, 1996 |
| Tengen, Germany | FluidSep® | 40 900 | 54.0 (32–91) | 6.9 (-7.1–15.8) | Field and O'Connor, 1996 |
| Walsall, England | Storm King® | 32 600 | 53.2 (13.9–95.0) | 14.3 (-0.2–33.9) | Field and O'Connor, 1996 |
| West Roxbury, Massachusetts | EPA Swirl | 14 800 | 28.1 (9.5–36.0) | 17.0 (2.8–32.0) | Field and O'Connor, 1996 |
| Columbus, Georgia | Storm King® | | 40.0 | | Boner, 2003; Turner, 2003 |
| Toronto, Canada | Storm King® | | 20–80 | | Averill et al., 1999; Schmidt et al., 1997 |

# CASE STUDIES

**BALLASTED FLOCCULATION.** Bremerton, Washington, is located on Puget Sound in Kitsap County approximately 24 km (15 mi) west of Seattle, Washington. Port Washington Narrows, a narrow estuary connecting two major embayments of the Puget Sound (Dyes and Sinclair Inlets), splits the city into Bremerton and East Bremerton. One wastewater system with both combined and sanitary sewers, including 15 CSOs, and one treatment plant serve approximately 37 000 residents in the city. An average flow of approximately 19 000 m$^3$/d (5 mgd) is received at the wastewater treatment plant but peak wet weather flows can exceed 151 400 m$^3$/d (40 mgd).

In 1993, the city was issued a consent order by the State of Washington Department of Ecology and also settled a lawsuit with the Puget Soundkeeper Alliance. Among other things, the consent order and agreement required the city to reduce the discharge of untreated CSOs from its Pine Road basin to the Port Washington

Narrows section of the Puget Sound to less than one per year. Water quality standards for treatment of CSOs in Washington are set by the State Department of Ecology at 50% removal of suspended solids and an effluent settleable solids concentration of less than 0.3 mg/L.

Combined sewer overflow planning began in 1989 after the State of Washington Department of Ecology issued regulations limiting CSO discharges. A CSO reduction plan was prepared in 1992 and updated in 2000. The CSO plan evaluated alternative methods of reducing CSOs throughout the city. Alternatives evaluated included the construction of relief sewers to the wastewater treatment plant, and treatment processes such as fine screening, primary sedimentation, filtration, vortex separation, constructed wetlands, and dissolved air flotation. After conducting engineering evaluations and pilot testing of two ballasted flocculation systems, the city amended its CSO reduction plan to implement ballasted flocculation at the Pine Road CSO (Eastside Treatment Plant). Results from the pilot plant testing are summarized in Table 3.17.

The Eastside Treatment Plant was built on the site of a primary treatment plant that was demolished in the mid-1980s. The old marine outfall was still useable and was converted to the outfall for the CSO treatment facility. The onshore CSO has been eliminated. The CSO treatment facility that was constructed includes 38 $m^3$ of short-term storage, influent screening, ballasted flocculation, and UV disinfection. The design peak flow for this facility is 76 300 $m^3/d$, with an overflow volume of 37 850 $m^3$ over 48 hours. The facility was designed with a Lamella overflow rate of 98 m/h at a flow rate of 38 200 $m^3/d$. It is expected to provide 90% removal of suspended solids up to a flow rate of approximately 54 500 $m^3/d$. During rare peak storms, the overflow rate is expected to exceed 180 m/h and during these events performance is expected to drop below 90% TSS removal. Effluent quality is not expected to degrade during peak storms because of reduced influent concentrations. The design effluent quality is summarized in Table 3.18.

Design criteria for the facility are summarized in Table 3.19. The facility footprint is approximately 13.7 m by 9.75 m, the project cost was approximately $4.1 million, and operation began in December 2001. The UV disinfection was designed to be expanded in the existing channel by 25% if necessary. Land is available on the site to construct at least one parallel HRC and UV train.

Bremerton faced a number of challenges during its first year of operation, including multiple equipment problems and the typical learning curve associated with new processes. Despite the startup problems, the facility always met its permit limits, although suspended solids and $BOD_5$ removals were often much less than was expected from pilot testing. Effluent fecal coliforms have averaged approximately 30 per 100 mL with a maximum value of 78. Performance in 2004 has been much improved, with reported solids removals ranging from 50 to 90%.

Initial operation of the Bremerton and other CSO facilities has highlighted design features that require adequate attention to minimize operating problems. Combined sewer overflow treatment facilities must inherently treat a wide range of flows, and facility components like flow meters, chemical feed systems, and sludge pumps should be adequately sized to handle the full range of flows expected. All water-quality characteristics that affect chemical dose requirements must be taken into consideration. As noted above, water quality characteristics for wet weather flows are often significantly different from dry weather flows and may not respond to treatment in the same way as diluted dry weather wastewater used for pilot testing. For example, the alkalinity in wet weather flows can be significantly less than dry weather in systems with moderate to hard water supplies and high collection system inflow. Hydraulic drops tend to result in foaming and should be avoided. Although high-rate clarification is effective at reducing suspended solids, significant turbidity may remain in the discharge.

**COMBINATION STORAGE/SETTLING TANKS.** The Sugar Creek Wastewater Treatment Plant is one of five major wastewater treatment plants owned and operated by Charlotte–Mecklenburg Utilities in Charlotte, North Carolina. This 75 700-$m^3$/d treatment facility is located adjacent to Little Sugar Creek and serves a sewershed to the south of Charlotte. In 1992, the sewer collection and conveyance system served by the Sugar Creek Wastewater Treatment Plant was experiencing overflows as a result of excessive rainfall-derived infiltration and inflow. Onsite effluent polishing lagoons at Sugar Creek were converted to a combination flow equalization/settling tank facility as part of a systemwide program to eliminate system overflows. The overall program included sewer system rehabilitation and increases in trunk sewer capacity.

Actual peak flow rates experienced by the Sugar Creek plant were unknown because the inlet Parshall flumes became submerged at 227 000 $m^3$/d. Flows in excess of the peak hydraulic capacity of Sugar Creek (approximately 151 000 $m^3$/d) were typically bypassed to an interceptor that fed the downstream McAlpine Creek Wastewater Management Facility. The new flow-equalization facility was sized for the difference between the interceptor and treatment plant capacity (208 000 $m^3$/d). Though

**TABLE 3.17.** Bremerton, Washington, ballasted flocculation pilot-plant data.

| Date[a] | Time | Flow ($m^3$/d) | Rise rate (m/h) | $FeCl_3$[b] (mg/L) | Polymer (mg/L) | Turbidity Raw | Turbidity Settl. | TSS (mg/L) Raw | TSS (mg/L) Settl. | TSS (mg/L) Removal (%) | BOD (mg/L) Raw | BOD (mg/L) Settl. | BOD (mg/L) Removal (%) | Comments |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| 8-Dec | 18:45 | 1740 | 134 | 45 | 0.50 | 49 | 2.9 | 86 | 8 | 91 | 78 | 26 | 67 | Varying |
| | 19:15 | | | | 0.75 | 41 | 2.0 | 76 | 2 | 97 | 78 | 38 | 51 | polymer dose |
| | 19:55 | | | | 1.00 | 45 | 2.5 | 68 | 8 | 88 | 78 | 39 | 50 | |
| 9-Dec | 14:20 | 1740 | 134 | 15 | 1.0 | 39 | 2.4 | 38 | 1 | 97 | 94 | 56 | 40 | Varying |
| | 15:15 | | | 25 | | 33 | 1.9 | 292 | 10 | 97 | 75 | 60 (33) | 20 | $FeCl_3$ dose |
| | 18:40 | | | 35 | | 29 | 3.3 | 64 | 6 | 91 | | | | |
| | 19:15 | | | 45 | | 26 | 2.7 | 132 | 22 | 83 | 85 | 83 | 2 | |
| 10-Dec | 12:00 | 1740 | 134 | 25 | 1.0 | 23 | 4.4 | 108 | 64 | 41 | 216 | 50 | 77 | 10-hour demon- |
| | 14:00 | | | | | 19 | 3.1 | 58 | 8 | 86 | 188 | 67 | 64 | stration run |
| | 16:00 | | | | | 26 | 3.5 | 342 | 26 | 92 | 342 | 64 | 81 | |
| | 18:00 | | | | | 21 | 2.9 | 128 | 1 | 99 | 236 | 76 | 68 | |
| 13-Dec | 11:30 | 1740 | 134 | 25 | 1.0 | 24 | 2.9 | 58 | 10 | 83 | 154 | 76 | 51 | 10-hour demon- |
| | 13:30 | | | | | 21 | 3.2 | 78 | 32 | 59 | 164 | 86 | 48 | stration run |
| | 15:00 | | | | | 26 | 4.1 | 736 | 20 | 97 | 318 | 86 | 73 | "Cut run 2 hours short" |

| | | | | | | | | | | | | | |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| 15-Dec | 19:30 | 1740 | 134 | 25 | 1.00 | 31 | 30.0 | 32 | 30 | 6 | 115 | 63 | 45 |
| | 19:32 | | | | | 28 | 37.0 | 18 | 28 | 0 | 115 | 28 | 76 |
| | 19:34 | | | | | 34 | 5.9 | 34 | 10 | 71 | 99 | 62 | 37 |
| | 19:36 | | | | | 22 | 6.7 | 30 | 12 | 60 | 96 | 56 | 42 |
| | 19:38 | | | | | 25 | 2.2 | 24 | 2 | 92 | 114 | 45 | 61 |
| | 19:40 | | | | | 26 | 4.3 | 28 | 2 | 93 | 135 | 38 | 72 |
| | 19:45 | | | | | 25 | 2.3 | 28 | 6 | 79 | 119 | 41 | 66 |
| | 20:00 | | | | | | 4.2 | 32 | 8 | 75 | 111 | 39 | 65 |
| | 22:00 | | | | | 29 | 1.9 | 28 | 10 | 64 | 131 | 41 | 69 |
| | 22:30 | | | | | 19 | 3.8 | 28 | 12 | 57 | 124 | 43 | 65 |
| 16-Dec | 0:00 | | | | | 14 | 2.7 | 12 | 2 | 83 | 121 | 43 | 64 |
| | 2:00 | | | | | 17 | 1.5 | 10 | 2 | 80 | 126 | 35 | 72 |
| | 4:00 | | | | | | 0.9 | 26 | 8 | 69 | 138 | 26 | 81 |
| | 6:00 | | | | | 23 | 1.9 | 20 | 2 | 90 | 126 | 28 | 78 |
| | 8:00 | | | | | 18 | 2.6 | 36 | 8 | 78 | 143 | 56 | 61 |
| | 10:00 | | | | | 16 | 3.4 | 90 | 10 | 89 | 183 | 47 | 74 |
| | 12:00 | | | | | 21 | 4.8 | 76 | 34 | 55 | 149 | 81 | 46 |

[a]For the demonstration runs on December 10 and 13, 2-hour composites were made by combining four 30-minute grab samples. The turbidity and pH values are averages of the 30-minute samples for the composites.

[b]$FeCl_3$ = ferric chloride.

**TABLE 3.18** Bremerton, Washington, CSO reduction plant removal during design storm.

| **Design storm condition** | **Value** |
|---|---|
| Influent TSS at normal flow (mg/L) | 150 |
| Normal removal efficiency (%) | >90 |
| Influent TSS during 4-hour peak condition (mg/L) | <50 |
| Removal efficiency during 4-hour peak condition (%) | 50 |
| Overall removal efficiency (%) | >82 |

**TABLE 3.19** Bremerton, Washington, Pine Road CSO design criteria.

| **Item** | **Value** |
|---|---|
| Influent | |
| Average flow during design storm ($m^3/d$) | 37 854 |
| Peak flow during design storm ($m^3/d$) | 75 700 |
| Short-term storage | |
| Total volume ($m^3$) | 378 |
| Volume available before plant startup ($m^3$) | 284 |
| Time before full during design event (hours) | 1 |
| Inlet screening | |
| Number | 1 |
| Capacity ($m^3/d$) | 75 700 |
| Screen opening (mm) | 6 |
| High-rate clarification | |
| Process trains (number) | 1 |
| Injection/coagulation zone | |
| Detention time (minutes) | |
| At 37 854 $m^3/d$ | 1.4 |

**TABLE 3.19** Bremerton, Washington, Pine Road CSO design criteria. *(continued)*

| | |
|---|---|
| High-rate clarification *(continued)* | |
| Detention time (minutes) *(continued)* | |
| At 75 700 $m^3/d$ | 0.7 |
| Length × width (m) | 2.9 × 2.9 |
| Maturation zone | |
| Detention time (minutes) | |
| At 37 854 $m^3/d$ | 4.1 |
| At 75 700 $m^3/d$ | 2.0 |
| Length × width (m) | 4.6 × 5.5 |
| Clarification | |
| Rise rate (gpm/sq ft[a]) | |
| At 37 854 $m3^3/d$ | 98 |
| At 75 700 $m^3/d$ | 183 |
| Length × width (m) | 5.5 × 5.5 |
| UV disinfection | |
| Maximum head loss at 20 mgd[b] (mm) | 254 |
| Minimum unfiltered UV transmittance (%) | 50 |
| Minimum delivered dosage (mW-sec/$cm^2$) | 30 |
| Number of channels | 1 |
| Length × width × depth (m) | 11.3 × 2.1 × 3.7 |
| Capacity at target of 200 MPN/100mL fecal (mgd) | 20 |
| Number of UV banks per channel | 2 |
| Number of modules per bank | 3 |
| Number of lamps per module | 16 |
| Total number of lamps | 96 |
| Total power required (kW) | 300 |
| Expandability (%) | 25 |

[a]gpd/sq ft × 0.002 = m/h.

[b]mgd × 3785 = $m^3/d$.

the primary purpose of the storage facility was wet weather flow management, it was also desired that the storage basin be able to provide dry weather flow equalization.

Storage volumes for both equalization of dry weather diurnal flows and wet weather peak flows were estimated. Mass-balance calculations using historical flow data from dry weather periods and diurnal flow patterns were used to estimate the dry weather storage volume. Diurnal flow equalization requirements were determined to be between 3400 and 7200 $m^3$, or approximately 6 to 12% of the average daily flow. An excess flow volume or flow exceedence represents conditions when the storage volume is full and the flow rate is higher than the plant capacity. Under these conditions, flow must be bypassed to the downstream plant. The number of annual excess flow volumes, or flow exceedences, for different combinations of expected peak flows, treatment capacity, and storage volume were estimated using the U.S. Army Corp of Engineers STORM program. Figures 3.20 and 3.21 present the results of the STORM modeling. These figures show that the use of the entire existing polishing lagoon for wet weather storage volume would reduce the

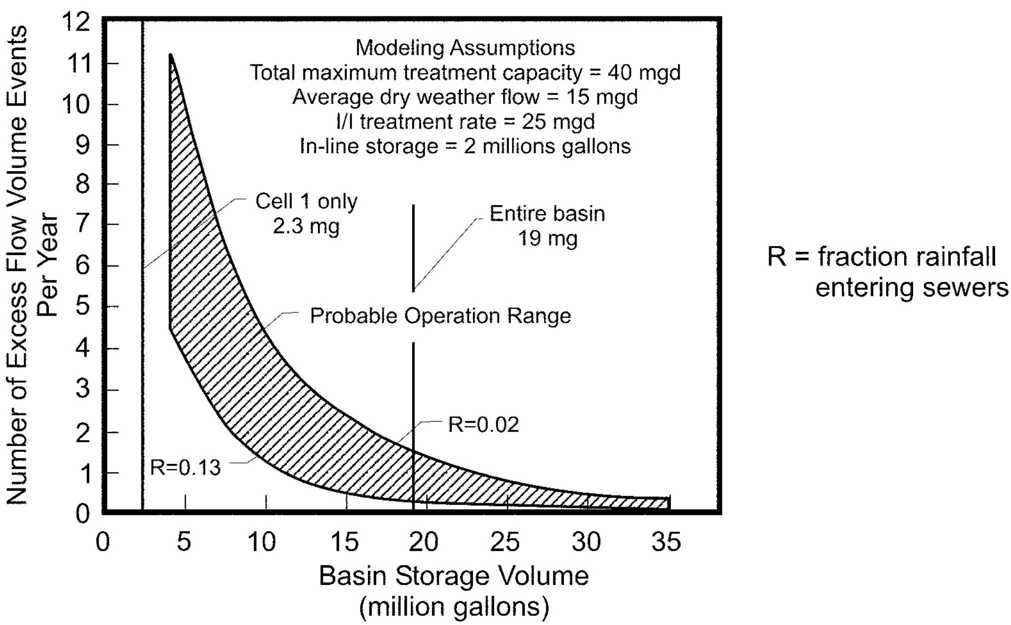


**FIGURE 3.20** Charlotte-Mecklenburg Utilities Sugar Creek storm curves (mil. gal × 3785 = $m^3$; mgd x 3785 = $m^3$/d).

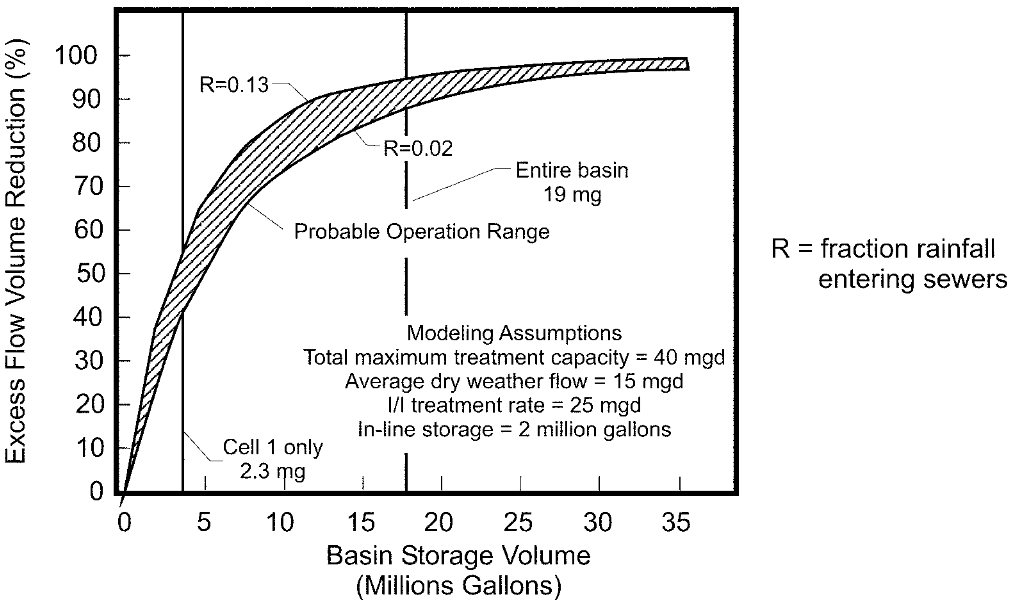


**FIGURE 3.21** Charlotte-Mecklenburg Utilities Sugar Creek storm curves (mil. gal × 3785 = $m^3$; mgd x 3785 = $m^3/d$).

number of excess flow events to between 1 to 2 events per year at a treatment plant capacity of 151 000 $m^3/d$.

Consideration was given to subdividing the existing lagoon into one, two, or three cells. This would allow the use of the available storage volume to be more closely matched to that needed for a given storm event to minimize the volume to be cleaned after typical storm events. Subdividing the lagoon would also allow the first cell to function as a primary clarifier for the flow diverted to the lagoon. The first cell was sized at 8700 $m^3$ to provide a 1-hour detention time for the design peak flow of 208 000 $m^3/d$. Further consideration was given to subdividing the remaining volume into two cells; however, this was not implemented because the second cell would only be used approximately 5 to 12 times per year and the third cell would only be used approximately one to four times per year. It was decided that the lagoon would be divided into an 8700-$m^3$ first cell and a 62 000-$m^3$ second cell. Disinfection using chlorine was provided where wet weather flows would overflow the first cell into the second cell.

The existing lagoon was constructed from membrane-lined earthen berms. Concrete lining was added to the interior side slopes and the bottom to provide a stable bottom that could be cleaned regularly. To facilitate regular cleaning water, cannons with a capacity of 36 L/s were installed at 52-m intervals around the perimeter of both cells. Vehicle access ramps were provided to allow the removal of solids using mechanical equipment.

**PILOT TESTING OF HIGH-RATE CLARIFICATION.** The Village Creek Wastewater Treatment Plant serves more than 750 000 people in Tarrant County and portions of Johnson County, Texas. Treated water is discharged to the Trinity River. The plant is permitted to treat an annual average daily flow of 545 000 $m^3/d$; however, it is estimated that, during wet weather periods, the plant can experience peak flows of up to 1 722 000 $m^3/d$. Plant peak flow capacity was estimated to be 965 000 $m^3/d$. The treatment facilities consist of screening, primary clarifiers, secondary treatment with activated sludge, filtration, and disinfection. Sludge is thickened, anaerobically digested and dewatered with belt filter presses (City of Fort Worth, 2002).

The City of Fort Worth, Texas, worked with U.S. EPA Region VI to explore the potential to apply ballasted flocculation as an alternative to constructing more primary and secondary treatment facilities or flow equalization storage to provide treatment for peak wet weather flows. In conjunction with these negotiations, the city undertook an intensive 2-week pilot test of four different high-rate clarification processes (CDM, 1999). The testing was conducted at the Village Creek plant from September 14 to October 9, 1998. The pilot testing was undertaken to demonstrate the feasibility of the technology; establish optimum coagulant, polymer, and ballast dosages at increasing overflow velocities; and evaluate the quantity, quality, and effect of the enhanced high-rate process sludge on primary clarifier performance and sludge thickening.

The four systems tested were

- USFilter (Cary, North Carolina) Microsep® process,
- Parkson (Ft. Lauderdale, Florida) Lamella® Plate Clarification process,
- USFilter/Kruger (Cary, North Carolina) Actiflo® process, and
- Infilco-Degrémont (Richmond, Virginia) DensaDeg 4D® process.

The USFilter Microsep® and Actiflo® processes both used ballasted flocculation. The primary difference was the inclusion of plate settling in the Actiflo® process. The Parkson Lamella® uses enhanced chemical coagulation followed by

**TABLE 3.20** Fort Worth, Texas, pilot test flowrates (CDM, 1999).

| Manufacturer | Units | Low | Medium | High |
|---|---|---|---|---|
| Parkson Lamella® clarifier | m/h | 36.7 | 48.9 | 73.3 |
| USFilter Microsep® | m/h | 48.9 | 73.3 | 97.8 |
| Infilco-Degrémont DensaDeg 4D® | m/h | 97.8 | 122 | 147 |
| USFilter/Kruger Actiflo® | m/h | 122 | 147 | 171 |

Lamella® clarification (CEPT with plates). The DensaDeg 4D® process is similar in concept to Actiflo®, except that it uses chemically conditioned recycled sludge to improve flocculation and ballast the floc.

The overflow velocities tested are listed in Table 3.20.

The pilot program was designed to develop dosage curves for coagulant and ballast versus increasing overflow velocity. Polymer doses ranged from 0.75 to 1.25 mg/L, and ballast concentrations were 5 and 10 g/L. Ferric sulfate was the only coagulant used in the testing. During the testing, three influent and effluent samples were collected each hour at each pilot unit. The pilot work found that TSS removal increased with increasing coagulant dose and with increasing polymer dose; however, the relationship was not as clear for the polymer as it was with the coagulant. The optimum doses found are summarized in Table 3.21.

The optimum ballast concentrations were reported to be as follows (CDM, 1999):

- USFilter/Kruger Actiflo® Process

  —At 122 m/h, 6 to 8 g/L

  —At 171 m/h, 8 to 10 g/L

- USFilter Microsep®

  —At 49 m/h, 5 to 7 g/L

  —At 98 m/h, 8 to 10 g/L

The results from the demonstration phase of the pilot testing on a blend of raw wastewater with secondary effluent to simulate wet weather wastewater and on raw wastewater are summarized in Tables 3.22 and 3.23, respectively. In general, the performance of all four units was similar, with the exception of the maximum hydraulic

**Table 3.21** Optimum coagulant and polymer doses (CDM, 1999).

| | Blended (mg/L) | Raw (mg/L) |
|---|---|---|
| USFilter/Kruger Actiflo® Process | | |
| Coagulant dose | 70–80 | 125 |
| Polymer dose | 0.75–1.0 | 1.3 |
| Parkson Lamella® Clarifier | | |
| Coagulant dose | 50–60 | 100 |
| Polymer dose | 0.75–1.0 | 1.0 |
| Infilco-Degrémont DensaDeg 4D® | | |
| Coagulant dose | 70–80 | 150 |
| Polymer dose | 0.75–1.0 | 1.75 |
| USFilter Microsep® | | |
| Coagulant dose | 80–85 | 150 |
| Polymer dose | 1.0–1.25 | 1.25 |

overflow rates that could be treated. The Actiflo® process (with chemical addition, sand ballast, and plates) and the DensaDeg® process were able to provide good treatment at much higher overflow velocities. The primary difference between Actiflo® and DensaDeg® was the additional time (120 minutes versus 20 minutes) required by the DensaDeg® process to achieve full performance.

The final recommended design overflow rates from the Fort Worth study for the four processes are provided in Table 3.24. Recommended chemical and microsand doses were

- Ferric sulfate, 70.0 to 125 mg/L;
- Anionic polymer,0.75 to 1.0 mg/L; and
- Microcarrier, 7.0 to 10.0 g/L.

Based on the success of the pilot testing program, the city of Fort Worth was able to negotiate renewal of the NPDES permit to allow diversion of primary effluent to a high-rate clarification process when flows exceed 965 000 $m^3/d$. Flows treated by high-rate clarification will be returned to the main flow upstream from the chlorine contact basins.

Construction of the new ballasted flocculation system is expected to be complete in 2005. The ballasted flocculation process will have one train designed to treat a

**Table 3.22** High-rate clarification performance during demonstration testing on blended wastewater.

| Process | TSS (mg/L) | $BOD_5$ (mg/L) | Total Kjeldahl nitrogen (mg/L) | Total phosphorus (mg/L) |
|---|---|---|---|---|
| USFilter/Kruger Actiflo® process | | | | |
| 122 m/h | 89 | 36 | 25 | 92 |
| 147 m/h | 92 | 64 | 30 | 95 |
| 171 m/h | 74 | 62 | 28 | 96 |
| Parkson Lamella® clarifier* | | | | |
| 37 m/h | 73 (82) | (57) | 34 | 76 |
| 49 m/h | 76 (81) | (41) | 19 | 69 |
| 73 m/h | 53 (62) | (41) | 30 | 71 |
| Infilco-Degrémont DensaDeg 4D® | | | | |
| 98 m/h | 90 | 63 | 28 | 95 |
| 122 m/h | 81 | 49 | 28 | 95 |
| 147 m/h | 85 | 37 | 40 | 88 |
| USFilter Microsep® | | | | |
| 49 m/h | 80 | 54 | 25 | 95 |
| 73 m/h | 43 | 38 | 29 | 88 |
| 98 m/h | 31 | 31 | 19 | 41 |

*After 20 minutes and (120 minutes) operating time.

**Table 3.23** High-rate clarification performance on raw wastewater.

| | TSS (mg/L) | $BOD_5$ (mg/L) | Total Kjeldahl nitrogen (mg/L) | Total phosphorus (mg/L) |
|---|---|---|---|---|
| USFilter/Kruger Actiflo® process | | | | |
| 98 m/h | 88 | 63 | 30 | 95 |
| Parkson Lamella® clarifier | | | | |
| 49 m/h | 82 | 58 | 25 | 82 |
| Infilco-Degrémont DensaDeg 4D® | | | | |
| 98 m/h | 92 | 51 | 20 | 94 |
| USFilter Microsep® | | | | |
| 73 m/h | 83 | 50 | 2 | 83 |

maximum flow of 416 000 $m^3/d$. The flow scheme for the ballasted flocculation process comprises the following:

- Influent channel,
- Two sludge hoppers for future grit removal,
- Five influent submersible pumps,

**Table 3.24** Recommended design overflow rates from Fort Worth, Texas, demonstration testing.

| | Overflow rates (m/h) |
|---|---|
| Infilco-Degrémont DensaDeg 4D® | 100 |
| USFilter/Kruger Actiflo® | 100 |
| Parkson Lamella® clarifier | 50 |
| USFilter Microsep® | 73 |

- Flow measurement weir,
- One coagulation/flow-splitting tank,
- Two injection tanks,
- Two maturation tanks,
- Lamella settling tanks,
- Sludge-handling facilities,
- Hydrocyclones, and
- Chemical storage and handling facilities.

The footprint for the facility measures approximately 18 m by 81 m (60 ft by 265 ft).

**AERATION TANK SETTLING.** Aeration tank settling (ATS) has been implemented as wet weather flow control in several plants in Scandinavia. In 2001, ATS was implemented at the Hirtshals wastewater treatment plant in Denmark. The implementation of ATS was part of a major project to implement STARcontrol® (USFilter/Krüger, now Krüger, Inc.—A Veolia Water Systems Company (Cary, North Carolina) at the plant. The STARcontrol® system is an advanced software program that optimizes the control of chemical and biological wastewater treatment facilities. In addition to ATS, the control system implemented at Hirtshals allows for the control of aerobic and anoxic phase lengths, dissolved oxygen setpoints, chemical doses, and return sludge flow rate. All of these initiatives were undertaken in response to an increasing load on the plant from the fishing industry in the town. The industrial contribution to the plant is approximately 70 to 80%.

The town of Hirtshals, Denmark, is served by a wastewater treatment plant designed to treat the flow from a population equivalent of 53 000 (One population equivalent is defined by the Urban Waste Water Treatment Directive of the European Environment Council (Brussels, Belgium) to be an organic biodegradable load of 60 g $BOD_5$/d. This corresponds to 12 g nitrogen/d, and 2.5 g phosphorus/d. The standard U.S. definition for a hydraulic population equivalent is 100 gpd = 0.3785 $m^3$/d population equivalents.) Biological nutrient removal is provided using the BioDenitro™ process, which is a phased isolation ditch process based on alternating aeration and mixing in the biological reactors. The BioDenitro™ process at Hirtshals consists of two aeration tanks followed by a traditional secondary sedimentation tank.

During normal operation (dry weather), the mode of operation consists of alternating aeration and mixing with two main phases and two intermediate phases. In the first main phase, influent flow is directed to one of the tanks, which is kept anoxic by stirring but not aerating. The effluent from the first tank flows to the second tank, which is kept aerobic by aeration. In the second main phase, the states of the reactors and the flow direction are reversed. As more time is generally needed for nitrification than for denitrification, intermediate phases with aerobic conditions in both tanks are applied between the main phases.

During rainstorms, ATS is used instead of the normal operation cycle. In the main phases of ATS, the flow direction is the opposite of normal operation and the anoxic tank is not stirred. Therefore both denitrification and sedimentation occur in the anoxic phase. See the process scheme in Figure 3.22. In the first main phase in Figure 3.22, the suspended solids settle in the left tank, whereas nitrification takes place in the right tank. Effluent is taken from the settling tank to retain suspended solids. At some time, it is better to change states in the two tanks so that the settling tank becomes the nitrification tank and vice versa. The reason is that more sludge is available for nitrification in the settling tank than in the nitrification tank. Therefore, the system is changed to the opposite main phase. Before changing to the opposite main phase, an intermediate phase is applied. In the intermediate phase, suspended solids settle in both tanks, that is, the suspended solids in the right tank

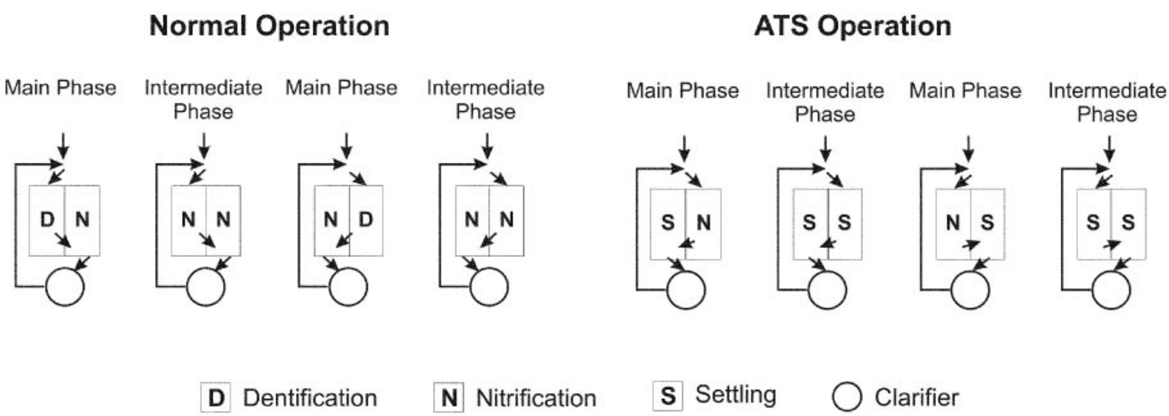


**FIGURE 3.22** Normal and ATS phase isolation ditch operation schemes.

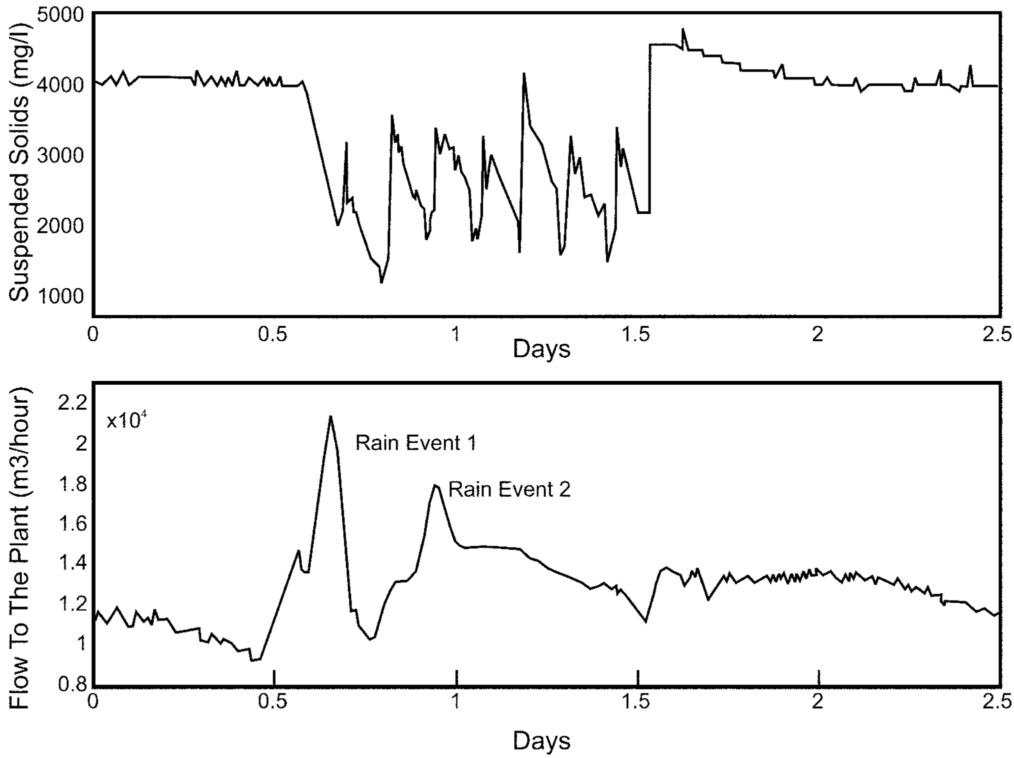


**FIGURE 3.23** Example of ATS in operation (suspended solids measured in the aeration tank effluent).

are "presettled" before flow is discharged from this tank. If this intermediate phase is not applied, the effluent is discharged from a tank that has just been mixed, hence would be rather high. Generally, the intermediate phase is much shorter than the main phases.

An online suspended-solids sensor is used to measure the suspended solids in the inlet to the settler (i.e., the outlet from the aeration tanks). An example of these online measurements can be seen in Figure 3.23. Before ATS operation mode is started, the effluent suspended-solids concentration is fairly constant at 4000 mg/L. Once ATS is started, the concentration of suspended solids leaving the aeration tanks drops rapidly because of sedimentation in the aeration tanks. During the

ATS event, an average aeration tank effluent suspended-solids concentration of 2500 mg/L is achieved. When the ATS operation mode is ended, both reactors are mixed again and the MLSS concentration increases to 4500 mg/L. This measurement is representative of the average suspended-solids concentration in the aeration tanks. This means that the aeration tanks held 12.5% more sludge during ATS control than during normal operation, which indicates a considerable reduction of the sludge load to the settler. The suspended-solids concentration in the aeration tanks is back to normal after 12 hours.

The control of ATS includes a special routine for the control of aeration and mixing phase lengths, dissolved oxygen setpoints, and sludge recirculation to ensure maximum exploitation of the system. The control goal is to increase the hydraulic capacity of the whole system without loosing too much nitrification and denitrification capacity at the same time. The timing for the start of an ATS control period is a crucial part of the control scheme. The automatic initiation of ATS control can be decided solely based on the influent flow rate so that, once a certain criterion is exceeded, the ATS control scheme is applied. However, by using flow prediction based on rain gauges located upstream from the plant, it is possible to prepare the plant for the increased hydraulic load 30 to 45 minutes in advance.

Today, all of the control loops work well, and process performance at the plant continues to improve. In addition to the improvements as a result of ATS control, the implementation of the total STARcontrol® system has reduced the nitrogen and phosphorus effluent concentrations by 45 and 65%, respectively. Chemical consumption for nutrient removal has been reduced by 60% and energy consumption (electricity) has been reduced by 10%. These improvements have been achieved in spite of an increase in load to the plant during the period.

Though significant water quality data for the final effluent were not available for the Hirtshals wastewater treatment plant with and without ATS, such data have been published from testing conducted on a similar plant—the 330 000 population equivalent Aalborg West wastewater treatment plant. Table 3.25 contains a summary of effluent data from the Aalborg during ATS operation.

During the wet weather event testing at the Aalborg West plant in 1994, the increase in plant flow was limited to approximately a 50% increase in flow to secondary treatment by the pump capacity. Although the increased hydraulic loading was less than most plants experience during wet weather peaks, there was no deterioration in nutrient removal.

**TABLE 3.25** Effluent water quality during ATS operation at the Aalborg West Wastewater Treatment Plant (Denmark) (Bundgaard et al., 1996).

| Parameter | Units | 8-17-1994 | 8-18-1994 | 9-15-1994 | 9-16-1994 |
|---|---|---|---|---|---|
| Rain | mm/d | 31.8 | 33.4 | 63.8 | 14.2 |
| Rainfall duration | h | 6 | 15 | 21 | 18 |
| Secondary effluent flow | $m^3/d$ | 80 790 | 147 500 | 174 530 | 175 540 |
| Secondary bypass | $m^3/d$ | 23 900 | 78 392 | 115 792 | 62 520 |
| Secondary bypass without ATS | $m^3/d$ | 31 000 | 113 668 | 178 320 | 121 914 |
| Reduction of bypass | % | 30 | 45 | 54 | 95 |
| Ammonia nitrogen | mg/L | 0.73 | 0.10 | 0.35 | 0.10 |
| Nitrate nitrogen | mg/L | 1.8 | 1.0 | 1.3 | 1.8 |
| Total phosphorus | mg/L | 0.29 | 0.11 | 1.4 | 0.28 |
| Orthophosphorus | mg/L | 0.20 | 0.19 | 0.59 | 0.14 |
| Suspended solids | mg/L | 6 | 7 | 15 | 11 |

**VORTEX SEPARATORS.** Two existing full-scale vortex separator installations illustrate the range of applications for this technology for treatment of wet weather flows. Use of vortex separators as high-rate clarification devices at satellite CSO treatment sites within collection systems was demonstrated at Columbus, Georgia, where six vortex separators preceded by screens and followed by compressed media filters were installed on a combined sewer system and their performance was monitored for six years (Boner, 2003).

At the South West Water Totnes WWTW in the United Kingdom, two chemically assisted vortex separators have been used to provide primary treatment before a high-purity oxygen activated sludge process. A third unit treats wet weather flows in excess of three times dry weather flow. In 1992, the performance of the Totnes vortex separators was evaluated in detail by the U.K. Water Research Centre to establish their suitability as a process to meet the requirements of the European Commission Urban Wastewater Directive for marine discharge (Dudley and Marks, 1993).

Faced with the regulatory mandates of the Clean Water Act and the Safe Drinking Water Act and a then forthcoming U.S. EPA CSO policy, the Columbus (Georgia) Water Works initiated a three-phased program to address wet weather induced water-quality problems in the middle reach of the Chattahoochee River. Included in the program was the construction of two satellite CSO treatment facilities, which use vortex separators for solids control and chemical disinfection. One of the facilities also served as a national full-scale demonstration program to test vortex separators followed by a compressed media filter and several alternative disinfectants as CSO treatment technologies. The demonstration project treatment facilities are shown schematically in Figure 3.24 and include coarse screens followed by six 9.75-m-diam vortex separators, a compressed media filter with a 760-mm bed of 25.4-mm fabric balls, and UV disinfection. Each vortex unit has a volume of approximately 380 m$^3$. An urban catchment area of approximately 390 ha contributes combined sewer flows to the demonstration plant. The sequence of operation for the CSO facility depends on the incoming flow rate. Wastewater flows up to 37 850 m$^3$/d

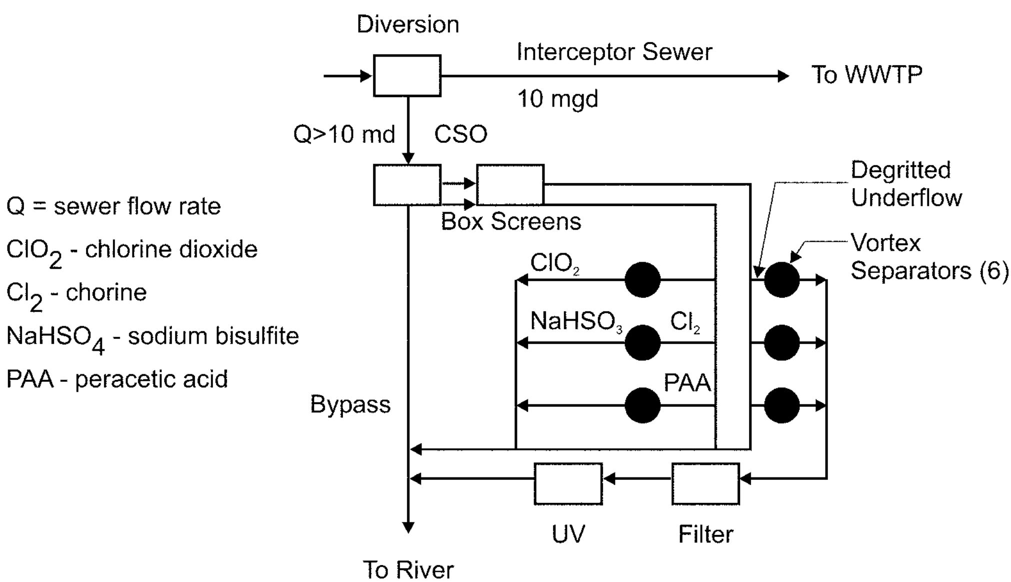


**FIGURE 3.24** Process flow schematic for Columbus, Georgia, vortex demonstration project (WWTP = wastewater treatment plant; mgd $\times$ 3785 = m$^3$/d).

(10 mgd) continue in the interceptor to the wastewater treatment plant. Once the flow exceeds the capacity of the interceptor sewer, flow is directed to the six vortex units, and disinfectant addition is initiated. Each vortex vessel provides approximately 3 minutes detention time for chemical disinfection. After the vortex tanks are full, flow is then directed to the compressed media filters and UV disinfection. Flow greater than the CSO facility capacity is bypassed to the river. All six vortex separators are operated in parallel until the tanks are full, at which time the sixth unit is used to concentrate the underflow from the remaining five units online after their underflows have been degritted using a 2.4-m-diam vortex grit-removal unit housed in an adjacent building.

This arrangement resulted in the organic solids and their related pollutants being concentrated in a significantly smaller portion of the flow (approximately 1% of the peak design flow for the facility) being transported through the collection system to the main wastewater treatment plant. There was, therefore, no need to upsize the collector/interceptor sewer and the removal of grit and sediments upstream provided major operational benefits as the potential for sediment accumulation in the relatively flat interceptor was averted.

Performance monitoring over a period of six years showed that vortex separators accomplish several treatment operations, including (1) the reduction of a significant number of CSO discharges with approximately 40% of the annual volume captured by interception and storage and 82% of the annual volume treated, (2) high level removals of oil and grease (90%), (3) grit and gross solids removals of more than 90%, (4) primary removals for the lighter fraction TSS contaminants on an annual basis (40%), (5) metal removals of 50%, and (6) phosphorus removal of 60%. Vortex vessels were also found to be effective contact chambers for chemical disinfection, resulting in water-quality objectives being met.

Vortex separators were equivalent to conventional primary treatment at loading rates up to approximately 12 m/h (40% removal of suspended solids). At loading rates of more than 12 m/h, vortex separators still effectively removed the readily settleable solids such as fecal solids, grit, sediments, wastewater debris, and other heavy particles but the light particulate TSS removal as measured by conventional small bore tube samplers became negligible.

Conventional small bore tube samplers do not typically measure gross solids and suspended solids larger than 63 μm and, as such, the coarse fraction of organic solids, sediments, and their associated pollutants that are known to be present in combined sewer flows but are typically not accounted for in water-quality studies and evaluations that use conventional small bore tube sampling equipment. For example, at Columbus, Georgia, the observation has been that, for every kilogram of light particulate TSS removal across the vortex vessels measured by the conventional TSS analytical procedure, there are more than 2 kg of unaccounted-for solids material removed by the vortex units. This quantification is based on quantities of sediment captured in dumpsters from the degritting of underflow lines from the primary vortex units.

Figure 3.25 shows the measured suspended-solids removal of the Columbus vortex separators. Figure 3.26 shows estimates of expected performance from an optimized CSO treatment facility at Columbus with two 9.75-m-diam vortex separators and a 57-$m^3$ filter. A flow of 189 250 $m^3$/d (50 mgd) results in an overflow velocity of approximately 26 m/h. During high flow events, the Columbus vortex facility is relying heavily on filtration for solids removal.

As a remote facility, the Columbus CSO plant is not staffed full time but remote monitoring is provided. Operators visit the facility to check chemical feed systems and water-quality samplers, provide routine maintenance, and check for residuals removal after storm events. During storm events, operators visit the facility to check

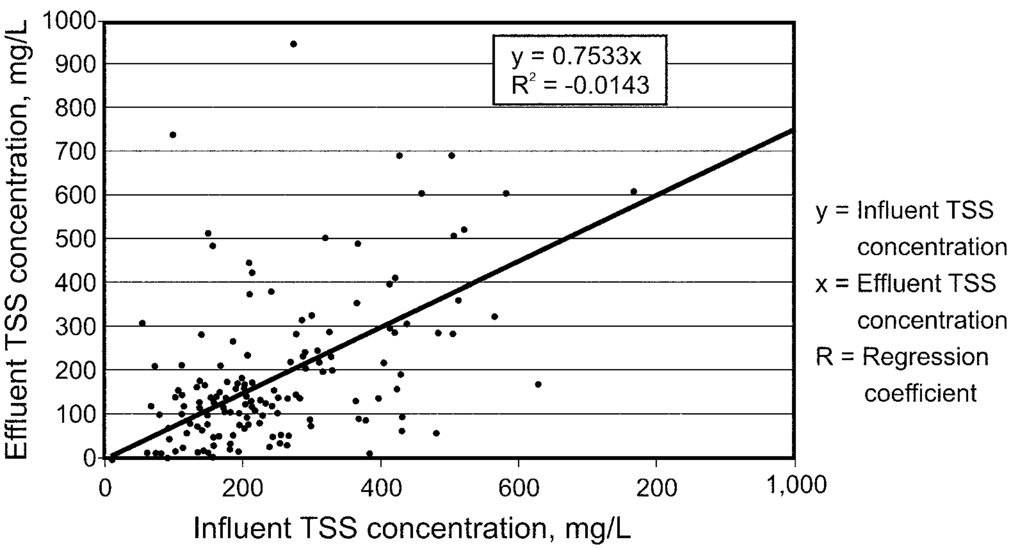


**FIGURE 3.25** Typical TSS removal in the vortex separators at the Columbus, Georgia, demonstration project.

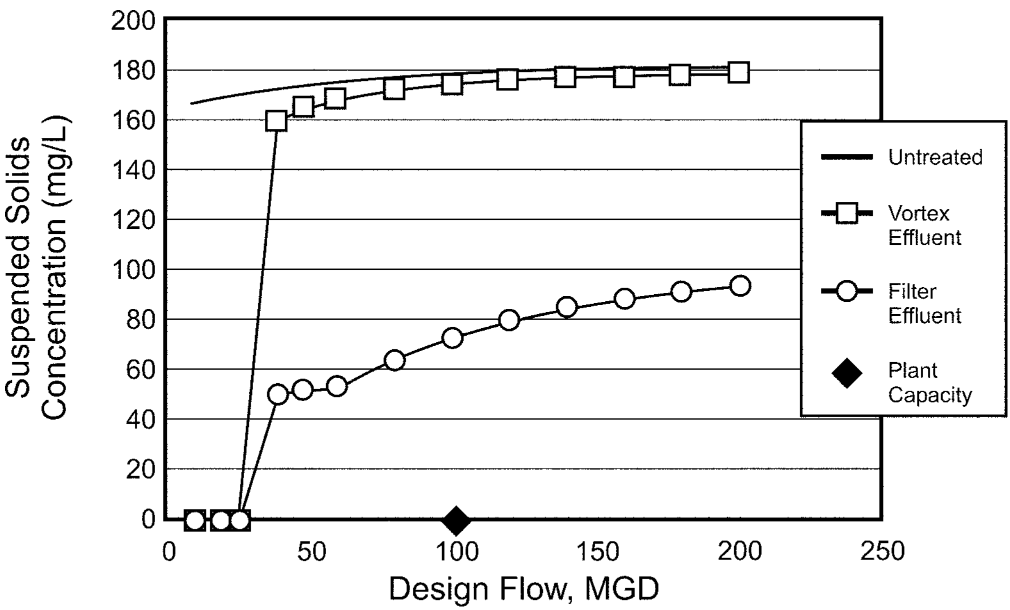


**FIGURE 3.26** Example suspended solids removal by vortex separators at Columbus, Georgia, demonstration project (mgd × 3785 = $m^3/d$).

equipment operation, take chlorine residual measurements, log operating conditions, and transport water-quality samples to the laboratory. Sodium hypochlorite, chlorine dioxide, peracetic acid, and UV radiation were tested as disinfectants. Sodium bisulfite was used to remove residual chlorine from the addition of chlorine. The vortex units were found to be cost-effective vessels for accomplishing high-rate chemical disinfection in addition to serving as primary clarifiers on an annual mass basis.

A significant advantage found for the vortex separators at Columbus was their effectiveness as preliminary and primary treatment units and the significant reductions in capital and operating costs associated with CSO control that they provided, especially given that the primary operation and maintenance issue observed for CSO treatment was in the handling and removal of grit and gravel. The vortex units have no moving parts and are self-cleansing on draindown.

The Columbus water-quality program led to the conclusion that cost-effective CSO controls can be achieved by using direct treatment processes such as the vortex separator with chemical disinfection followed by compressed media filtration and

UV disinfection. The CSO controls developed and tested in Columbus satisfy the U.S. EPA CSO policy and meet water-quality standards in the Chattahoochee River as demonstrated by postconstruction monitoring and calibrated watershed modeling.

Testing of the vortex units at Totnes WWTW was conducted to demonstrate the ability of vortex separators with chemical addition to meet the European Commission Urban Wastewater and Bathing Waters Directives (Dudley and Marks, 1993). Minimum $BOD_5$ and TSS removals of 20 and 50%, respectively, are required by the European Commission for primary treatment. To discharge treated wastewater to marine waters, the European Commission requires removals of 70% for $BOD_5$ and 90% for TSS, with 99% removal of indicator bacteria.

The vortex units at Totnes comprised two parallel trains, with each treatment train consisting of a 4.24-m vortex unit used for grit removal and chemical coagulation (labeled Swirl-Flo Separator, Hydro International US, Portland, Maine) followed by an 8.54-m-diam settling unit (labeled Swirl-Flo Clarifier, Hydro International US, Portland, Maine). The water depth in the clarifier vortex units was 8.138 m. A third, smaller, 2.52-m vortex unit (sludge decant tank) concentrates sludge from both of the large vortex separators. Decant water from the sludge concentrator was returned to the coagulant tanks. Figure 3.27 illustrates the vortex separator configuration at the Totnes WWTW.

With a design capacity of 28 000 population equivalents, the design flow of the Totnes vortex separators was approximately 2300 $m^3/d$ at the time of the performance testing. Testing of the vortex units was conducted at 2300 and 4600 $m^3/d$. At design flow, the coagulation (separator) and clarification units had working volumes of 40 and 480 $m^3$, respectively. At 200% of design flow, the working volumes were 47 and 360 $m^3$, respectively. During performance testing, one vortex train was operated with chemical addition and one without. To avoid returning chemicals from the decant unit to the train without chemical addition, the overflow from the unit without chemical addition was stored in a tank rather than being recycled.

Performance testing was conducted at design flow and two times design flow. Also included was tracer testing, operation with low chemical dose, operation with high chemical dose, analyses for bacteria and virus, characterization of sludge, and evaluation of capital and operating costs. A proprietary coagulant that has iron sulfate as its active ingredient was used as the coagulant at doses of approximately 200 mg/L at design flow and approximately 350 to 380 mg/L at twice design flow. This was added to the small vortex units (separators). Ferric chloride was added to the large (clarifiers) vortex units as a coagulant. Flow to each vortex separator train was

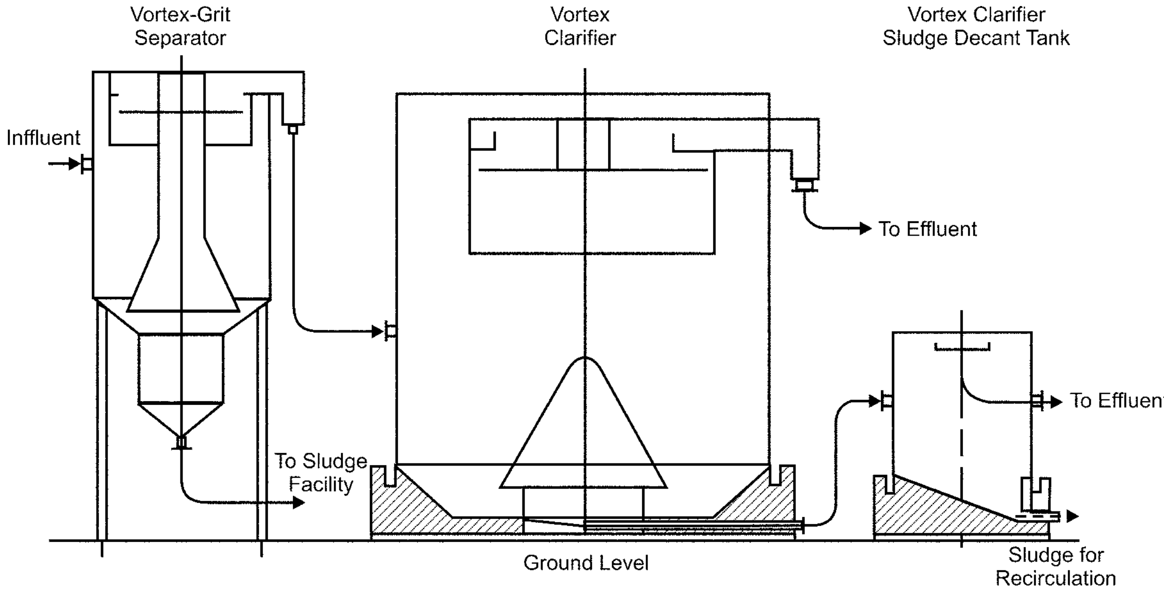


**FIGURE 3.27** Vortex separator process configuration at Totnes WWTW.

2300 $m^3/d$ to simulate design conditions and 5200 $m^3/d$ for 200% of design. Overflow rates for the clarifier units at design flow were approximately 1.7 m/h and approximately 3.8 m/h at twice design flow.

Tracer testing showed that the small vortex units had little dead space and behaved increasingly like plug-flow reactors as the flow rate increased. Conversely, the large units acted slightly less like plug-flow reactors at the higher flow rate. Using a tanks-in-series model, the large vortex units acted like 3.2 and 2.6 tanks-in-series at design and two times design flow while the smaller unit acted like 3 and 4 tanks-in-series, respectively, at design flow and twice design flow.

Results from the low dose test at the design flow are reproduced in Table 3.26. Grab samples were taken every hour for four days for the major water quality parameters. Bacterial sampling was conducted hourly and daily spot samples were taken for nonroutine microbiological analyses. Average removals for $BOD_5$ were 37% with chemical addition and 23% without chemical addition. Suspended-solids removals were 55 and 47% with and without chemicals, respectively. Phosphorus removal was 37% with chemical addition. At the low dose, however, the bacteria content was too high to meet bathing water standards.

**Table 3.26** Average water quality during low dose trials at design flow at the Totnes WWTW.

| Parameter | Raw wastewater* | Effluent quality*: Coagulator with chemical | Clarifier with chemical | Clarifier without chemical |
|---|---|---|---|---|
| pH | 8 | 6.8 | 6.7 | 7.3 |
| Turbidity | 132 | 104 | 63 | 109 |
| Soluble BOD | 156 | 105 | 126 | 126 |
| Total BOD | 319 | 223 | 200 | 244 |
| Total Kjeldahl nitrogen | 38 | 35 | 35 | 36 |
| Suspended solids | 283 | 257 | 128 | 151 |
| Total phosphorus | 7.5 | 6.1 | 4.7 | 8.7 |
| Oils, fats, and grease | 40 | 19 | 7.4 | 15.6 |

*All units in milligrams per liter except pH in standard units and turbidity in nephelometric turbidity units.

Results from the high dose test at design flow are reproduced in Table 3.27. Grab samples were taken every hour for four days for the major water-quality parameter and hourly samples were taken for bacteria. Average removals for $BOD_5$ were 73% with chemical addition and 19% without chemical addition. Suspended-solids removals were 92 and 47% with and without chemicals, respectively. Phosphorus removal was 97% with chemical addition and 10% without.

Results from the high flow trials are summarized in Table 3.28. At higher flow rates, vortex overflow velocities increase and hydraulic detention times decrease. As a result, performance is expected to decrease; however, for the low dose trials at 200% design flow, removals were unexpectedly better. This was attributed to the slighter weaker raw wastewater during the high flow, low dose trials. At slightly less than 70% removal efficiency for both $BOD_5$ and suspended solids, performance was lower at the higher chemical dose but still higher than that required by European Commission directives. More than 99% of bacteria were removed at the high chemical dose, resulting in effluent concentrations of approximately 200 000 for total coliform, 70 000

**Table 3.27** Average water quality during high dose trials at design flow at the Totnes WWTW.

| | | Effluent quality* | | |
|---|---|---|---|---|
| **Parameter** | **Raw wastewater*** | **Coagulator with chemical** | **Clarifier with chemical** | **Clarifier without chemical** |
| pH | 8.4 | 5.5 | 5.5 | 7.7 |
| Turbidity | 99 | 90 | 15 | 86 |
| Soluble BOD | 131 | 77 | 77 | 96 |
| Total BOD | 306 | 151 | 83 | 249 |
| Total Kjeldahl nitrogen | 32 | 28 | 23 | 31 |
| Suspended solids | 255 | 204 | 19 | 135 |
| Total phosphorus | 7.6 | 3.3 | 0.2 | 6.8 |
| Oils, fats, and grease | 33 | 13 | 1 | — |

*All units in milligrams per liter except pH in standard units and turbidity in nephelometric turbidity units.

for fecal coliform, and 15 000 for fecal streptococci. Bacteria removals were sufficient to meet the European Commission Bathing Water Directive limits of 10 000 total coliform and 2000 fecal coliform.

Sludge from the low dose trials could be readily digested anaerobically and thickened to the same extent as conventional primary sludge. However, the ability to dewater the sludge was significantly reduced. In contrast, sludge from the high dose trials inhibited anaerobic digestion and resulted in lower gas production. The high dose sludge did not thicken or dewater well. An important observation from the time series of water-quality parameters measured showed that the vortex separators tended to produce a consistent effluent quality and showed an ability to absorb shock loadings.

Overall, the Totnes study concluded that vortex separators are an appropriate treatment process to meet European Commission directives and recommended operation without chemicals when the discharge is away from bathing waters and with chemicals when the discharge is near bathing waters.

**Table 3.28** Water quality results from trials at 200% design flow at the Totnes WWTW.

| | Low dose | | High dose | |
|---|---|---|---|---|
| **Parameter** | **Average effluent*** | **Average removal (%)** | **Average effluent*** | **Average removal (%)** |
| Soluble BOD | 40 | 44 | 34 | 47 |
| Total BOD | 78 | 57 | 67 | 65 |
| Total Kjeldahl nitrogen | 17 | — | 17.4 | 6 |
| TSS | 80 | 42 | 52 | 69 |
| Turbidity | 34 | 64 | 23 | 79 |
| Phosphorus | 4.06 | 1.58 | 0.17 | 96 |
| Oils | 5.3 | 77 | 3 | 88 |
| Transmittance | 59 | — | 36.8 | — |

*All units in milligrams per liter except turbidity in nephelometric turbidity units and transmittance in percent of light transmitted.

## PROCESS SELECTION

Table 3.29 provides a summary of expected performance and hydraulic loading rates for various wet weather clarification alternatives, providing an overview of relative site area requirements and the potential for water-quality improvement.

Selection of the best technological solution to wet weather flow problems is a subjective and sometimes controversial process that depends on water-quality objectives and environmental regulations, characteristics of the individual collection and treatment facilities, local economic conditions, policy set by the system owners, and preferences of the community and operations staff. Clarification is a strong candidate to be part of any wet weather treatment alternative because of its relatively low capital and operating costs. Characterization of the range of expected influent wastewater flows and quality during wet weather periods is essential to establishing the relative performance and cost of the wet weather clarification alternatives discussed here. Another recommended, and often mandatory, first step is to determine the hydraulic and treatment capacity of the existing treatment plant. To

**TABLE 3.29** Alternate wet weather clarifier designs.

| Primary settling process | Removal efficiency (%) | | Peak overflow rate (m/h) | Reference |
|---|---|---|---|---|
| | $BOD_5$ | TSS | | |
| Conventional | 25–30 | 40–50 | 3.4–5.0 | (WEF, 1998b) |
| CEPT | 45–65 | 60–85 | 3.0–5.0 | (CDM, 1995; Morrisey and Harleman, 1992; Odegaard, 1998) |
| High polymer CEPT | — | 40–80 | 30–40 | (Averill et al., 1999) |
| Plates | 30–60 | 60–90 | 10–15 * | (Murcott and Harleman, 1992) |
| CEPT w/ plates | 40–60 | 60–90 | 30–40 * | (Rogalla et al., 1992) |
| Dense sludge | 40–60 | 70–90 | 25–100 * | (Metcalf and Eddy, 2003; Murcott and Harleman, 1992) |
| Ballasted floc | 40–60 | 70–90 | 100–130 * | (EPRI, 1999; Young and Edwards, 2000) |
| Vortex separators (w/o chemicals) | 25–30 | 40–50 | 4–10 | (Andoh et al., 1996; Dudley and Marks, 1993; Dudley et al., 1994) |
| Vortex separator (w/ chemicals) | 37–86 | 55–94 | 4–40 | (Andoh and Harper, 1994; Andoh et al., 1996; Averill et al., 1996; Dudley and Marks, 1993; Dudley et al., 1994) |

* Lamella overflow rate.

not do so, or to do this in a cursory manner based on standard criteria, can be costly. Dynamic process simulation is invaluable in evaluating the response of biological treatment processes, including the level of the sludge blanket in secondary clarifiers to wet weather flows and loads. Such evaluations of existing facilities will often spotlight bottlenecks that can be removed at a sometimes modest cost. The ultimate goal of any wet weather treatment program should be the protection of receiving waters from adverse water-quality effects that would result from inadequate treatment of wet weather flows. From a rational standpoint, any combination of treatment plant and operational modifications that enable a plant to meet discharge water-quality standards should be acceptable. Then the goal becomes determination of the most economical approach. While reliable cost estimates must come from site-specific studies, in many cases the lowest cost approaches are those that maximize the capacity of existing facilities by removing bottlenecks, rerating unit processes, implementing alternative flow configurations, and providing for biological bypass and blending. Approaches requiring construction of new facilities must be evaluated within the context of the individual situation. Construction of new, conventional wet weather primary or secondary clarifiers should have the highest capital cost but operation and maintenance requirements are well established, the volume of the tanks provides storage if empty at the start of storm, and additional annual costs are low. Conversion of conventional primaries to CEPT during wet weather minimizes capital costs to the extent that the size of the primary clarifiers can be minimized but incurs additional annual costs in the form of chemicals and additional sludge production. Operating cost effects such as those associated with chemical use, increased sludge production, or reduced aeration costs will be proportional to the expected duration of wet weather and, in many cases, will be relatively low. High-rate clarification processes offer dramatically reduced footprints and often increased pollutant-removal efficiencies but incur varying degrees of additional annual costs. Advantages of reduced land-area requirements; however, can be substantial in highly developed urban areas with limited land for facility expansions, high land costs, and the need to minimize the effects on aesthetics for plant neighbors.

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# Chapter 4

# Secondary Clarifier Design Concepts and Considerations


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# INTRODUCTION

The purpose of this chapter is to present the underlying secondary clarifier design and discuss general design and selection considerations. Most often, final clarifiers are discussed in conjunction with suspended-growth systems, primarily because of sludge settleability issues and the dependency of the biological process on the return sludge. However, many of the elements covered in this section can also be applied to clarifiers following attached-growth systems. Though clarifiers have served suspended- and attached-growth processes for decades, opinions differ as to what constitutes an optimal design. Several references (Ekama et al., 1997; Tekippe, 1986; and Tekippe and Bender, 1987) are available to help readers develop an appreciation of the theories, variety of design criteria, and various geometric details used in recent years. In addition, the behavior of various clarifier configurations may be predicted fairly accurately using calibrated computer models.

Secondary clarifiers do not function in isolation and should not be designed without considering upstream and downstream processes for the following reasons:

- Clarification efficiency is directly related to sludge quality (i.e., how a sludge settles, compacts, and flocculates), which is caused by conditions (overaeration, lack of aeration, low food-to-microorganism ratio [F/M], etc.) in the bioreactor.

- Poor sludge settleability will result in lower return activated sludge (RAS) solids and higher RAS flowrates for the same bioreactor mixed liquor suspended solids (MLSS).
- Excessive turbulence in the MLSS conveyance system created by pumps or significant drops in the hydraulic profile could break up the floc, resulting in the need to reflocculate.
- Inefficient influent screening may clog vacuum sludge collection system and certain sludge pumps.
- Provided flocculent sludge develops in the bioreactor, proper clarifier design will ensure lower effluent solids and smaller effluent filters, if filters are required.

## FUNCTIONS OF A FINAL CLARIFIER

The primary function of a final clarifier is clarification, which is a solids-separation process that results in the removal of biological floc from the liquid stream. During the subsequent thickening process, sludge particles are conveyed to the bottom of the tank, resulting in a concentrated underflow (RAS). In underloaded and critically loaded clarifiers, the RAS solids concentration is a function of the recycle ratio. A secondary function is to store sludge during peak flow periods. If the clarifier fails in either of these functions, the performance of the biological process may be affected. As well, because of solids carryover, the effluent may not meet specified discharge limits.

It should be noted that thickening in clarifiers is a root cause of several performance-related problems. In addition, clarifier underflow concentrations of more than 1.0 to 1.5% solids are difficult to achieve. For these reasons, consideration should be given to operating the clarifier with a shallow sludge blanket (minimum thickening) and using sludge-thickening devices (e.g., gravity belt thickener, centrifuge, or dissolved air flotation thickener) for thickening, which can achieve significantly higher solids concentrations.

The key factors that affect clarifier performance are listed in Table 4.1 (adopted from Ekama et al., 1997).

Whereas all of these are important considerations, flow and sludge characteristics are central to sizing the clarifier. The remaining factors enhance clarifier performance and improve process reliability (Ekama et al., 1997).

**TABLE 4.1** Factors that affect clarifier performance (adapted from Ekama et al., 1997).

| Category | Factors |
|---|---|
| Hydraulic and load factors | Wastewater flow (ADWF, PDWF, PWWF)* |
| | Surface overflow rate |
| | Solids loading rate |
| | Hydraulic retention time |
| | Underflow recycle ratio |
| External physical features | Tank configuration |
| | Surface area |
| | Depth |
| | Flow distribution |
| | Turbulence in conveyance structures |
| Internal physical features | Presence of flocculation zone |
| | Sludge-collection mechanism |
| | Inlet arrangement |
| | Weir type, length, and position |
| | Baffling |
| | Hydraulic flow patterns and turbulence |
| | Density and convection currents |
| Site conditions | Wind and wave action |
| | Water temperature variation |
| Sludge characteristics | MLSS concentration |
| | Sludge age |
| | Flocculation, settling, and thickening characteristics |
| | Type of biological process |

*ADWF = average dry weather flow; PDWF = peak dry weather flow; and PWWF = peak wet weather flow.

## CLARIFIER CONFIGURATIONS

Clarifier shape determines whether the actual flow pattern approaches radial or plug-flow. Radial flow occurs in circular, rectangular, hexagonal, and octagonal tanks. Plug-flow clarifiers are rectangular in shape. Circular and rectangular clarifiers are the most popular. A well-designed rectangular clarifier can be expected to perform similarly as a well-designed circular unit.

The shape of new clarifiers provided may be dictated by consistency and operator familiarity. Clarifiers are often designed to closely match existing units. For some plants, saving surface area may be of paramount importance to allow room for other process units. In such cases, rectangular tanks with common wall construction may be the choice. Table 4.2 compares circular and rectangular clarifiers.

**TABLE 4.2** Comparison of rectangular and circular clarifiers.

| | Rectangular clarifiers | Circular clarifiers |
|---|---|---|
| Advantages | Less land and construction cost in a multiple unit design<br>Longer flow path and less chance for short-circuiting than center-feed/peripheral overflow circular clarifiers<br>More even distribution of sludge loads on collectors<br>Can be shallower<br>Low head loss for flow distribution<br>Can be easily covered for odor control<br>More effective foam/scum trapping and positive removal<br>Not proprietary | Short detention time for settled sludge<br>Better effect of dynamic filtration<br>Simple and more reliable sludge-collecting system<br>Low maintenance requirements |
| Disadvantages | Longer detention time for settled sludge (except for Gould-type designs,[a] which have very short detention times)<br>Possibly less effective for high solids loading[b]<br>Increased maintenance of collectors | Center feed/peripheral units have higher potential for short-circuiting<br>Lower limits for effluent weir loading<br>Generally proprietary<br>More susceptible to wind effects<br>High headloss for flow distribution |

[a]See Chapter 9.

[b]Lack of data at high loadings; most rectangular clarifiers are operated at lower solids loadings.

Additional discussion on the geometric features of clarifiers may be found in Chapters 8 and 9.

## BASICS—THE SCIENCE OF DESIGN

**SEDIMENTATION PROCESS.** Settling basins handling wastewater must separate a variety of materials in the clarification zone. As shown in Figure 4.1, depending on the concentration of the suspended solids and the tendency of the particles to flocculate, four distinct types of settling processes are typically recognized in wastewater treatment plant design:

- Type I, or discrete nonflocculent settling: particles in the top left corner (Figure 4.1) are completely dispersed with no tendency to flocculate. These particles settle independently at their terminal velocity.

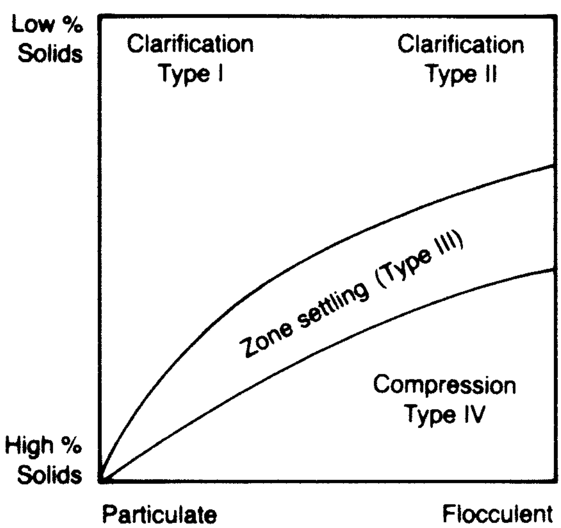



FIGURE 4.1 Relationship between solids characteristics and sedimentation processes.

- Type II, or flocculent settling: particles at the top right of Figure 4.1 are dispersed but have a strong affinity to flocculate. With time, the particles coalesce and settle as flocculated particles.
- Type III, or zone settling: in this settling regime, particles that that have a strong tendency to coalesce do so quickly and settle together as a matrix. All of the particles within the matrix settle at the same velocity. As they settle, the particles retain their relative position to each other.
- Type IV, or compression settling: as the solids settle to the bottom of the tank, the particles come into mechanical contact. The resulting compressive forces squeeze out the water and the sludge is thickened.

While all four types of settling may occur in secondary clarifiers, type III governs design. Type I occurs to a limited extent at the top part of the clarifier where the flocculated particles undergo discrete settling because of very low particle concentration (Ekama et al., 1997). Below this layer, true flocculent settling (type II) is encountered. Types I and II contribute to clarification, the actual separation of the solids from the liquid stream. Type III occurs in the middle to lower middle part of the clarifier and is responsible for the conveyance of the solids to the bottom. Type IV is encountered at the bottom of the clarifier, where thickening of the settled sludge occurs. As shown in Table 4.3, the four classes of settling involve different particle behavior and, therefore, different capacity-controlling factors.

***Type I Settling (Discrete Settling).*** Type I settling is the predominant mechanism in gravity grit chambers. It occurs to a limited extent in secondary clarifiers. Each particle is assumed to settle independently and with a constant (or terminal critical) velocity. The mathematical treatment of type I settling is presented in Chapter 3.

**Table 4.3** Type of settling and controlling factor.

| Types of settling | Controlling factors |
|---|---|
| I. Discrete settling | Overflow rate |
| II. Flocculent settling | Overflow rate and depth |
| III. Zone settling | Solids flux |
| IV. Compression settling | Solids retention time and sludge depth |

***Type II Settling (Flocculent Settling).*** Type II settling occurs when particles initially settle independently but flocculate as they proceed to the bottom of the tank. As a result of flocculation, the settling velocities of the aggregates formed change with time, and a strict mathematical solution is not possible. Laboratory testing is required to determine appropriate values for design parameters.

Type II settling can occur during clarification following fixed-film processes, primary clarification of wastewater, and clarification of potable water treated with coagulants. It can also occur above the sludge blanket in clarifiers following activated sludge treatment; however, design procedures based on type III settling are typically used to design these units.

A batch-type laboratory procedure was developed to estimate the necessary surface overflow rate (SOR), detention time or basin depth, and percent removal of suspended solids. The procedure, described in most textbooks (Reynolds and Richards, 1996; Metcalf and Eddy, 2003), follows:

1. Use a batch settling column equal to the proposed clarifiers depth and 120 mm by 200 mm (5 in. by 8 in.) in diameter, with sampling ports at equal intervals (Figure 4.2).
2. Determine the initial suspended-solids concentration of the suspension under study.
3. Mix the suspension thoroughly and transfer the contents rapidly into the column to ensure a uniform mixture. Care should be taken to avoid shearing of particles.
4. The procedure should be carried out under quiescent conditions and the temperature within the column should not vary more than 1°C (1.8°F).
5. Samples are collected from each port at selected intervals. The total time that samples are collected should at least equal the detention time of the clarifier.
6. The percentage removal of total suspended solids is computed for each sample.
7. Percent removal values are plotted as numbers on a set of coordinate axes labeled tank depth (*H*) on the ordinate and sampling time (*t*) on the abscissa (Figure 4.3).
8. Curves of equal percentage removal (isopercent curves $R_1$ through $R_6$) are drawn through the points, interpolating where necessary.
9. A series of detention times are selected. The percentage removal and SOR corresponding to each are computed according to

$$\mathrm{SOR} = V_c = H/t \tag{4.1}$$

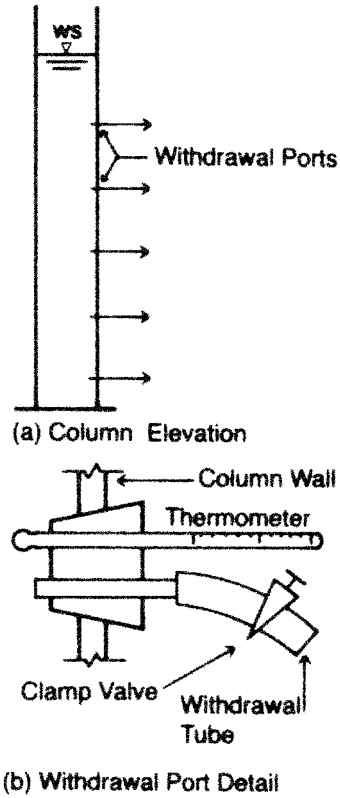



**FIGURE 4.2** Batch settling test.

Where

$H$ = the settling column height (m),
$t$ = the detention time selected (min), and
$V_c$ = settling velocity (m/min).

and overall percentage removal, as given by

$$R = \Sigma\ (\Delta h/H)\{(R_n + R_{n+1})/2\} \tag{4.2}$$

Where

$R$ = overall removal (%),
$\Delta h$ = vertical distance between adjacent isopercent curves (m),
$H$ = total height of settling column (m), and
$R_n$ and $R_{n+1}$ = isopercent curve numbers $n$ and $n + 1$.

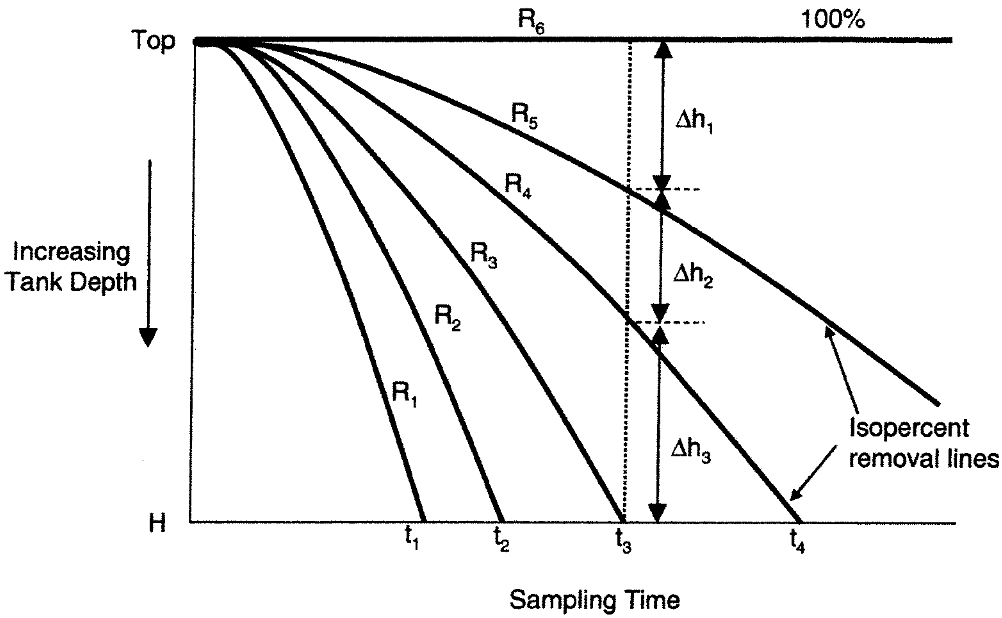



**FIGURE 4.3** Batch settling curves.

For example, as shown in Figure 4.3, the overall solids removal at detention time $t_3$, and depth $H$ is

$$(\Delta h_1/H)\{(R_5 + R_6)/2\} + (\Delta h_2/H)\{(R_4 + R_5)/2\} + (\Delta h_3/H)\{(R_3 + R_4)/2\}$$

10. Plot computed SOR versus percentage removal. Knowing SOR, percentage removal can be obtained from the graph.
11. Adjust the SOR by appropriate scaleup factors. The U.S. Environmental Protection Agency (U.S. EPA, 1975) suggests that the prototype SOR be adjusted as follows:

$$\text{SOR} = \text{Laboratory value} \div \text{Scaleup factor (1.25 to 1.75)}$$

As this procedure indicates, settling tanks are typically designed using an SOR, detention period, or both and assuming an ideal settling basin. This design method often fails to predict or explain the behavior of tanks under operating conditions because it does not account for concentration or density gradients, wind movement, flow variation, differences in tank shape, inlet–outlet structures, and temperature variations. Scaleup factors such as those suggested in step 11 are required to compensate for these. However, some effort has been made to examine the reliability of the laboratory test procedure and the influence of some of the factors mentioned.

Temperature is an important factor in type II clarifier design, especially those operating at low solids levels, such as clarifiers following fixed-film processes. Increases in water viscosity at lower temperatures retards particle settling in clarifiers and requires extended detention times to maintain the same removal efficiency.

Zanoni and Blomquist (1975) have examined the repeatability of the laboratory design procedure. They found that column diameter (100 mm versus 150 mm [4 in. versus 6 in.]) and number of sampling ports (four versus seven) produced only minor differences in results. Thackston and Eckenfelder (1972) have presented a procedure modification that accounts for the actual hydraulic regime in the clarifier. However, the method requires a tracer curve from a clarifier with a hydraulic regime similar to the one proposed. Inlet and outlet turbulence in clarifiers reduces the effective settling area.

***Type III Settling (Hindered Settling or Zone Settling).*** Type III settling is a predominant mechanism in secondary clarifiers. While type II and type IV settling may occur to a limited extent, it is type III that governs design. In suspensions undergoing hindered settling, the solids concentration is typically much higher than in discrete or flocculent processes. As a result, the contacting particles tend to settle as a zone or blanket and maintain the same position relative to each other.

As settling continues, a clear liquid is produced above the settling zone and particles near the clarifier bottom become compressed and are in close physical contact. Thus, the solids concentration in the sludge blanket increases with depth and solids are continuously removed as they reach the design underflow concentration. Key variables that affect clarifier performance are listed in Table 4.4.

Determination of maximum allowable SLR could be refined using experimentally determined settling velocities and solids flux analyses. The method of design now widely accepted is based on work by Coe and Clevenger (1916), Dick and Ewing (1967), Dick and Young (1972), and Yoshioka et al. (1957). It involves determining the total solids flux that can be applied to a clarifier. The total flux consists of two components: settlement of the sludge induced by gravity and bulk movement of sludge and water induced by sludge withdrawal from the clarifier bottom.

The gravity flux component is based on the settling velocity of the sludge, which is assumed equal to the sludge interface settling velocity. The bulk flux component is

**TABLE 4.4** Variables affecting clarification and thickening.

| Clarification | Thickening |
|---|---|
| Wastewater | Wastewater |
| Flowrate | MLSS flowrate |
| Wastewater temperature | |
| Tank | Tank |
| Surface area, solids loading rate, and SOR | Surface area |
| Depth | Depth |
| Weir length, position, and weir loading | Sludge-collection device |
| Inlet device | |
| Tank configuration | |
| Sludge-collection device | |
| Hydraulic pattern | |
| Wave and wind action | |
| Sludge | Sludge |
| Mass loading | Settling rate |
| Sludge settling rate | Compaction characteristics |
| Compaction characteristics | MLSS concentration and solids loading |
| Degree of nitrification | Recycle ratio |
| Sludge blanket control | Sludge blanket control |
| Biological process | |
| Process mode | |
| Biochemical oxygen demand loading | |

calculated from the velocity within the tank induced by sludge withdrawal. If 100% solids capture is assumed (i.e., no effluent suspended solids), then the solids flux past a horizontal plane within the clarifier may be obtained by adding the gravity and bulk fluxes. At steady state, this also represents the solids flux that can be applied to a clarifier producing a specified underflow concentration at a specified withdrawal rate.

As with all laboratory design procedures discussed thus far, several factors not accounted for limit the usefulness of the solids flux theory in predicting the nonideal performance of settling tanks. These include conditions at the inlet, the sludge removal outlet related to velocity distribution, density currents, and other related factors. Wilson and Lee (1982) and Riddell et al. (1983) showed that the procedure for applying the solids flux theory could be simplified. According to the authors, their procedure also makes it simpler for the designer to account for nonideal performance, loading variations, and change in settling characteristics.

Keinath et al. (1977) used a systems approach, based on the solids flux method, to design and operate an activated sludge/clarifier system. The approach enables design engineers to evaluate the economic tradeoffs between alternative system designs and establish a least-cost design. Once the system has been constructed, the same approach can be used by plant operations personnel to establish the operational state of the system and subsequently make rational decisions regarding required control actions or responses.

In a settling basin that is operating at a steady state, a constant flux of solids is moving downward (Figure 4.4). The total mass flux, $SF_t$, of solids is the sum of the

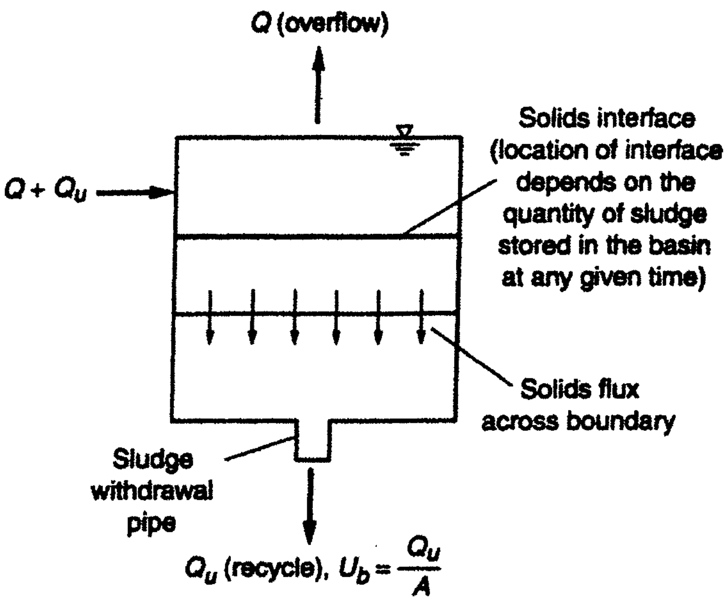



**FIGURE 4.4** Settling basin at steady state ($u_b$ = bulk downward velocity, m/h or ft/hr, and $A$ = required area, $m^2$ or sq ft) (Metcalf and Eddy, 2003).

mass flux resulting from hindered settling due to gravity, $SF_g$, and the mass flux resulting from bulk movement of the suspension, $SF_u$. The solids flux across any arbitrary boundary resulting from hindered settling is;

$$SF_g = X_i V_i \tag{4.3}$$

Where

$SF_g$ = solids flux resulting from gravity, kg/m²·h (lb/sq ft/hr),
$X_i$ = solids concentration at point in question, g/m³ (lb/cu ft), and
$V_i$ = settling velocity of solids at concentration $X$, m/h (ft/hr).

The solids flux resulting from underflow, $SF_u$, is

$$SF_u = X_i U_b \tag{4.4}$$

and

$$U_b = Q_u / A \tag{4.5}$$

hence,

$$SF_u = X_i Q_u / A \tag{4.6}$$

Where

$SF_u$ = solids flux resulting from underflow, kg/m²·h (lb/sq ft/hr),
$U_b$ = bulk downward velocity, m/h (ft/hr),
$Q_u$ = underflow flowrate, m³/h (cu ft/hr), and
$A$ = required area, m² (sq ft).

The total solids flux, $SF_t$, in kg/m²·h (lb/sq ft/hr), is the sum of these two components:

$$SF_t = X_i V_i + X_i U_b \tag{4.7}$$

In eq 4.7, the solids flux resulting from gravity (hindered) settling depends on the solids concentration and the settling characteristics of the solids at that concentration. The procedure entails the following steps:

- Settling tests are conducted at different solids concentrations ($C_1$, $C_2$, and $C_3$) and a set of settling curves (interface height versus time) is generated as shown in Figure 4.5.
- From the settling curves, the hindered settling velocity, $V_1$, $V_2$, and $V_3$ (slope of the linear portion of the respective curves in Figure 4.5a), is determined for the

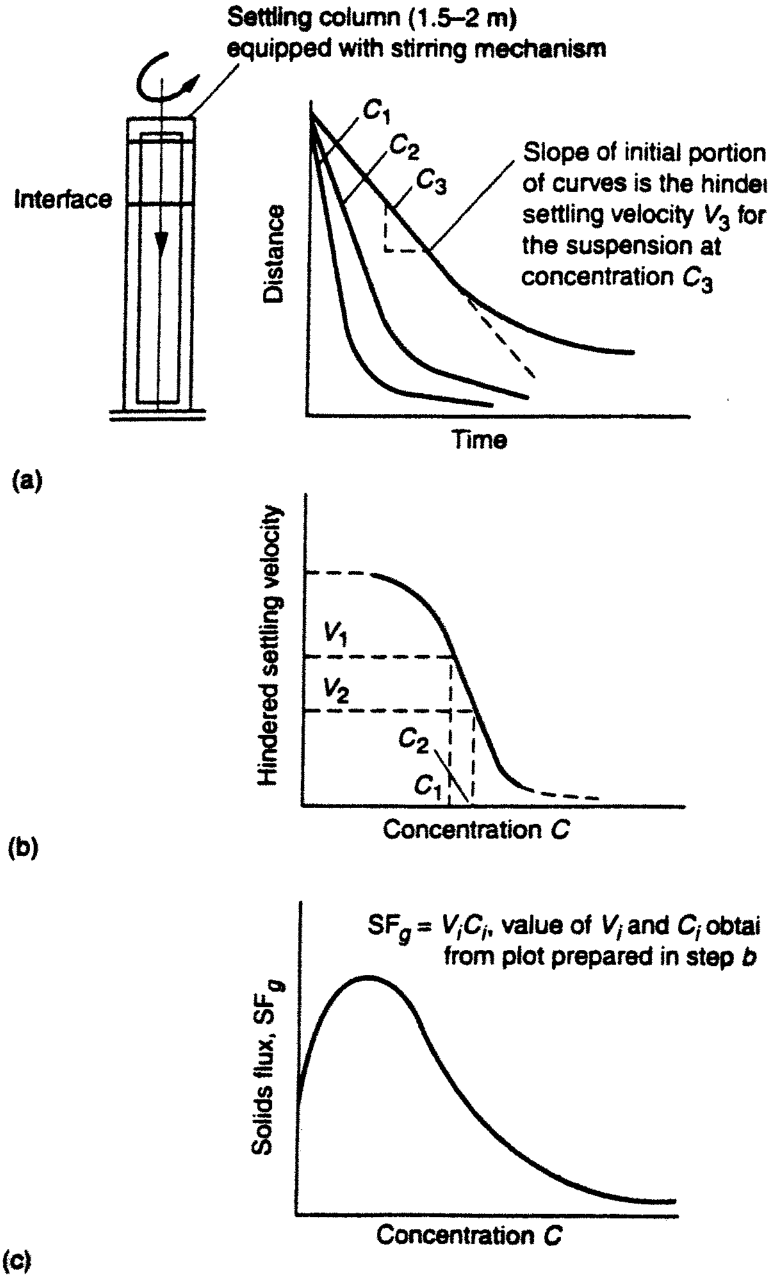
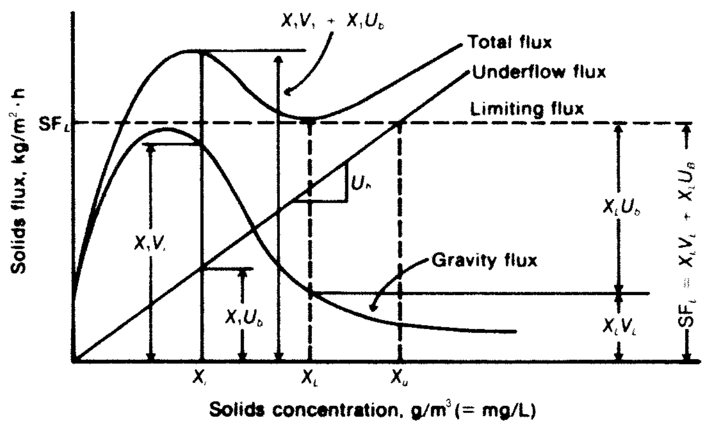



**FIGURE 4.5** Procedure for developing solids flux curves.

MLSS concentrations ($C_1$, $C_2$, and $C_3$) and plotted as a function of the solids concentration (Figure 4.5b).

- The gravity solids flux ($SF_g$) is computed using eq 4.3 and plotted against the corresponding solids concentration as illustrated in Figure 4.5c.

The solids flux resulting from bulk transport is a linear function of the concentration with slope equal to $U_b$, the underflow velocity (Figure 4.6). The total flux, which is the sum of the gravity and the underflow flux, is also shown in the figure. Increasing or decreasing the flowrate of the underflow causes the total flux curve to shift up or down. Because the underflow velocity can be controlled, it is used for process control. The required cross-sectional area of the thickener is determined by drawing a horizontal line tangent to low point on the total flux curve (Figure 4.6). The point of intersection of this line with the vertical axis represents the limiting solids flux, $SF_L$ that can be processed through the basin. The corresponding underflow concentration is obtained by dropping




Definition sketch for settling data analysis using the solids flux method of analysis. Where:
$X_i$ = solids concentration at the point in question,
$U_b$ = bulk downward velocity, m/h, caused by recycle,
$V_i$ = settling velocity of solids at concentration $X_i$
$X_L$ = influent solids concentration at limiting flux (mg/L),
$X_u$ = underflow solids concentration at limiting flux (mg/L),
$SF_L$ = limiting solids flux (kg/m²·h), and
$V_L$ = settling velocity of solids at limiting flux (m/h).

**FIGURE 4.6** Solids flux curve analysis.

a vertical line to the $x$-axis from the intersection of the horizontal line and the underflow flux line. If the quantity of solids fed to the settling basin is greater than the limiting solids-flux value, the solids will build up in the settling basin and, if adequate storage capacity is not provided, ultimately overflow at the top. Using the limiting solids-flux value, the required area derived from a solids balance is given by

$$A = \frac{(1+\alpha)QX}{SF_L} \tag{4.8}$$

Where

$A$ = area, $m^2$ (sq ft),
$\alpha$ = recycle ratio ($Q_r/Q$),
$X$ = influent solids concentration, $g/m^3$ (lb/cu ft),
$SF_L$ = limiting solids flux, $kg/m^2 \cdot h$ (lb/sq ft/hr),
$Q_r$ = recycle flowrate, $m^3/h$ (cu ft/hr) and
$Q$ = flowrate to clarifier, $m^3/h$ (cu ft/h).

An alternative graphical method of analysis is presented in Figure 4.7 (Yoshioka et al., 1957). The basic theory is that for a given underflow concentration, there is a

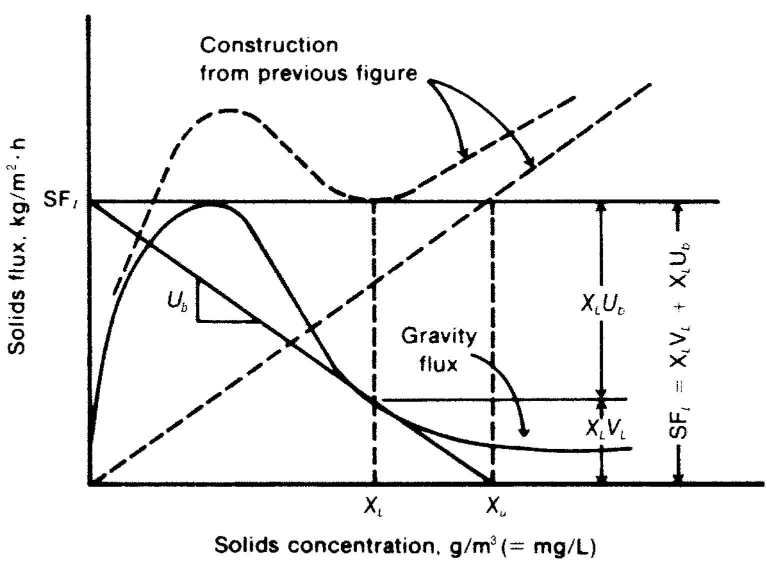



**FIGURE 4.7** Graphical solids flux solution.

maximum amount of solids that can pass through the clarifier (limiting flux). A line passing through the underflow concentration (abscissa) and tangent to the gravity flux curve when extended provides the associate limiting flux on the ordinate. The geometric relationship of this method to that given in Figure 4.6 is shown by the dashed lines in Figure 4.7. Based on the limiting flux value, the required clarifier area can be determined using eq 4.8.

The reader is referred to Chapter 6 for a detailed mathematical treatment of the flux theory.

Several models have been developed linking the initial settling velocity (ISV or $V_i$) with solids concentration. Of these, the most common are the exponential model (Vesilind, 1968) and the power model (Dick and Young, 1972). The generic forms of these two models are as follows:

$$\text{ISV} = a \exp(-nX_i) \qquad \text{(Exponential model)} \tag{4.9}$$

$$\text{ISV} = a' X_i^{-n} \qquad \text{(Power model)} \tag{4.10}$$

Where

ISV = initial (hindered) settling velocity (m/d),
$X_i$ = MLSS (g/L), and
$a$, $a'$, $n$ = sludge-settling constants.

Riddell et al. (1983) have developed graphical methods based on both models. As shown in Figure 4.8, the exponential model yields a family of curves relating $A/Q$, which is the inverse of SOR, $R$ (RAS rate) for various MLSS concentrations ($X_{LP}$). The dashed line represents the minimum or critical RAS rate ($R_c$) required to obtain the minimum clarifier area for a given MLSS. Left and right of the dashed line represent portions of the MLSS curves for $R < R_c$ and $R > R_c$, respectively. For practical application, the theoretical clarifier area obtained from the graphical method is multiplied by a safety factor (SF) to account for variations in MLSS concentrations, sludge volume index (SVI) values, and flowrates. The curves for the power model, shown in Figure 4.9, provide minimum RAS rates for various safety factors and power model coefficients (Riddell et al., 1983).

The use of the graphical model may be illustrated by the assuming the following:

MLSS = 3500 mg/L,
SF = 2.0,
$Q$ = 28 575 $m^3/d$ (7.5 mgd), and
$R$ = 60%.

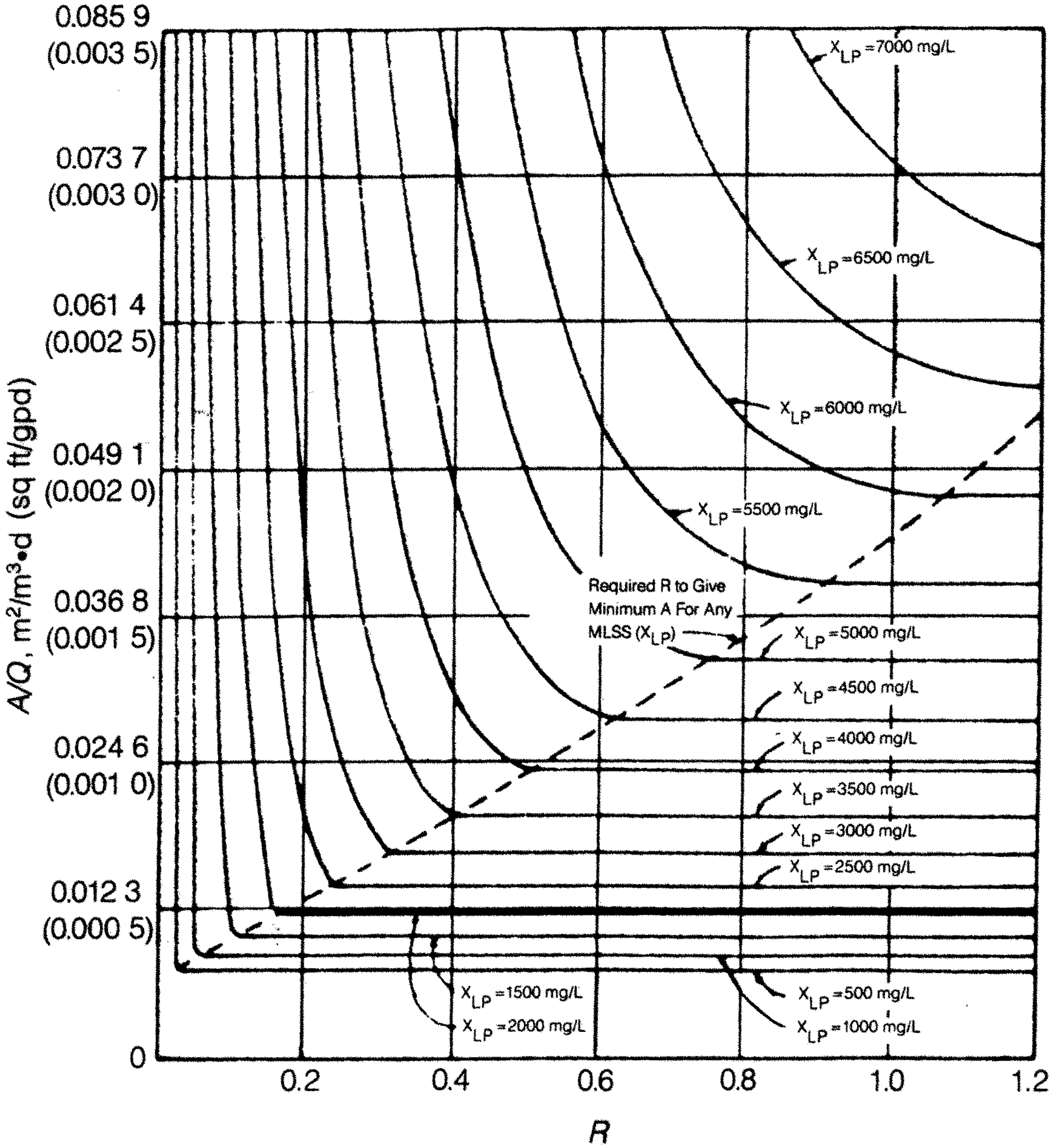



**FIGURE 4.8** $A/Q$ versus $R$ curves for various MLSS ($X_{LP}$) (Riddell et al., 1983).

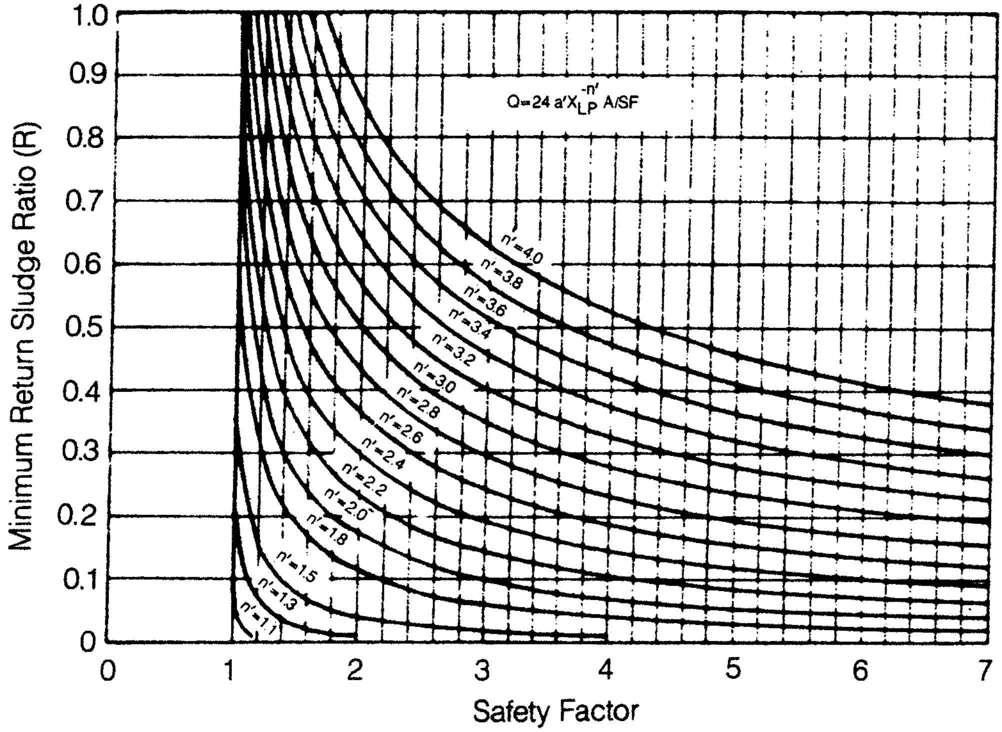



**FIGURE 4.9** Return ratios required for the power model at various safety factors (Riddell et al., 1983).

From Figure 4.8, the intersection of 3500 mg/L MLSS line and the dashed line (minimum RAS line) yields $A/Q = 0.02$ m$^2$/m$^3$·d. Hence, the minimum clarifier surface area ($A$) = (0.02 m$^2$/m$^3$·d) * (28 575 m$^3$/d) * 2.0 = 1143 m$^2$.

Wilson and Lee (1982) presented the following equation to determine the maximum allowable hydraulic loading rate as a function of ISV at the design MLSS concentration:

$$Q/A = 24 * \text{ISV}/\ \text{CSF}$$

Where

$Q$ = flow or limiting hydraulic capacity (m$^3$/d),
$A$ = clarifier surface area (m$^2$),
ISV = $V_o$ exp (-$k$*MLSS) (m$^3$/m$^2$·h),

CSF = clarifier safety factor,
$V_o$ = sludge settling characteristic ($m^3/m^2 \cdot d$),
$k$ = sludge settling characteristic (L/g), and
MLSS = mixed liquor suspended solids (g/L).

The sludge-settling characteristics ($V_o$ and $k$) are typically obtained by linear least-square regression of ISV against MLSS data over the range of concentrations. In addition, Wilson (1996), based on review of plant operating data, concluded that $V_o$ is dependent on wastewater temperature and is approximately equal to 0.3 to 0.5 times the temperature in degrees Celsius. He also suggested that the value of $V_o$ is depressed by the volatile solids level of the sludge and increased by the addition of a polyelectrolyte.

The Wilson and Lee model assumes a sufficiently high value of sludge removal and rate of return sludge pumping. Typically, one should provide for rates of 100 to 150%. The equation includes a safety factor (CSF) for scaleup. For CSF values up to three, the equation provides results consistent with other, more basic clarifier analyses. A CSF of 2 would be typical for systems known to have stable sludge-settling properties, limited flowrate fluctuations, or step-feed flexibility. A *minimum* safety factor of 1.5 is considered necessary by the authors. In the above analysis, the engineer must recognize that the ISV will change with MLSS concentration and other conditions as shown in Figure 4.10. Maximum anticipated operational MLSS or the corresponding minimum ISV should be used in the equation.

***Type IV Settling (Compression Settling).*** In type IV settling, particles have reached such a concentration that a structure is formed and further settling can only occur by compression. This type of settling typically occurs in the lower layers of a deep sludge mass such as near the bottom of secondary clarifiers and sludge thickeners (Tchobanoglous and Schroeder, 1985).

The sludge consolidation rate in this region is proportional to the difference in the height, $H$, at time, $t$, and the height to which the sludge will settle after a long period of time. This can be presented as

$$H_t - H = (H_2 - H)\, e^{-i(t-t_2)} \tag{4.11}$$

Where

$H_t$ = height of settled sludge at time ($t$),
$H$ = height of settled sludge after a long period of time (approximately 24 hours),
$H_2$ = height of settled sludge at time ($t_2$),
$i$ = constant for a given suspension, and
$e$ = base of the naperian logarithm system.

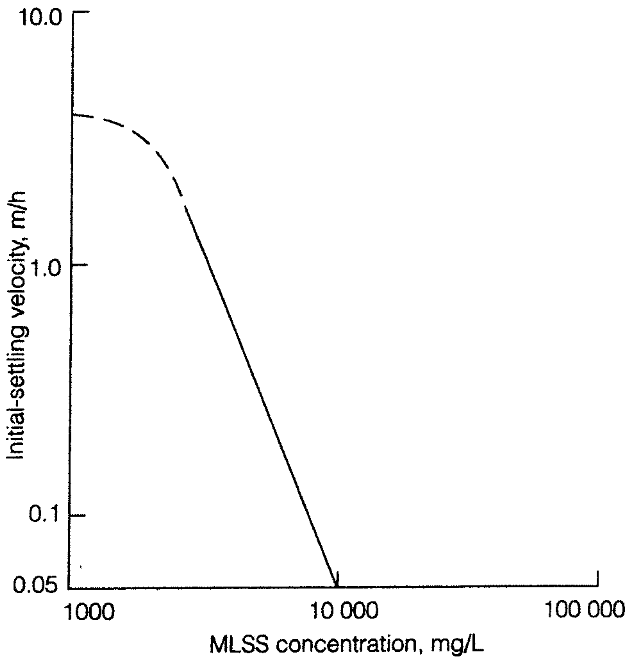



FIGURE 4.10 The dependency of the ISV on MLSS.

The equation form points out that if thickening is desired, sufficient time or depth must be provided to for this to happen to levels predicted by laboratory thickening analysis. However, it should be pointed out that thickening in clarifiers is the source of many operational problems and should be avoided if possible. The above equations apply to batch thickening and not to continuous thickening, which occurs in final clarifiers.

**FACTORS AFFECTING SLUDGE SETTLEABILITY.** The primary factor affecting sludge settleability is the microbial makeup. A well-designed and well-operated activated sludge system provides an environment promoting the proliferation of desired microorganisms (floc formers) and suppressing the growth of nuisance organisms (filaments) that contribute to poor sludge settleability and foaming. The filament content of the sludge is influenced by the following factors (Ekama et al., 1997):

- Wastewater characteristics: industrial content, soluble substrate, temperature, pH, total dissolved solids, oil and grease content, septicity, combined or separate sewers, characteristics of recycle streams, etc.

- Biological reactor: configuration, operating conditions (anoxic/anaerobic/oxic), solids retention time (SRT), MLSS, dissolved oxygen levels, etc.

Some of these factors are discussed below.

***Microbial Makeup.*** Activated sludge microorganisms that settle and thicken well are generally referred to as floc formers. These organisms include a mixture of bacteria, protozoa, and metazoa. Some of the more common types are listed in Table 4.5.

The ability of the sludge to flocculate, settle, and thicken is primarily affected by nonfloc formers, or filamentous organisms. When viewed under a microscope, they are typically long and stringy in appearance. Such filaments protruding from flocs are believed to prevent biomass compaction. Some researchers (Jenkins et al., 2003; Sezgin et al., 1978) contend that an ideal floc contains just the right mixture of filamentous microorganisms and floc formers, with the filaments forming the backbone of the floc (Figure 4.11a). They contend that if the floc lacks enough filaments, it is likely to breakup (Figure 4.11b) and effluent quality deteriorates. If too many filaments exist, bulking may develop (Figure 4.11c).

To date, approximately 60 different filamentous organisms have been implicated with poor settling sludge, and the number is growing. Table 4.6 lists the 18 most prevalent filamentous organisms identified at 270 treatment plants (525 samples) in the United States (Jenkins et al., 2003).

Jenkins et al. (2003) linked dominant filament types to causative operating conditions as shown in Table 4.7.

**TABLE 4.5** Common types of bacteria and protozoa.

| Bacteria | Protozoa |
|---|---|
| *Pseudomonas* | *Paramacium* |
| *Archromobacter* | *Aspidisca* |
| *Flavobacterium* | *Vorticella* |
| *Alcaligenes* | |
| *Arthrobacter* | |
| *Citromonas* | |
| *Zooglea* | |
| *Acinetobacter* | |

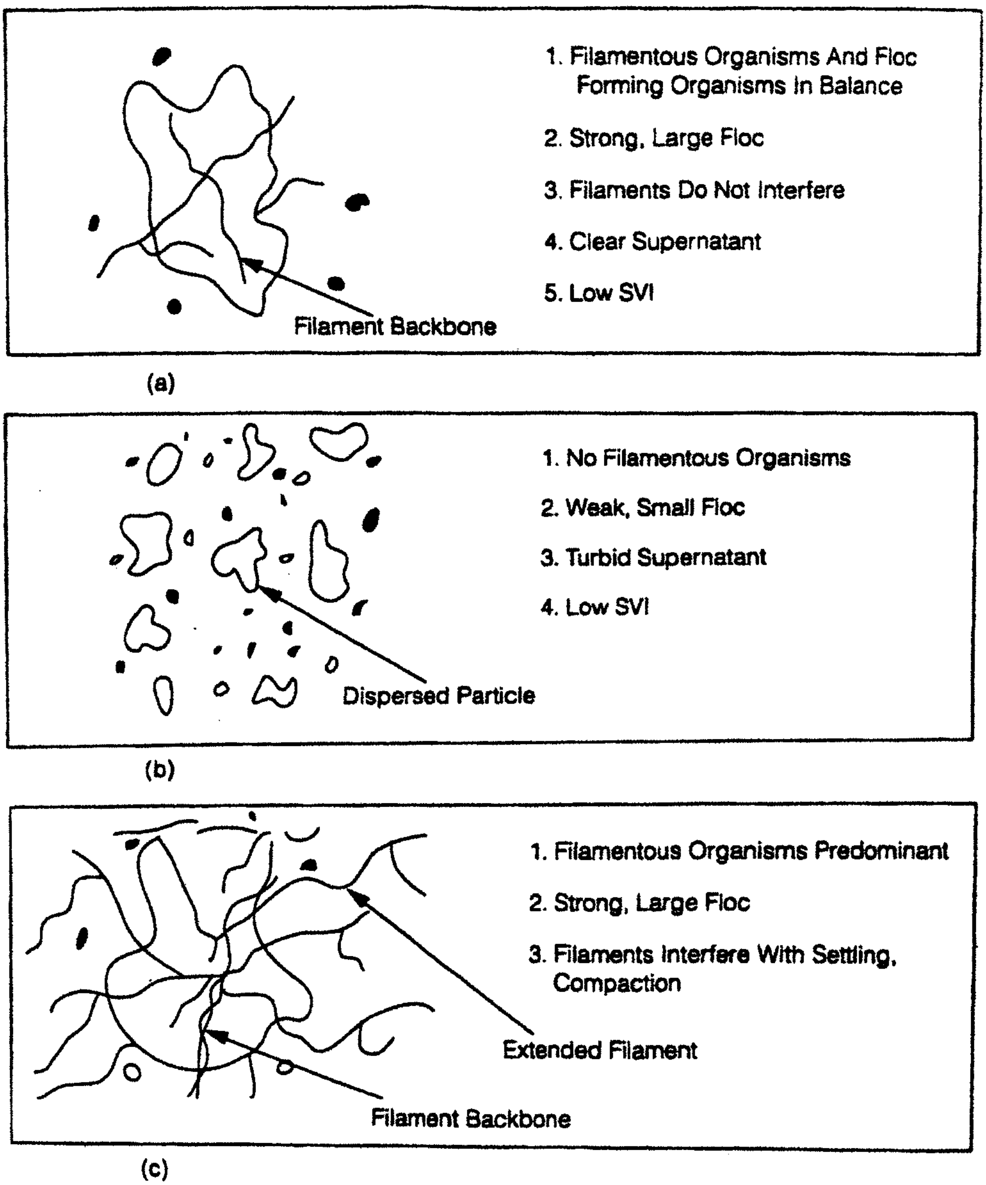



FIGURE 4.11 Effect of filamentous organisms on activated sludge structure: (a) ideal, nonbulking floc; (b) pin-point floc; and (c) filamentous, bulking (reprinted from *Secondary Settling Tanks*, ISBN: 190020035, with permission from the copyright holder, IWA).

**Table 4.6** Dominant filamentous organisms identified in wastewater treatment plants in the United States (Jenkins et al., 2003).

| Filamentous organism | Treatment plants with bulking or foaming in which filaments were (%) | |
|---|---|---|
| | Dominant | Secondary |
| *Norcardia* spp. | 31 | 17 |
| Type 1701 | 29 | 24 |
| Type 021N | 19 | 15 |
| Type 0041 | 16 | 47 |
| *Thiothrix* spp. | 12 | 20 |
| *Sphaerotilus natans* | 12 | 19 |
| *Microthrix parvicella* | 10 | 3 |
| Type 0092 | 9 | 4 |
| *Haliscomenobacter hydrossis* | 9 | 45 |
| Type 0675 | 7 | 16 |
| Type 0803 | 6 | 9 |
| *Nostocoida limicola* | 6 | 18 |
| Type 1851 | 6 | 2 |
| Type 0961 | 4 | 6 |
| Type 0581 | 3 | 1 |
| *Beggiatoa* spp. | 1 | 4 |
| Fungi | 1 | 2 |
| Type 0914 | 1 | 1 |
| Other | 1 | — |

Finally, there are different types of filaments. Some are short, whereas others are long and coiled. Filamentous organisms that are short may not affect sludge settleability, even when present in significant numbers as much as a smaller number of long and coiled filaments. Hence, filament length is a better indication of sludge settleability than the number of filaments.

**TABLE 4.7** Filament type and causative agent (Jenkins et al., 2003).

| Causative condition | Filament types |
|---|---|
| Low F/M | *Microthrix parvicella, Haliscomenobacter, Nocardia* spp., types 0041, 0092, 0581, 0675, 0803, 0914, and 1851 |
| Low dissolved oxygen | Type 1701, *S. natans, H. hydrossis,* and *M. parvicella* |
| Presence of sulfide | *Thiothrix* spp., *Baggiatoa* spp., types 021N and 0914 |
| Readily metabolizable soluble organics | *S. natans, Thiothrix* spp., *H.* hydrossis, *N. limicola* II, *N. limicola* III, and types 021N, 0914, 1701, and 1851 |
| Low pH | Fungi |
| Nitrogen deficiency | *Thiothrix* spp. and type 021N |
| Phosphorus deficiency | *S. natans, H. hydrossis,* and *N. limicola* III |

The designer's task is to design a system that discourages the growth or accumulation of bulking and nuisance microorganisms. The designer also must provide in the design flexibility that allows the operator to control any nuisance organisms that may appear in the system.

***Nonsettleable Solids.*** Nonsettleable solids are those that, because of their size being too small or their density being too close to that of the surrounding fluid, settle at a negligible rate. Consequently, these solids are not removed in a typical final clarifier. They have low tendency to flocculate or have sheared from floc particles because of excessive turbulence in the aeration basin or in the conveyance system. Formation and escape of too many small and dispersed solids represent clarification failure leading to potential effluent noncompliance.

The degree of flocculation has a direct effect on clarifier performance and can be quantified by performing the dispersed suspended solids (DSS) test. The DSS test, originally developed by Parker et al. (1970) and used by Das et al. (1993), is defined as the supernatant suspended solids concentration following 30 minutes of settling in a Kemmerer sampler (Wildlife Supply Company, Buffalo, New York). In essence, DSS is a "snapshot" of the state of flocculation (or breakup) at the time of sampling. Parker and Stenquist (1986) reported a close approximation of DSS to effluent suspended solids (ESS) from a clarifier with a flocculator center well and not subjected to short-circuiting, and denitrification.

Wahlberg, et al. (1995) developed the flocculated suspended solids (FSS) test, which measures the flocculation potential of a mixed liquor sample. It is not to be confused with the DSS, test, which quantifies the actual state of flocculation of a sample. The FSS test, performed under ideal flocculation and settling conditions, is operationally defined as the supernatant suspended solids concentration after 30 minutes of flocculation followed by 30 minutes of settling.

Table 4.8 summarizes the guidance provided by Wahlberg (2001) for interpreting DSS/FSS data for municipal publicly owned treatment works.

A discussion of nonsettleable solids with respect to primary clarification may be found in Chapter 2.

**Table 4.8** Interpretation of DSS/FSS data (Wahlberg, 2001).

| Condition at the clarifier | Potential causes |
|---|---|
| High FSS (>10 mg/L) | Dispersed growth in aeration basin or deflocculation of clarifier influent |
| Influent DSS > effluent DSS, and effluent DSS = FSS | Good flocculation in the clarifier; no hydraulics problems |
| Influent DSS = effluent DSS, effluent DSS = FSS, and ESS = FSS | Mixed liquor well flocculated before entering the clarifier and there are no hydraulics problems |
| Regardless of the influent DSS, effluent DSS = FSS and DSS < ESS | Good flocculation but there are hydraulics problems |
| Regardless of influent DSS, effluent DSS > FSS and DSS < ESS | Both flocculation and hydraulics problems |
| Regardless of influent DSS, effluent DSS > FSS and DSS = ESS | Flocculation problems and tank hydraulics good |

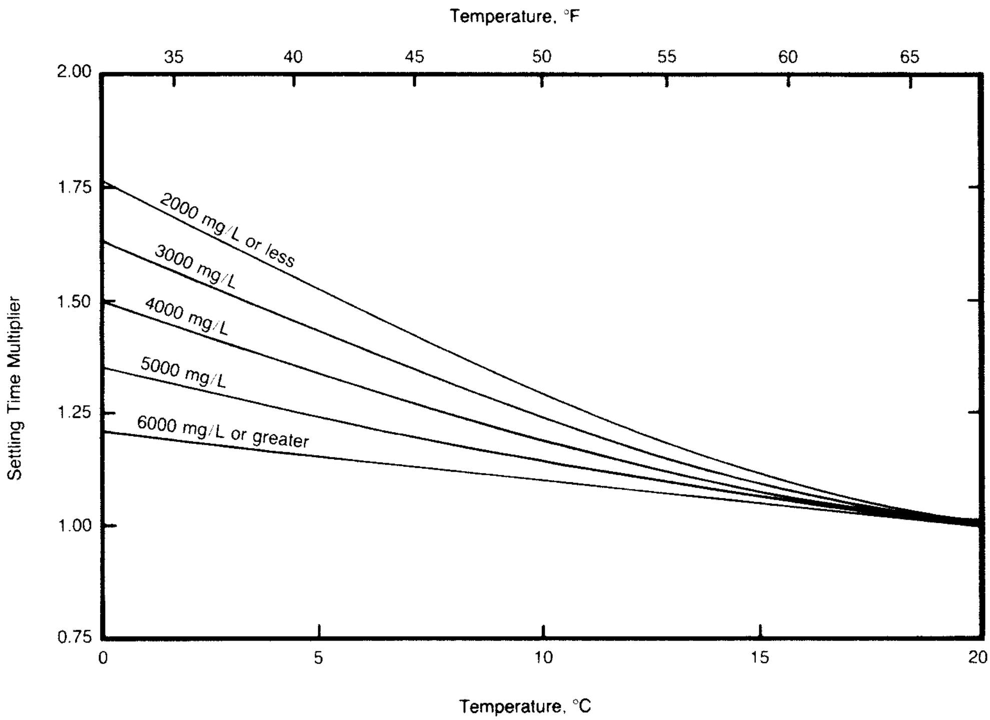



**FIGURE 4.12** Effect of temperature on settling detention time.

***Effect of Temperature.*** Temperature is one of the key factors affecting the sedimentation process in secondary clarifiers. Reed and Murphy (1969) have investigated the effect of temperature on type III settling, which governs the design and performance of secondary clarifiers following the activated sludge process. They noted that the settling times at $0^0$C increased by a factor of 1.75 over those at $20^0$C for a MLSS concentration of 2000 mg/L (Figure 4.12). However, this temperature effect became less pronounced as the solids concentration increased.

As noted previously, Wilson (1996) noted the temperature dependency of $V_o$, the sludge settling constant in the Vesilind ISV equation (eq 4.9). Based on review of plant data, he concluded that $V_o$ (m/h) is equal to 0.3 to 0.5 times the temperature in degrees Celsius or equal to 1.0 to 1.5 times temperature (°C) when $V_0$ is expressed in feet per hour.

**MEASUREMENT OF SLUDGE SETTLEABILITY.** Sludge settleability is central to the health of the biological system. Ironically, settleability is influenced by conditions in the activated sludge basin but manifests itself in the clarifier. Poor settling sludge causes lower underflow (RAS) solids concentration because of poor compaction. When the RAS solids concentration required by the recycle ratio is not achieved, fewer solids are removed from the tank than applied to it (Ekama et al., 1997). If this condition persists, the sludge blanket can propagate to the surface of the clarifier, resulting in loss of solids in the effluent. Consequently, measuring sludge settleability is fundamental to the operation and control of the biological system. Two basic approaches are used in measuring sludge settleability:

- Volume of settled sludge after a given period of time and
- Settling velocity of the sludge/liquid interface during zone settling.

The following is a brief discussion of the various parameters used in expressing sludge settleability. A more detailed discussion of the topic may be found in Ekama et al. (1997).

***Sludge Volume Index.*** Historically, the SVI has been used most commonly as a measure of sludge settleability. It is the volume in milliliters occupied by 1 g of the suspended solids following 30 minutes of settling of the aeration basin MLSS. The test may be carried out in a 1- or 2-L settling column. *Standard Methods* (APHA et al., 1999) specifies gently stirring the sample during settling to eliminate or minimize wall effects. Dick and Vesilind (1969) noted that, for the same samples, slow stirring yielded consistently lower SVI values than the unstirred tests. In addition, stirred SVI appears to be less affected by solids concentration. In spite of these benefits, many plant operators continue to use the unstirred settled volume test. Sludge volume index is expressed as follows:

$$\text{SVI (mL/g SS)} = V_{30}(1000\ \text{mg/g})/(XV_t) \tag{4.12}$$

Where

$V_{30}$ = sludge volume after 30 minutes of settling (mL),
$X$ = mixed liquor concentration before the test (mg/L), and
$V_t$ = volume of settling column (L).

The popularity of SVI is partly because of the ease of measurement. However, SVI is not always a good measure of settleability. Dick and Vesilind (1969) have provided several reasons for deficiencies of the SVI. Perhaps the most significant is the dependency on mixed liquor concentration. The authors found that for good settling sludge, the MLSS concentration above which the SVI was influenced by the solids concentration was relatively high. For poor settling sludge, the critical MLSS concentration was low.

***Dilute Sludge Volume Index.*** Dilute sludge volume index (DSVI) was developed to overcome the above-mentioned problem with the traditional SVI test. In this test, an effort is made to keep the 30 minutes settled volume between 150 and 250 mL/L by dilution. Final effluent before chlorine addition is typically used for dilution to minimize the interference from foreign material.

$$\mathrm{DSVI} = \mathrm{DSV}_{30}/X_{\mathrm{dil}} \tag{4.13}$$

Where

$\mathrm{DSV}_{30}$ = settled volume of the diluted sludge after 30 minutes of settling and
$X_{\mathrm{dil}}$ = MLSS concentration following the necessary dilution.

The upper limit of 250 mL/L was selected for $\mathrm{DSV}_{30}$ because the SVI is influenced by solids concentration above this level. Because of the relative insensitivity of $\mathrm{DSV}_{30}$ to solids concentration, it provides a common basis for comparing sludge settleabilities at different facilities.

***Stirred Specific Volume Index at 3.5 g MLSS/L.*** Wall effects plague the traditional SVI test. White (1975, 1976) found that this could be eliminated by slowly stirring the contents of the settling column while it settles. Stirring also minimizes short-circuiting and bridge formation. Consequently, the stirred specific volume index at 3.5 g MLSS/L ($\mathrm{SSVI}_{3.5}$) represents the field conditions more closely than the traditional SVI. It is defined as the volume occupied by 1 g of solids following 30 minutes of settling in a gently stirred (at 1 rpm) settling column at a standard initial concentration of 3.5 g MLSS/L (3599 mg/L). Determination of $\mathrm{SSVI}_{3.5}$ entails (1) performing a range of settling tests at various MLSS values ranging from 2000 to 6000 mg/L, (2) calculating the SSVIs for each concentration, (3) developing a SSVI-concentration graph, and (4) obtaining the SSVI value at 3500 mg/L by interpolation.

A series of studies by Bye and Dold (1996, 1998, 1999) compared the above settleability parameters and their effect on zone settling velocity. They developed a simple mechanistic model to evaluate the effects of sludge characteristics and test parameters on SVI-type indices. Their investigations revealed that sludge settleability and compactability, settling column height, and solids concentration have an interactive effect on the SVI. They also concluded that SVI may show a marked dependency on solids concentration and that, although the DSVI test eliminates the influence of solids concentration on SVI, it may not bear any relationship to the settleability of the test sample.

**CLARIFIER ANALYSIS.** ***Flux Theory.*** The flux theory, described under type III settling, has been used as the basis for designing and analyzing final clarifiers. It is important to understand that the inherent assumption of the flux theory is that solids are continuously removed from the clarifier as they reach the design underflow concentration. A detailed mathematical analysis of the flux theory is presented in Chapter 6.

Using full-scale data collected by the Dutch research agency, Stichting Toegepast Onderzoek Waterbeheer (STOWA), the Dutch Foundation of Applied Water Research, Ekama and Marais (1986) qualitatively verified the flux procedure. The data revealed a typical solids concentration–depth profile presented in Figure 4.13 (Ekama et al., 1997), which consists of the following four zones: the clear water zone

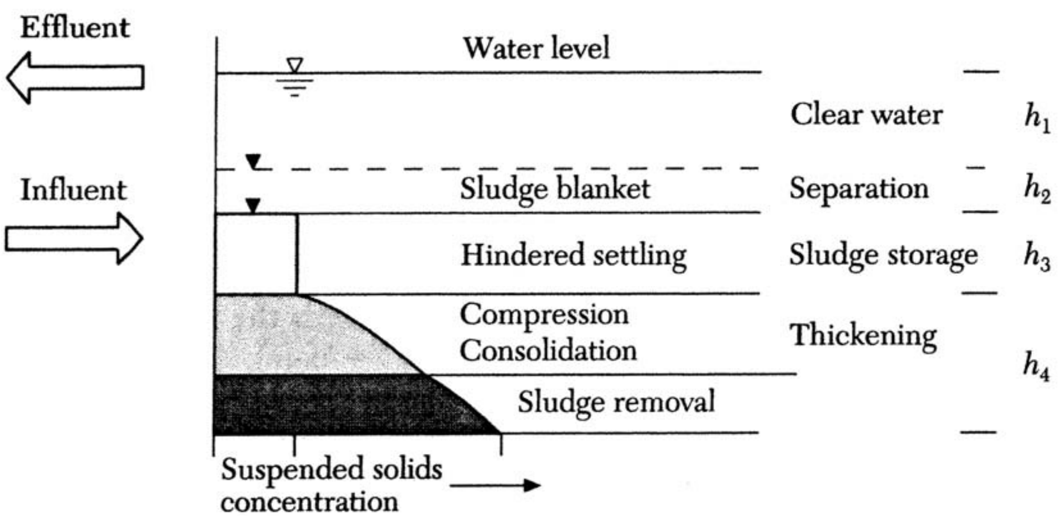



FIGURE 4.13 Typical solids concentration–depth profile assumed in flux analysis (reprinted from *Secondary Settling Tanks*, ISBN: 190020035, with permission from the copyright holder, IWA).

($h_1$), the separation zone ($h_2$), the sludge storage zone ($h_3$), and the thickening and sludge-removal zone ($h_4$).

The fundamental premise of the flux theory is that under overloaded conditions (applied solids flux greater than the limiting flux), a critical zone settling layer (sludge storage zone, $h_3$) develops in the sludge blanket, which limits the conveyance of solids to the bottom of the tank. Consequently, all of the solids that enter the storage zone from the separation zone are not transferred to the thickening zone below and the excess solids accumulate in the storage zone, causing it to expand. As it expands, the solids concentration remains constant throughout the storage layer. The depth of the separation and thickening zones ($h_2$ and $h_4$), however, do not increase. The continued expansion of the storage layer will result in the sludge blanket reaching near the effluent weir, causing a loss of solids with the effluent. At this point, the storage layer cannot expand further and the storage capacity of the clarifier is exhausted. The solids flux that could not be transferred through the storage layer is lost with the effluent.

When the applied solids flux is less than the critical flux (underloaded condition), all of the applied solids can be effectively transferred to the tank bottom and there is no need for solids storage. As s result, the sludge blanket is composed of the separation ($h_2$) and the thickening ($h_4$) zones only.

***State Point Analysis.*** Based on the solids flux approach, Keinath (1985) and Keinath et al. (1977) advanced the concept of *state point,* which is the operating point of a clarifier. State point is the point of intersection of the clarifier overflow rate (OFR) and underflow rate (UFR). It links the operation of the activated sludge basin with that of the clarifier. Consequently, the state point analysis can be used by designers and operators to assess the redistribution of the solids between the aeration basin and settler, optimize the system, and perform "what if" analysis using site-specific and up-to-date settleability data (flux curves).

An overview of the mathematical basis for the state point analysis is presented in Chapter 6. In essence, the state point analysis incorporates the following five factors that influence the transport of solids through the clarifier:

- MLSS concentration,
- Clarifier surface area available for thickening,
- Influent flowrate,

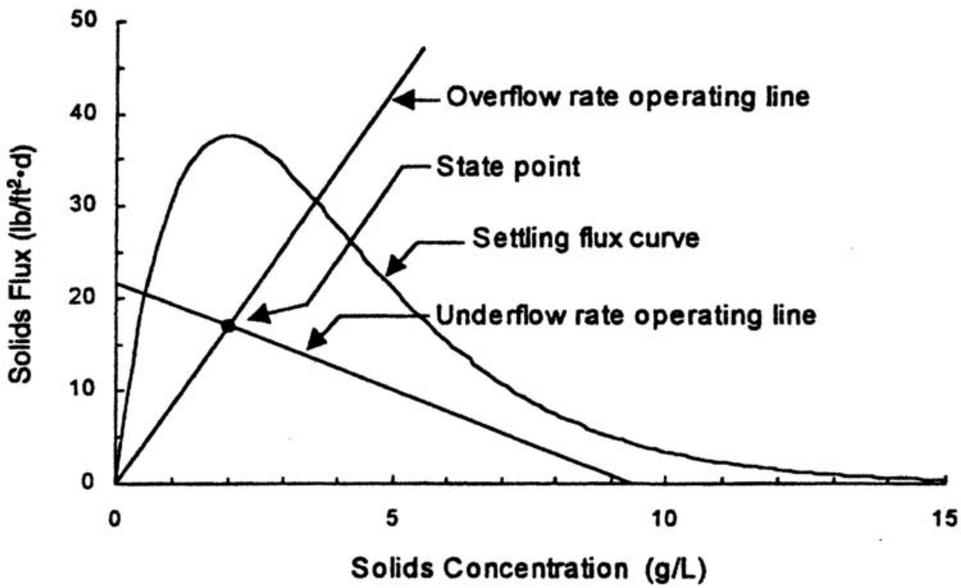



**FIGURE 4.14** Elements of state point analysis (reprinted with permission from Water Environment Research Foundation (2001) *WERF/CRTC Protocols for Evaluating Secondary Clarifier Performance).*

- Return sludge flowrate, and
- Settling characteristics of the mixed liquor.

The components of the state point analysis are illustrated in Figure 4.14. As summarized in Table 4.9, the position of the state point and the location of the UFR line relative to the descending limb of the flux curve determine whether the clarifier is underloaded, critically loaded, or overloaded, as can be seen in Figures 4.14 through 4.19.

In addition to the above corrective actions, critically loaded or overloaded conditions may be relived by lowering the SVI. These strategies are discussed in Clarifier Performance Enhancements section.

The Water Environment Research Foundation/Clarifier Research Technical Committee (CRTC) Protocol (Wahlberg, 2001) provides guidance with respect to the development and application of the state point analysis. The state point approach can be used to analyze the behavior of existing facilities as well as the

**TABLE 4.9** Interpretation of the state point analysis.

| Location of state point | Location of underflow line | Condition of clarifier | Potential corrective action |
|---|---|---|---|
| Within the flux curve (Figure 4.14) | Below the descending limb of the flux curve | Underloaded | None |
| Within the flux curve (Figure 4.15) | Tangential to the descending limb of the flux curve | Critically loaded | Increase RAS rate to become underloaded |
| Within the flux curve (Figure 4.16) | Above the descending limb of the flux curve | Overloaded | Increase RAS rate to become underloaded |
| On the flux curve (Figure 4.17) | Below the descending limb of the flux curve | Critically loaded | Reduce clarifier feed solids to become underloaded<br>Convert to step-feed or Lower MLSS (SRT) |
| On the flux curve (Figure 4.18) | Above the descending limb of the flux curve | Overloaded | Increase RAS rate to become critically loaded |
| Outside the flux curve (Figure 4.19) | | Overloaded | Reduce clarifier feed solids to become underloaded<br>Convert to step-feed or Lower MLSS (SRT) |

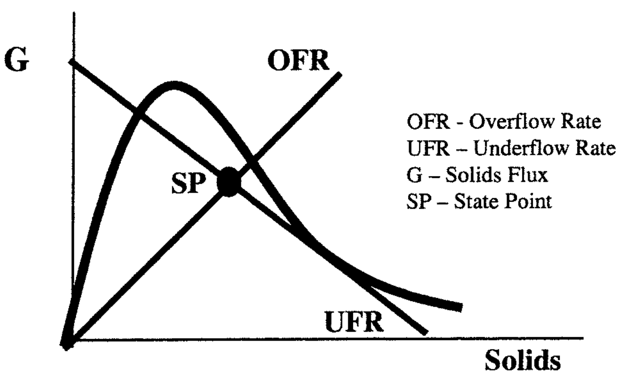



**FIGURE 4.15** Critically loaded clarifier.

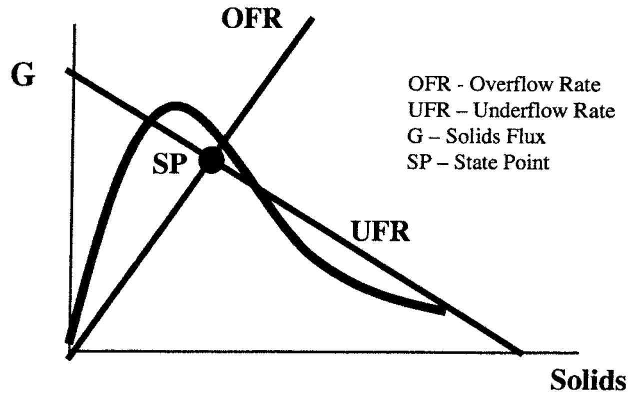



FIGURE 4.16 Overloaded clarifier.

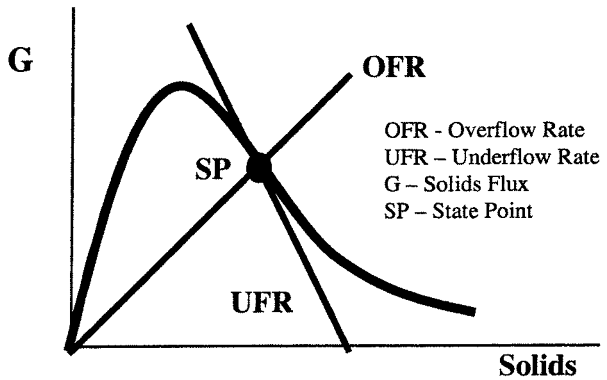



FIGURE 4.17 Critically loaded clarifier.

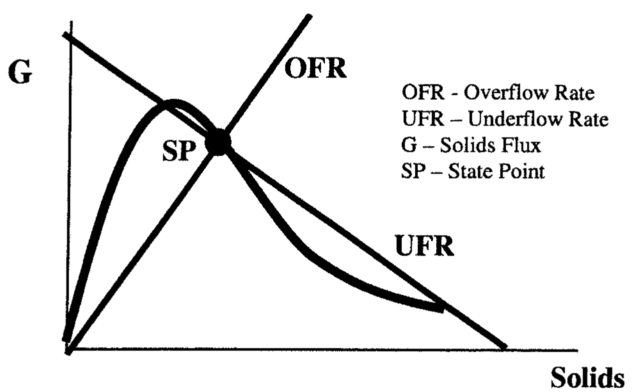



FIGURE 4.18 Overloaded clarifier.

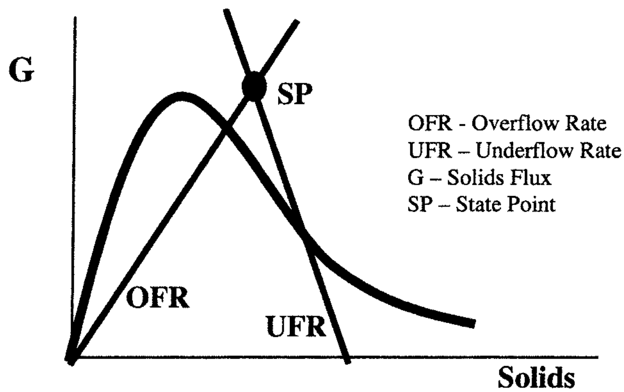



**FIGURE 4.19** Overloaded clarifier.

effect of potential operating scenarios on proposed clarifiers during the design phase. Metcalf and Eddy (2003) present an example on the use of state point analysis in operation and design.

***Other Approaches.*** *THE DAIGGER APPROACH.* Daigger (1995) and Daigger and Roper (1985) developed a convenient clarifier operating diagram (Figure 4.20) by plotting allowable solids loading rate (SLR) as a function of RAS solids concentration for a range of unstirred SVI values. These lines represent the limiting flux for the SVI shown. Finally, lines representing various underflow (RAS) rates are superimposed. Similar operating diagrams can be generated using $SSVI_{3.5}$ and DSVI values (Daigger, 1995).

The clarifier operating point can be located on the diagram by using two of the following operating parameters: actual SLR, underflow rate, or RAS solids concentration. The third parameter, if available, can be used as a check. If the operating point is below and left of the line corresponding to the current SVI, the clarifier is operating below the limiting flux associated with the operating SVI. This means, the clarifier should not be subjected to thickening failure. If the operating point falls on the line representing the current SVI, the clarifier solids loading equals the limiting flux and the clarifier is operating at its failure point. If the operating point falls above and right of the line representing the operating SVI, the clarifier is overloaded with respect to solids loading and thickening failure is likely. The Daigger operating chart can be used to (a) optimize existing system operation, (b) determine the operating

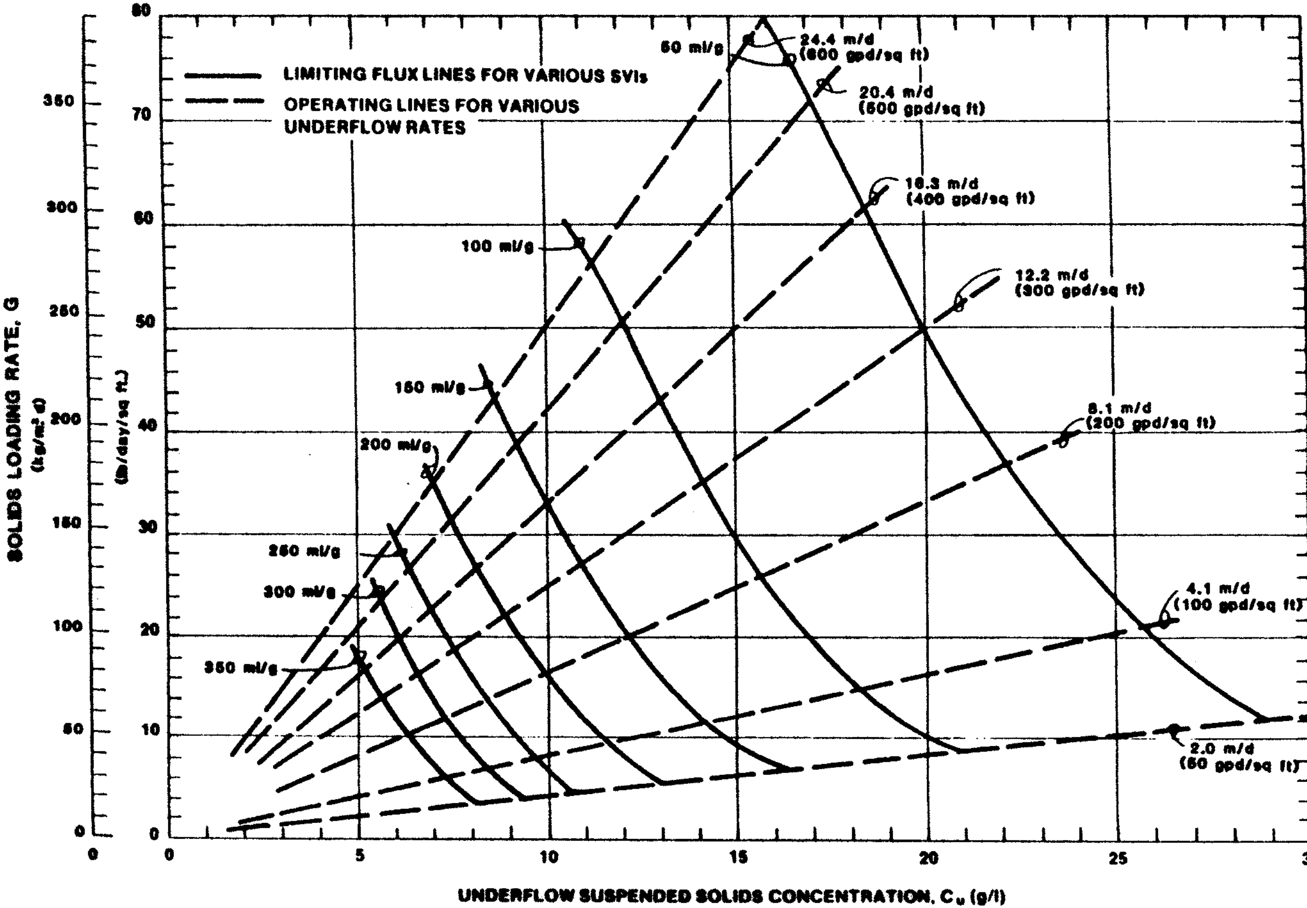



**FIGURE 4.20** Daigger operating chart (Daigger, 1995).

conditions when a different process configuration (such as step-feed) is implemented, and (c) examine clarifier behavior under potential operating scenarios during the design phase. Jenkins et al. (2003) present a detailed illustration of the practical applicability of the Daigger operating chart.

*THE KEINATH APPROACH.* Keinath (1990) sought the broader database of Wahlberg and Keinath (1988) to develop a design and operating chart presented in Figure 4.21. The database included information from 21 full-scale plants that varied considerably with respect to size, geographic location, mode of operation, method of aeration, and type and amount of industrial wastewater input. None of the sludges tested were chemically amended.

Results obtained using the Keinath operating charts differ substantially from the Daigger approach discussed above, especially at high SVI values. Much of this difference can be attributed to the effect of stirring during the SVI test. For a single mixed liquor tested by Wahlberg and Keinath (1988), a stirred SVI of 122 mL/g was measured in contrast to an unstirred value of 189 mL/g. Most full-scale plant SVI data are based on unstirred test results. For such plants, sufficient stirred test data are needed to successfully use the Keinath nomograph, or a correlation between stirred and unstirred test data must be developed. Daigger (1995) developed such a correlation but good correlation is neither transferable from plant to plant, nor over a wide range of MLSS concentrations.

Keinath (1990) outlined the use of the design and operating chart (Figure 4.21) for designing secondary clarifiers according to the thickening criterion and evaluating various economic tradeoffs to determine a cost-effective design. He also presented examples to demonstrate the effect of corrective strategies such as RAS control or conversion to step-feed on ameliorating thickening overload conditions in an operating secondary clarifier. For example, a plant with an MLSS concentration of 2 g/L, a flowrate of 4000 $m^3/d$, a 50% RAS pumping rate, and a stirred SVI of 125 mL/g led to the prediction of a 6-g/L RAS concentration and a clarifier limiting SLR of 90 $kg/m^2 \cdot d$. This requires a clarifier surface area of 133 $m^2$. A higher MLSS concentration would lead to a larger clarifier area but smaller aeration tank volume if SLR were the governing criterion for tank sizing. The nomograph permits clarifier areas to be easily determined for various sets of conditions so that the most optimum conditions can be found.

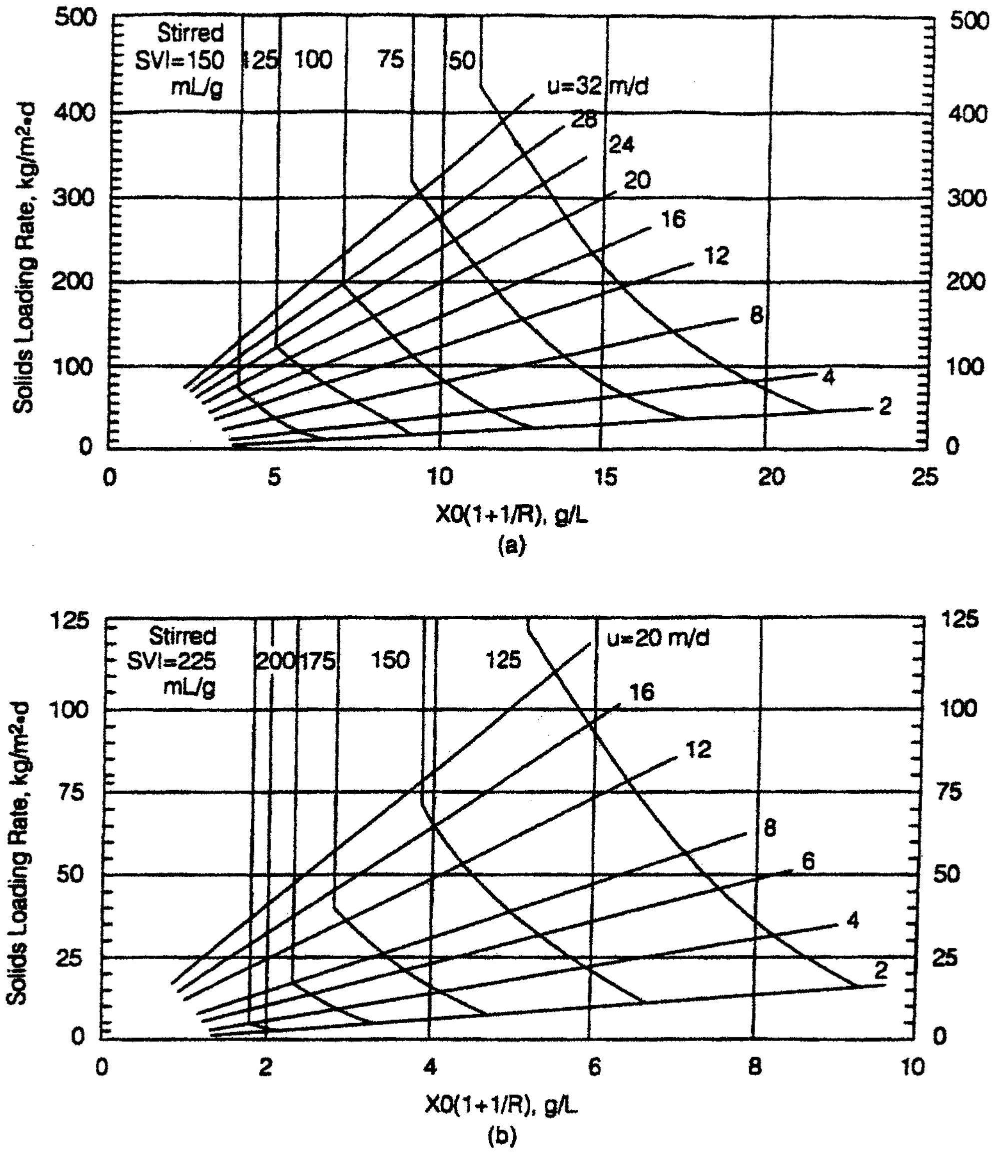



**FIGURE 4.21** Keinath operating chart (Keinath, 1990).

*THE WILSON APPROACH.* Wilson (1996) presented a simplified method of evaluating secondary clarifier performance using the settled sludge volume (SSV or $V_{30}$) from a 30-minute settling test, which is routinely conducted by plant operators. He proposed the SSV as a good surrogate for ISV (which also represents the clarifier surface SOR), providing it is adjusted, where appropriate, for temperature, volatile solids content, and chemical addition. The following equations derived by Wilson allow engineers and operators to determine whether a clarifier is overloaded.

$$R_{\min} = \mathrm{SSV}/(10^3 - \mathrm{SSV}) \tag{4.14}$$

$$\mathrm{ISV} = V_o{*}\exp(-4{*}\mathrm{SSV}/10^3) \tag{4.15}$$

Where

$R_{\min}$ = minimum RAS rate (%),
ISV = initial settling velocity (m/h),
SSV = 30-minute settled volume (mL/L), and
$V_o$ = sludge-settling characteristic (m/h).

Figure 4.22 presents a family of curves relating ISV (or clarifier SOR) to SSV for various values of $V_0$, assuming $V_0$ (in m/h) is 0.3 to 0.5 times temperature in degrees Celsius. Wilson concluded that the model compares well with the empirically validated German Abwassertechnische Vereinigung (ATV) approach as well as the model developed by Daigger (1995).

The Wilson approach entails determining ISV, which is also the maximum surface overflow rate ($\mathrm{SOR}_{\max}$), from Figure 4.22 or eq 4.15 and $R_{\min}$ from eq 4.14. These values are then compared with SOR and RAS rates determined from plant operating data. Finally, the CSF and return safety factor (RSF) are calculated as follows:

$$\mathrm{RSF} = \text{Plant RAS rate}/R_{\min}$$

$$\mathrm{CSF} = \mathrm{SOR}_{\max}/\text{plant SOR}$$

A CSF value of less than 1.0 indicates clarifier overload. If CSF and RSF are both greater than 1.0, the clarifier is underloaded. If CSF is more than 1.0 and RSF is less than 1.0, the clarifier is most likely overloaded and the operating condition should be confirmed using other methods, such as the Daigger approach.

*THE EKAMA–MARAIS APPROACH.* Ekama et al. (1997) characterized final clarifier behavior based on solids loading limited by (1) the solids flux (criterion I) and (2) the surface overflow rate (criterion II). This is illustrated in Chapter 6.

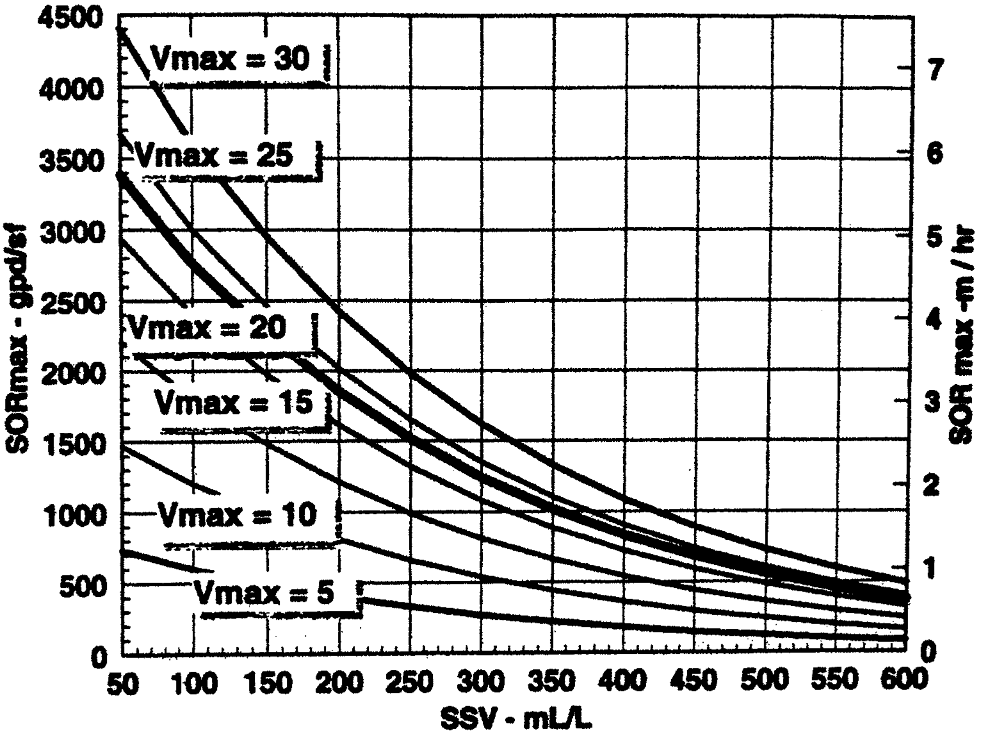



FIGURE 4.22 Wilson model (Wilson, 1996).

**DESIGN PARAMETERS OF IMPORTANCE.** Design of clarifiers typically requires specification of acceptable values for the following design parameters: SOR, applied solids flux, side water depth, and weir loading. Because of the light, fluffy nature of biological sludge, it is also important for the designer to have some idea of the expected degree of flow variation. Factors that affect clarification efficiency include aeration basin MLSS concentration, clarifier depth, recycle rate, and SOR.

***Solids Loading Rate.*** Establishing the maximum allowable SLR is of primary importance to ensure that the clarifier will function adequately. Most design engineers prefer to keep the maximum SLR (including full RAS capacity) in the range of 100 to 150 kg/m$^2$·d (20 to 30 lb/d/sq ft). Rates of 240 kg/m$^2$·d (50 lb/d/sq ft) or more have been observed in some well-operating plants with low SVI, well-designed clarifiers, and effective solids removal. Approaches to determining the limiting SLR are presented above.

***Overflow Rate.*** Overflow rates (SOR) used by design engineers, based on average dry weather flow (ADWF) and full-floor area, have been observed to vary from 0.5 to 2 $m^3/m^2 \cdot h$ (300 to 1 000 gpd/sq ft). Some plants are known to operate without difficulty at the upper end of this range and produce a high-quality effluent. In many documented cases, diurnal or maximum pumping peak rates of 2.7 to 3.1 $m^3/m^2 \cdot h$ (1600 to 1800 gpd/sq ft) do not exceed a secondary clarifier's capacity. In other cases, however, poor clarification efficiency is encountered at lower average and peak SORs.

A survey of consulting firms resulted in preferred SORs, shown in Table 4.10 (WEF, 1998). Randall et al. (1992) recommend average and maximum SORs based on the clear water zone , which is the free settling zone above the maximum height of the sludge blanket. Their recommendations, presented in Table 4.11, show peak criteria to be three times the average, which may not apply in many cases.

These capacity ratings were developed from clarifier designs in operation before 1970. Improvements in the design of inlet and outlet structures, sludge collectors, and sludge removal will increase allowable rates. It is projected that fully optimized clarifier designs will have 15 to 20% higher hydraulic capacity than the pre-1970 clarifier designs having the same side water depth (WEF, 1998)

A correlation between effluent suspended solids and SOR developed for several plants, shown in Figure 4.23, indicates that an effluent total suspended solids (TSS) of less than 20 mg/L can be achieved at SORs ranging from 1.0 to 2.0 m/h. Such cor-

**Table 4.10** Preferred overflow rates ($m^3/m^2 \cdot h$ [gpd/sq ft]) (WEF, 1998).

| | Circular clarifiers | | Rectangular clarifiers | |
|---|---|---|---|---|
| **Flow** | **Range** | **Average** | **Range** | **Average** |
| Average | 0.68–1.19 | 0.95 | 0.68–1.19 | 0.95 |
| | (400–700) | (560) | (400–700) | (560) |
| Peak | 1.70–2.72 | 2.09 | 1.70–2.72 | 2.10 |
| | (1000-16 000) | (1230)[a] | (1000–16 000) | (1240)[b] |

[a]10 of 15 firms use 2.04 $m^3/m^2 \cdot h$ (1200 gpd/sq ft).
[b]8 of 13 firms use 2.04 $m^3/m^2 \cdot h$ (1200 gpd/sq ft).

**TABLE 4.11** Clarifier overflow rate limitations (Randall et al., 1992).

| Hydraulic condition | Moderate CWZ* 1.83–3.05 m | Deep CWZ 3.05–4.57 m |
|---|---|---|
| Average SOR ($m^3/m^2 \cdot h$) | 0.091 CWZ | 0.182 CWZ |
| Maximum SOR ($m^3/m^2 \cdot h$) | 0.278 CWZ | 0.556 CWZ |

*CWZ = clear water zone.

relations can be misleading because they do not account for the effects of temperature, peaking factors, SVIs, geometrical details, RAS flowrate, and RAS concentration. Because the literature is limited in this area, designs for specific sites should be conservative or based on experimental testing (Tekippe and Bender, 1987). Unbalanced load testing at existing plants undergoing expansion is encouraged. If such testing is not feasible, bench-scale investigations should be undertaken to provide reasonable design criteria.

***Side Water Depth.*** Selection of side water depth is based on the size of the unit or the type of biological process preceding it. The general trend in design practice is to make circular clarifiers deeper. Recommended values range from 2.4 to 4.6 m (8 to 15 ft). The distance of the sludge blanket from the effluent weir has a direct relationship

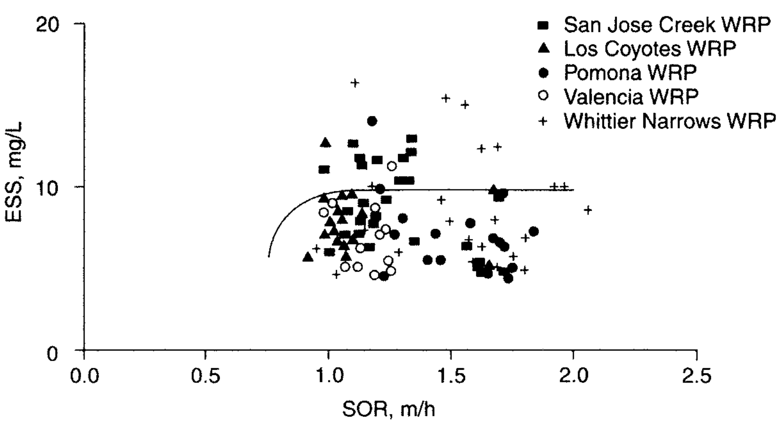



**FIGURE 4.23** Effect of SOR on effluent suspended solids (ESS) (Stahl and Chen, 1996).

to effluent quality (Miller and Miller, 1978). Based on historical operating data, Parker (1983) has demonstrated the effect of depth on effluent quality. At similar SORs, the average concentration of suspended solids in the effluent from a settler decreased as depth increased. Variability in effluent quality also decreased with increasing depth. In the ATV standards, the tank depth is calculated from four functional depths: (1) clear water zone, (2) separation zone, (3) sludge storage zone, and (4) thickening and sludge-removal zone. The side water depth (SWD) determined by this method is typically more than 4 m (13 ft). In a survey of 20 consulting engineering firms specializing in U.S. waste treatment plant design, Tekippe (1984) found the depth used for most large activated sludge secondary clarifiers ranged from 4 to 5 m (12 to 15 ft). In the final analysis, the decision to increase clarifier depth will be based, in large part, on the cost versus the anticipated improvement in effluent quality. Some form of economic analysis may be necessary to reach a decision. Additional discussion on clarifier depth may be found in Chapter 8.

***Weir Loading.*** The present consensus is that weir placement and configuration have greater effects on a clarifier's performance than weir loading, particularly in the absence of excessive sludge blanket depths and high flow energies near the weirs. However, misaligned weirs can cause flow imbalance within clarifiers. If upstream flow splitting is not proper, misaligned weirs can also interfere with flow distribution between clarifiers. Many state regulations limit maximum allowable weir loadings to 120 $m^3/m \cdot d$ (10 000 gpd/ft) for small treatment plants (less than 4000 $m^3/d$ [1 mgd]) and 190 $m^3/m \cdot d$ (15 000 gpd/ft) for larger plants. Experience of many operators and design engineers has led to a general agreement that substantially higher weir loading rates would not impair performance, provided other design parameters are selected consistent with good design practice.

For radial-flow (circular or square) clarifiers, a single peripheral weir is typically considered adequate, especially if some baffling is provided to prevent an updraft wall effect that results in TSS approaching the weir. Other engineers prefer to handle this problem by locating double inboard launders at a distance of approximately 30% of the tank radius from the outer wall. The double launder concept increases construction cost but, as demonstrated by Anderson (1945), improves performance over that of simple peripheral weirs without baffling.

For rectangular tanks, launders that extend 25 to 30% of the tank length from the effluent end and are spaced at approximately 3-m (10-ft) intervals have worked well. Some engineers continue to believe that a simple full-width weir at the effluent end is sufficient. Regardless, providing extensive launder structures to meet arbitrary criteria of 120 to 190 $m^3/m \cdot d$ (10 000 to 15 000 gpd/ft) seems unwarranted unless necessary to meet certain state criteria.

Algae growth is a problem with many clarifiers having weirs and open troughs. Strategies that have been found to be effective in minimizing algae growth include installing trough covers, feeding chlorine solution, hydraulic spray washing, and mounting algae brushes on rotating mechanism.

**HYDRAULIC CONSIDERATIONS.** ***Internal and External Factors.*** Hydraulic issues are pivotal to the performance of clarifiers. These include issues external to the clarifiers such as flow distribution and the internal hydraulic behavior of the units.

Equal flow distribution to the operating clarifiers is essential for ensuring uniform performance among the units. To achieve consistent and reliable operation, the design of the flow distribution system should be such that proper feed distribution is achieved under the range of expected flow conditions with one clarifier out of service. In addition, turbulence should be minimized in flow distribution structures to prevent floc breakup.

Internal tank hydraulics is more complex. They influence the following, which are linked to clarifier performance:

- Extent of flocculation,
- Energy dissipation,
- Density currents,
- Uniformity of effluent flow,
- Extent of short-circuiting, and
- Resuspension of settled sludge.

***Effect of Flow Variation.*** Generally, clarifier area is selected based on average and peak flows. Though such a procedure can produce an extremely conservative design in some cases, it is considered necessary because little is known regarding the mechanisms by which flowrate variation affect clarifier efficiency, except for a few installations or in extreme cases, and generalized quantitative relationships are not available.

According to Collins (1979), horizontal transport of solids away from the clarifier inlet is a direct function of both amplitude and frequency of flow variation. He reports that transients created by intermittent pump operation are damaging to effluent quality. Porta (1980) reported that implementation of measures to control surges created by influent pumping eliminated the need for an additional four clarifiers. Chapman (1983), investigating small-diameter clarifiers, noted that the practice of controlling the clarifier recycle at a constant proportion of the plant inflow magnifies influent transients. Based on a study in Phoenix, Arizona, Wilson (1983) found that failure of final clarifiers did not occur as long as the average daily SOR did not exceed the settling velocity of the mixed liquor solids. A U.S. EPA report (1979) on activated sludge clarification suggests that the effect of flow peaks is small until a threshold value of approximately 41 $m^3/m^2 \cdot d$ (1000 gpd/sq ft) is reached. In some cases, deterioration in effluent quality lagged flowrate variation considerably.

Chapman (1984) and Dietz and Keinath (1984) have attempted to characterize the time varying response of a settler to step changes in feed flowrate. Both studies found that settler response to a step increase in flowrate was rapid with process time constants of approximately 30 minutes. Chapman observed an initial overshooting of final steady-state values in some experiments. Step decreases in the feed flowrate resulted in the concentration and variability of the effluent suspended solids decaying exponentially to a new steady-state value; however, the response time was longer than for step increases. Chapman (1984) found that changes in the MLSS concentration in the feed resulted in changes in the effluent suspended solids concentration, but the time constant for the response was long—approximately 5 hours.

Both Chapman (1984) and Dietz and Keinath (1984) recommend that treatment plants take steps such as equalization, system storage, and careful pump selection to control influent surges.

***Flow Regimes.*** Hydraulic regimes for reactors used in wastewater treatment plants are typically classified as plug-flow, complete-mix, or arbitrary flow.

Plug-flow reactors convey liquid through the tank as a plug without longitudinal mixing. Every particle is assumed to remain in the tank for an amount of time ($t$) equal to $V/Q$, where $V$ and $Q$ are the tank volume and flowrate, respectively. Complete-mix reactors provide complete and instantaneous feed mixing. Retention time distribution (RTD) in a complete-mix tank may range from near zero to infinity, with an average value of $V/Q$. In practice, true plug-flow or complete-mix conditions can be approached but never achieved. Arbitrary flow reactors provide a degree of mixing that places them somewhere between plug-flow and complete-mix as far as RTD; all clarifiers, in fact, fall into this category. All three regimes are often characterized by dispersion curves produced by a slug or continuous input of dye or salt to

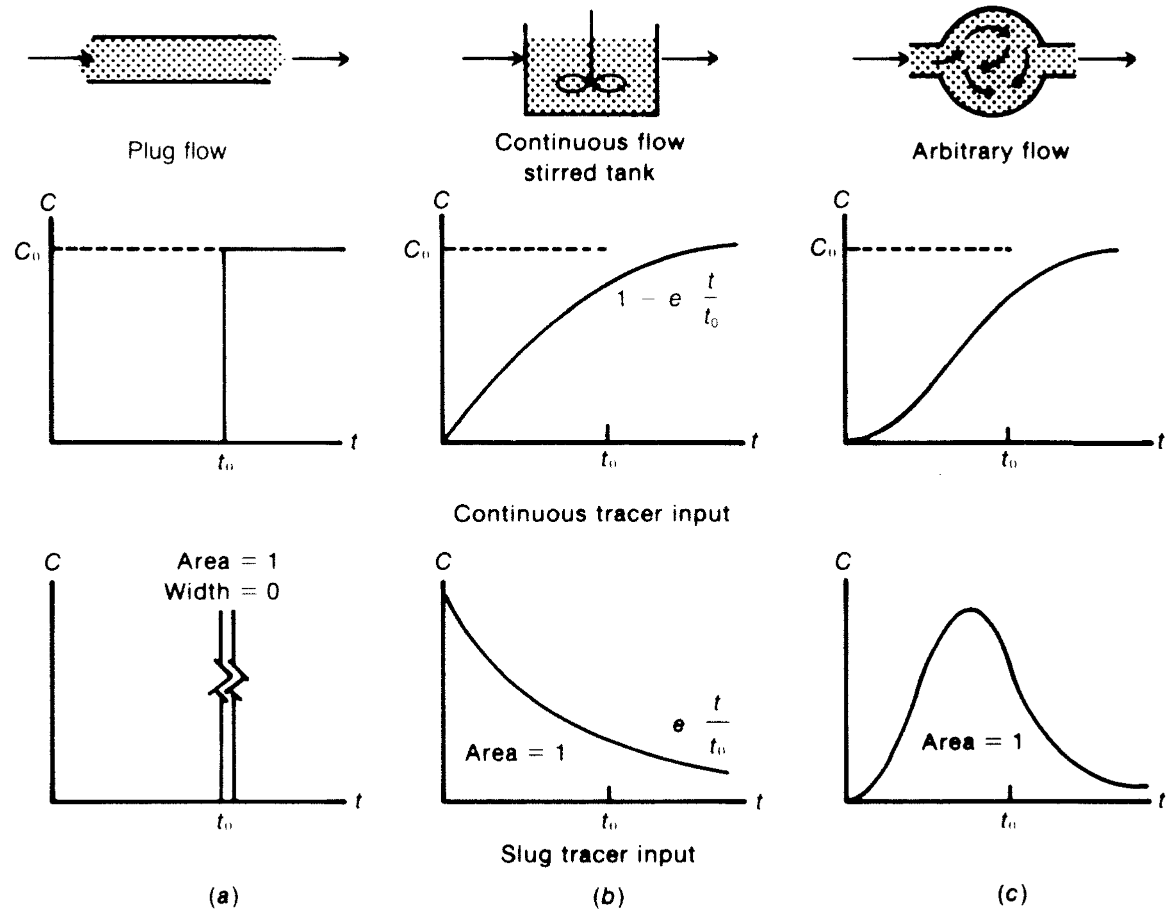



**FIGURE 4.24** Tracer response curves: (a) plug-flow, (b) complete-mix, and (c) arbitrary flow.

the feed. The curves in Figure 4.24 would be obtained by measuring tracer concentrations in the tank effluent.

Wide variation in clarifier efficiency has been observed, even in units of similar design. This results partly from making unverified, simplifying assumptions such as plug-flow hydraulic regime and uniform SOR. As a result, it has become increasingly important to have an understanding of the flow pattern in the tank and its relationship to the efficiency of biochemical oxygen demand (BOD) and solids removal.

Early research centered on characterizing the hydraulic regime in clarifiers using dispersion curves that provide some idea of the RTD in the tank (Reynolds and

Richards, 1996). It was typically assumed that efficiency would improve as the regime approached plug-flow. This approach has drawn criticism from recent researchers because the findings are based primarily on measurements made only at the effluent end of basins, with no effort made to study the conditions within the basin. Investigators have questioned the accuracy of plug-flow assumptions in clarifiers, indicating that plug-flow constitutes less than 40% of the effective flow area (not including dead zones), whereas mixing areas constitute more than 60%.

Hall (1966) states that the hydraulic regime that will achieve maximum solids removal is not the classical plug-flow but one with controlled turbulence and mixing that encourages flocculation. Tebutt (1969) has made a similar suggestion. Clements and Khattab (1968) have shown that velocity variations across the horizontal dimensions perpendicular to the flow seriously affect sedimentation efficiency for both circular and rectangular basins. Velocity variations with depth have little effect on sedimentation, provided scour is avoided.

Crosby and Bender (1980) and Bender and Crosby (1984) indicated that dye-dispersion tests give good indication of fluid movements within a clarifier. They developed several test procedures, which were later incorporated into an American Society of Civil Engineers (ASCE)/CRTC clarifier testing protocol that can provide insight to fluid behavior and solids distribution in a clarifier. The "flow pattern/solids distribution" procedure allows the analyst to produce a snapshot of the solids distribution at a particular cross section at a particular time. The "weir-wall solids" procedure provides information regarding direction preferences at the effluent weir and was used to determine the effect of sludge-removal mechanisms on clarifier performance. The following are some of the conclusions of the study.

The means for controlling hydraulic balance between clarifiers is often inadequate.

Balance between inlet ports on an individual clarifier is sometimes poor. Density currents are real, longitudinally persistent, and detrimental to effluent quality. Influent baffling fails to intercept these jets in some cases. Sludges that settle rapidly seem to produce higher velocity density currents, higher turbulence, and higher effluent turbidity than slower settling sludges.

Albertson (1992) found that, whereas detention efficiency does not govern clarification efficiency, it would limit clarifier SLR. According to Ekama et al. (1997), to maximize the detention efficiency, clarifier design must minimize energy gradients at the influent and effluent, control density currents, maximize the cross-sectional use of the basin, and prevent sludge blanket from encroaching on the clarification volume. Full-scale studies (ASCE/CRTC) have revealed that nonideal flow behavior was strongly linked to clarifier SLR, which created density currents and reduced the available clarification volume.

Ekama et al. (1997) defined two modes of short-circuiting in final clarifiers—the feed solids and liquid prematurely reaching the underflow and effluent, respectively. Under ideal conditions, the "first-in–first-out" criteria would be satisfied for both the solids and liquid components of the feed. Because short-circuiting in clarifiers can only be minimized and not eliminated, the above criteria cannot be achieved but should remain a goal. In dye tests performed by Lively et al. (1968) in a center draw-off clarifier (34.1 m in diameter, 3.66 m deep), it was observed that the dye appeared in the underflow within 10 minutes and peaked in the overflow at 40% of the theoretical detention time of 2.5 hours. The clarifiers were operated at an SOR of 1.55 m/h and SLR of 3.9 $kg/m^2 \cdot h$.

Lumley and Horkeby (1988) conducted similar investigations in 60-m-long rectangular clarifiers and found that the solids retention time was 76 to 91% of the nominal retention time. The modal peak of tracer in the effluent occurred at 54 to 76% of the nominal retention time.

***Flow Control.*** When multiple clarifiers that operate in parallel are designed, it is essential to maintain accurate flow distribution at all times, but especially when one or more units are taken out of service. In most plants, parallel clarifies are of the same size, so equal flow distribution is sought.

When tanks are not equal in size, flows should be distributed in proportion to surface area. For circular tanks, separate flow-splitting structures or pipe symmetry are used. For rectangular tanks, the inlet gates on a common-feed channel are most often used. Positive flow-splitting structures (such as feeding symmetrical splitting weirs from an upflow chamber leaving no residual horizontal velocity components) are effective and are the preferable method. Without positive flow-splitting structures, flow measurement and feedback are necessary to ensure proper splitting and control. Open-channel flumes have been used successfully for such measurements.

The advantages of hydraulic controls are low maintenance and low initial cost. To obtain a reasonably proportioned hydraulic split, a significant head loss must be taken through the weir or orifice, typically in excess of 150 mm (0.5 ft) for small plants and more for larger ones. This may not always be cost effective when energy costs are calculated for the life of the project. A hydraulic split requiring head loss may be impossible when additional clarifiers are added in parallel to existing units where the head loss of the new units must use only what is available in the existing hydraulic profile.

The concept of using effluent weir elevations, with minimal tank inlet head loss, to control flow split is often grossly inadequate. Unequal head loss in the influent channel feeding parallel tank is common and leads to poor distribution. It is also necessary to adjust the weirs to account for settling that occurs through the life of the clarifiers. Finally, weir settings must be precise because clarifier weir loadings are kept low to minimize effects on settling efficiency.

Flow-measurement devices and flow-control valves have the advantage of minimum head loss and good accuracy. The selection of proper valve size for the range of flowrates anticipated is crucial to a successful operation. Because a flow-measuring instrument generates a control signal and a signal-controlled automatic valve operator is required, there are devices in the system that are much more complex than those involved in flow proportioning, and the resulting maintenance requirements are much higher. Where flow measurement and feedback to a motor-controlled valve are used, a dampening, delayed-response system is important to prevent “hunting” or cycling involving overcompensation of the valve operator. Such fluctuations, even of small magnitude, will establish an inlet surge phenomenon that is detrimental to quiescent settling. The continual maintenance of the instruments and controllers required with this method is a distinct disadvantage.

In extremely large units, an auxiliary control gate must be used to adjust inlet/outlet flows. This gate is smaller than the main (shutoff) gate and will give much better control because less drastic flow changes will result when the gate is repositioned a small amount than would be experienced with the same reposition of the main gate.

Proper valve selection is necessary to obtain control. Gate valves and slide gates are not suitable as control devices because of frequent clogging from solids. Additional discussion on flow distribution may be found in Chapters 8 and 9.

## CLARIFIER PERFORMANCE ENHANCEMENTS

**PROCESS CONFIGURATION.** In general, the operator can take positive steps to ensure good settleability. In low F/M systems, a good approach is to operate at least the first portion of the aeration tank in plug-flow configuration. This configuration can minimize the growth of low F/M types of organisms that result in a bulking

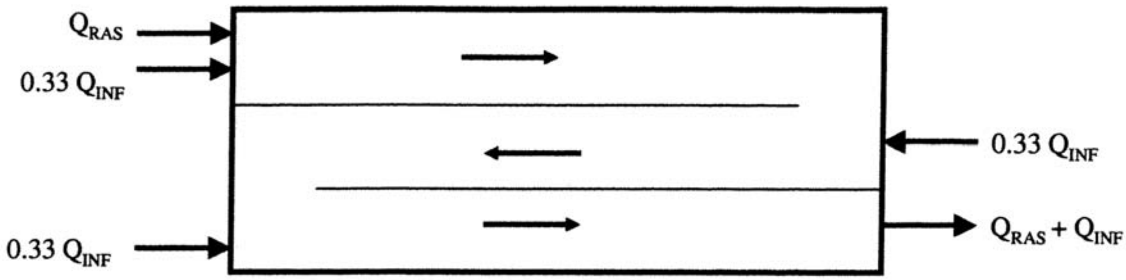



**FIGURE 4.25** Typical step-feed configuration.

sludge. An initial plug-flow zone provides a high F/M, which acts as a selector favoring floc-forming organisms when adequate dissolved oxygen is provided.

Providing step-feed capability so that some or all of the influent flow can be added at each of several points along the length of the aeration tank is sometimes recommended for operational flexibility. Typically, influent ($Q_{INF}$) is equally split between two to four addition points and return sludge ($Q_{RAS}$) is added only to the first pass of the aeration tank as shown in Figure 4.25. This type of design, for a given tank volume and F/M (or SRT), allows lower SLRs on the final clarifiers, thus allowing required treatment levels to be attained without affecting clarifier loading. Step feed also allows the oxygen demand to be more evenly distributed along the length of an aeration tank. Additional discussion on step-feed conversion may be found in Chapter 3.

**SELECTORS.** To promote the growth of floc-forming microorganisms while suppressing filamentous growth, special reactors called selector tanks can be provided ahead of conventional aeration basins. The goal is to maintain high enough F/M in the initial contact zone to achieve rapid soluble organic matter uptake rates. Although a single selector tank can be effective in controlling filaments, Jenkins et al. (2003) notes that that multiple compartments maintain plug-flow, enhance substrate gradient, and improve kinetic selection. A typical selector tank configuration is illustrated in Figure 4.26. Selectors can be aerobic, anoxic, or anaerobic. Table 4.12 summarizes the characteristics of the three types of selectors, the primary mechanisms, and design criteria. The reader is referred to Jenkins et al. (2003) and the Water Environment Federation (WEF, 1998) for detailed discussions on selector effects and design approaches.

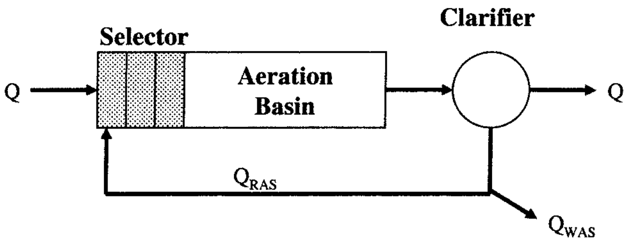



**Figure 4.26** Typical selector configuration (WAS = waste activated sludge).

**Table 4.12** Comparison of selectors.

| Type and characteristics | Features |
|---|---|
| **Aerobic** | |
| Environmental condition: | Advantages |
| Adequate dissolved oxygen | Simple process<br>No MLSS recycle required |
| Primary substrate removal mechanisms[a]:<br>• Storage<br>• Aerobic respiration | Disadvantages<br>No reduction in oxygen requirements<br>May need to redesign aeration system to meet high oxygen demand in the selector zone |
| | Design criteria[a]<br>Multiple selector compartments:<br>• Initial contact zone F/M = 10 to 12 kg COD[b]/kg MLSS·d<br>• Overall selector F/M = 3.0 to 4.0 kg COD/kg MLSS·d<br>• Dissolved oxygen: 1 to 2 mg/L |
| **Anoxic** | |
| Environmental condition: | Advantages |
| dissolved oxygen absent | • Alkalinity recovery |
| Adequate nitrate nitrogen | • Reduction in oxygen requirements as a result of aeration credit |
| Primary substrate removal mechanisms[a]: | • Can also achieve total nitrogen removal |

**TABLE 4.12** Comparison of selectors. *(continued)*

| | |
|---|---|
| • Storage<br>• Denitrification | Disadvantages<br>• Nitrification is a prerequisite<br>• MLSS recycle may be required if RAS denitrification is inadequate<br>• Tight control of recycle dissolved oxygen load and backmixing necessary to preserve the integrity of the anoxic zone and to prevent low dissolved oxygen bulking<br>• Mixers required |
| | Design criteria[a]<br>Multiple selector compartments:<br>• Initial contact zone F/M = 6 kg COD/kg MLSS·d<br>• Overall selector F/M = 1.5 kg COD/kg MLSS·d<br>Single selector basin:<br>• F/M ≤ 1.0 to 1.5 kg $BOD_5$/kg MLSS·d<br>• Anoxic SRT = 1 to 2 days |
| **Anaerobic** | |
| Environmental condition: dissolved oxygen and nitrate nitrogen absent | Advantages<br>• Simple design<br>• No MLSS recycle required<br>• Can also achieve biological phosphorus removal |
| Primary substrate removal mechanisms[a]:<br>• Polyhydroxyalkanote storage<br>• Hydrolysis of stored polyphosphate or fermentation of stored glycogen | Disadvantages<br>• Tight control of recycle dissolved oxygen and nitrate nitrogen loads and backmixing necessary to preserve the integrity of the anaerobic zone and prevent low dissolved oxygen bulking<br>• Mixers required<br>• No reduction in oxygen requirements |
| | Design Criteria[a]<br>Multiple selector compartments:<br>• Initial contact zone F/M = 6 kg COD/kg MLSS·d<br>• Overall selector F/M = 1.5 kg COD/kg MLSS·d<br>• Anaerobic HRT: 0.75 to 2.0 h |

[a]Jenkins et al. (2003).

[b]COD = carbonaceous oxygen demand.

Good design practice entails microbiological analyses to identify dominant organisms, initial determination of the viability of selector zones, pilot- and full-scale studies of proposed selector systems, and the design of selectors based on pilot- and full-scale test results. Examples of pilot- and full-scale investigations of selector system performance have been reported (Daigger et al., 1985, and Wheeler et al., 1984).

**FOAM CONTROL.** Foam formation typically occurs in the aeration basin and is conveyed to the secondary clarifiers with the mixed liquor. Foam accumulation on the liquid surface can lead to a deterioration of effluent quality. In addition, it is unsightly and a nuisance to operating and maintenance staff.

Foam control methods, described in detail by Jenkins et al. (2003), include

- Selectors (aerobic, anoxic, or anaerobic),
- Selective surface wasting from activated sludge basins,
- Surface chlorine spray,
- Cationic polymer addition to activated sludge basins, and
- Automatic mean cell residence time control using online MLSS and RAS solids concentrations.

**DISSOLVED OXYGEN AND FOOD-TO-MICROORGANISM RATIO.** The rate of $BOD_5$ removal in a plug-flow system requires supply of most of the air to the first portion of the aeration tank. If air addition does not match the oxygen demand profile, the dissolved oxygen concentration may drop below a critical value and bulking organisms may form. Incorporation of selector technology can reduce some of these problems more common in early years of the process. Figure 4.27 illustrates the importance of required dissolved oxygen as a function of loading and dissolved oxygen uptake rate in a continuously mixed system for controlling the growth of filaments (Jenkins et al., 2003). Matching oxygen demand to air supply is typically achieved by tapered aeration or step-feed operation.

**CHEMICAL ADDITION.** Chemicals may be added to enhance clarifier performance by eliminating excess filaments or inducing flocculation. Some bulking sludges can be controlled by RAS or sidestream chlorination. A typical design for a low (5- to 10-hour) hydraulic residence time (HRT) system uses 0.002 to 0.008 kg

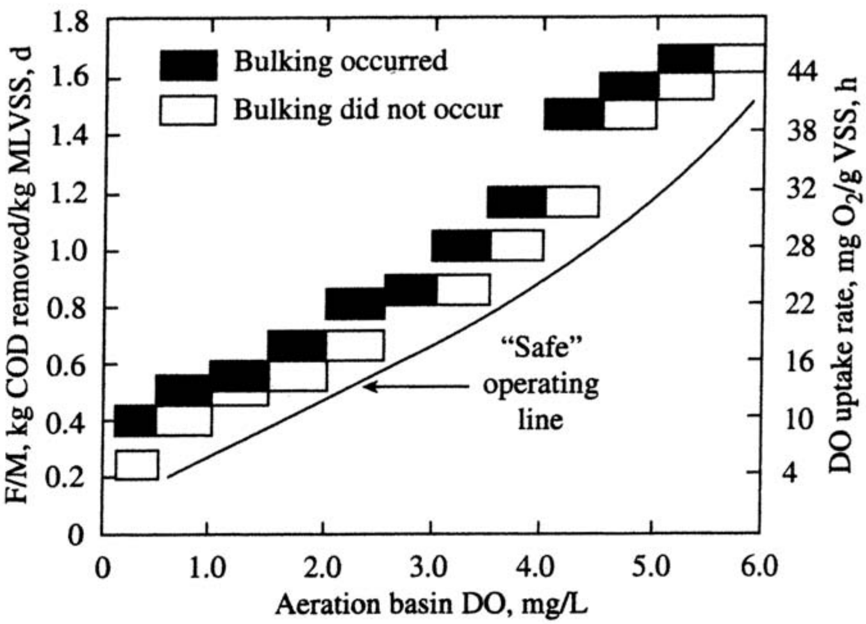



FIGURE 4.27 Bulking and nonbulking conditions in completely mixed aeration basins (COD = carbonaceous oxygen demand; DO = dissolved oxygen; MLVSS = mixed liquor volatile suspended solids) (Jenkins et al., 1993).

chlorine ($Cl_2$)/kg MLSS·d (2 to 8 lb $Cl_2$/d/1 000 lb MLSS), with the chlorine added to the RAS system. Longer HRT systems might need chlorine added to a sidestream or multiple points in the aeration tanks (Figure 4.28). Hydrogen peroxide can be substituted for chlorine in many cases. Further design and sizing details can be found elsewhere (Jenkins et al., 2003). Note that RAS chlorination can interfere with nitrification. One full-scale study (Ward et al., 1999) revealed that, to maintain biological nutrient removal (BNR) capability, the chlorine dose needs to be less than 0.001 kg $Cl_2$/kg MLSS·d (1 lb/d/1000 lb mixed liquor volatile suspended solids). The study also reported that, following chlorine inhibition, nitrification was established quicker than phosphorus removal when chlorine addition was ceased.

Chemical coagulants can be added to induce flocculation. For example, the addition of cationic polymers at concentrations of typically less than 1 mg/L has been shown to be effective in improving mixed liquor settleability. The selection of inorganic salts, polymers, or other flocculent aids should be based on laboratory studies.

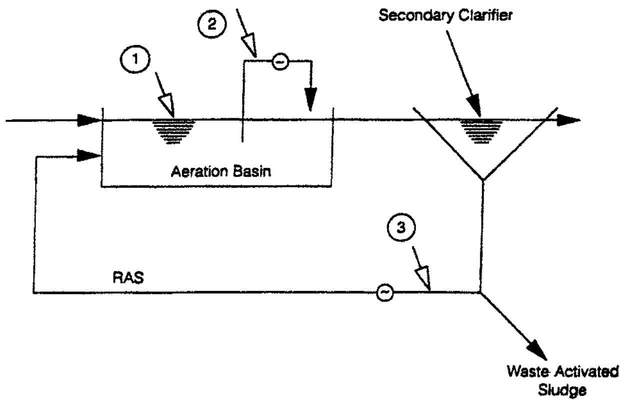



FIGURE 4.28 Chlorine dosing points for bulking control (Jenkins et al., 2003).

**HYDRAULIC IMPROVEMENTS.** Clarifier hydraulic performance is critical to good solids separation. Proper structural design and strategically placed devices as described below can significantly improve clarifier hydraulics. A few critical points are noted below. Additional material may be found elsewhere in this publication as noted.

Inlets must dissipate influent mixed liquor energy, distribute flow evenly in vertical or horizontal directions, reduce density short-circuiting and current effects, minimize blanket disturbances, and promote flocculation. Das et al. (1993) demonstrated that velocities in excess of 0.6 m/s (2 ft/sec) would cause deflocculation of biological flocs. If properly harnessed, the incoming energy can be used to promote flocculation, resulting in improved clarifier performance (Kalbkopf and Herter, 1984; Parker and Stenquist, 1986). The reader is referred to Chapter 8 for additional discussion.

Energy dissipating inlets and inlet diffusers promote reflocculation and provide uniform distribution of flow to the flocculating feed well. Haug et al. (1999) indicated that a specifically designed energy dissipating device called LA-EDI, enabled sustained SORs of 2.65 m/h (1558 gpd/sq ft) and a 3-hour peak rate of 2.89 m/h (1700 gpd/sq ft) to be maintained without degrading effluent quality. Chapter 8 provides a comprehensive discussion on inlet design.

A number of researchers have demonstrated that hydraulic performance and suspended solids removal can be improved by the strategic placement of baffle plates. This is further discussed in Chapters 8 and 9. Hydraulic modeling using computational fluid dynamics (CFD) and dye testing are commonly used to optimize the geometry and placement of interior baffles. The reader is referred to Chapters 6 and 7 for an in-depth discussion of CFD modeling,

Inclined plates or tubes significantly increase the allowable upflow velocity in a clarifier (based on horizontal tank area). They have also been installed upstream of the final clarifier to reduce the solids loading rate. Chapter 3 provides a detailed analysis of plate settlers.

**AERATION TANK SETTLING.** In aeration tank settling (ATS), solids settling is initiated in the final stages of a plug-flow aeration basin, thereby reducing solids loading to the final clarifiers. This strategy has been used successfully to minimize solids washout during wet weather conditions. The ATS concept is further discussed in Chapter 3.

## MISCELLANEOUS ITEMS

**SPECIAL CONSIDERATIONS WITH NUTRIENT REMOVAL SLUDGES.** Special care is required in designing clarifiers to handle sludges from nutrient removal facilities. The final sedimentation process plays a pivotal role in effluent nitrogen and phosphorus levels because of the following issues unique to BNR sludges:

- Certain operating conditions crucial to BNR operations appear to favor filamentous growth, resulting in poor settling sludge.
- Effluent solids from a BNR plant have relatively high phosphorus content in the range of 5 to 10% of volatile suspended solids on a dry weight basis (Randall et al., 1992). Consequently, good solids capture in the clarifiers becomes critical for achieving phosphorus compliance.
- Biological nutrient removal requires a minimum SRT to be maintained, which is typically in the 5- to 10-day range and several folds higher than the SRT required for BOD removal. Because of the higher MLSS requirement, settling of the BNR sludge is likely to be thickening limited whereas secondary sludge settling is typically clarification limited.
- Rapid sludge removal and control of the sludge blanket is important in BNR operations. Deep sludge blanket leads to anaerobic or anoxic conditions. The

former causes secondary phosphorus removal while the latter causes denitrification, which may lead to rising sludge. Appropriate RAS rates should be selected to minimize both the mass of nitrate recycled to the anaerobic zone and denitrification.

- Deep sludge blanket also creates conditions conducive to secondary phosphorus removal in secondary clarifiers. Wilson et al. (1990) observed better phosphorus removal when some nitrate was present in the effluent. The presence of nitrate eliminates anaerobic conditions, which triggers phosphorus release.

**CLARIFIERS FOLLOWING FIXED-FILM PROCESSES.** In clarifiers following fixed-film processes, type II settling is often the predominant mechanism, particularly if flocculation occurs as the particles settle. Metcalf and Eddy (2003) note that clarifiers designed for trickling filters should be similar to designs used for activated sludge process clarifiers with appropriate feedwell size and increased side water depth. They recommended SORs as a function of side water depth. For example, the average and maximum SORs at a side water depth of 4 m (13 ft) are approximately 1.2 and 2.2 m/h (680 and 1300 gpd/sq ft), respectively. At 5 m (16.5 ft) side water depth, the recommended average and maximum SORs are approximately 2.4 and 2.7 m/h (1417 and 1623 gpd/sq ft), respectively.

A survey of combined processes (Harrison et al., 1984) indicated that trickling filter/solids contact (TF/SC), roughing filter/activated sludge, and biofilter/activated sludge processes all had mean SVIs of less than 100 mL/g. In these systems, the high dissolved oxygen and organic loading conditions limit the proliferation of filamentous organisms (Harrison et al., 1984).

According to Parker and Bratby (2001), the dispersed solids generated by trickling filters can be bioflocculated in a solids contact tank. However, the floc remains fragile and susceptible to breakup because of head loss involved with its transfer to the final clarifiers. For this reason, it is often necessary to provide the clarifier with a flocculator center well to reflocculate the solids and enhance solids separation. Parker and Bratby (2001) also noted that rapid sludge removal should be provided to eliminate anaerobic conditions and loss of bioflocculation, which could potentially result in elevated effluent suspended solids. In addition, in nitrifying systems, deep sludge blanket could lead to denitrification and sludge flotation. Based on stress testing of TF/SC systems, Parker and Bratby (2001) reported that flocculator clarifiers could withstand surface SORs to 3.5 m/h (2050 gpd/sq ft) without failure. Parker et al. (1996) reviewed performance data of the Corvallis, Oregon, TF/SC system. The data revealed that low SVIs of 30 to 53 mL/g resulted in high SLRs of 184 to 364 kg/m$^2$·d; whereas, at poor sludge settleability, clarifier failure occurred at relatively low solids loadings.

For clarifiers following integrated fixed film activated sludge systems, Sen et al. (2000) recommends average surface SORs of 0.7 to 1 m/h (400 to 600 gpd/sq ft) for a side water depth of less than 4.3 m (14 ft), with a SLR of 98 to 146 kg/m$^2$·d (20 to 30 lb/d/sq ft). Under peak flow conditions, the SOR should not exceed 1.7 m/h (1000 gpd/sq ft). Deeper clarifiers with flocculator centerwell and baffles to prevent wall currents can tolerate SORs in excess of 1.7 m/h (1000 gpd/sq ft). The recommended range of clarifier hydraulic application rate for a moving bed biofilm reactor is 0.5 to 0.8 m/h (295 to 472 gpd/sq ft) (Metcalf and Eddy, 2003).

**INTERACTION WITH OTHER PROCESSES.** Pumping of mixed liquor to the final clarifiers should be avoided if possible. In most cases, the hydraulic profile can be designed to permit gravity flow between the aeration basin and final clarifiers. Large drops in hydraulic profile should also be avoided. If gravity flow is not possible, provisions should be made to pump with an absolute minimum of energy gradient to prevent floc shearing and breakup. Even after such precautions, some breakup should be expected and clarification systems that promote floc reformation and growth should be selected. A gently aerated feed channel or clarifiers with flocculating feed wells are uniquely suited to enhance floc formation.

In activated sludge basins using high-energy aeration or mixing, careful design of turbines, jets, and surface aerators can avoid floc breakup. Wahlberg et al. (1994) observed that, when surface aerators are located near the aeration basin discharge, the clarity of basin effluent was poor because of floc breakup. In contrast, for oxidation ditches where mechanical aerators are typically located away from the discharge weir, the sheared floc is able to reflocculate, resulting in low turbidities. According to Grady et al. (1999), diffused aeration systems delivering more than 90 m$^3$ air/min/1000 m$^3$ tank volume is likely to cause floc shear. In the case of mechanical aerators, the authors indicated that the energy input should be limited to less than 60 kW/1000 m$^3$ to avoid floc shear.

As detailed by Keinath et al. (1977), the sizing and perhaps type of final clarifier selected can be greatly influenced by the size of the aeration basin. A higher MLSS concentration requires a smaller aeration basin and a larger clarifier surface area. As discussed in Chapter 3, in flocculent suspensions, the settling velocity decreases with increasing MLSS concentration.

The flow configuration implemented in the upstream aeration basin affects clarifier loading. In step-feed arrangements, the clarifier "sees" a lower MLSS concentration compared with a conventional system operated at the same SRT. Consequently, clarifier solids loading is reduced significantly.

Effective removal of screenings from influent flow is critical for ensuring trouble-free clarifier operation. If influent screens are inadequate or if shredders are used in lieu of screens, the unscreened or shredded material can reform into balls because of turbulence in the aeration basin and clog sludge removal systems and sludge pumps.

**COST OPTIMIZATION.** Sludge settleability is central to the selection of a design MLSS concentration, which determines to a large extent the relative split of tank volumes between the aeration basin and the final settler. As the design MLSS concentration increases, the size and cost for the aeration basin decreases while the cost of the settler increases. The combined cost of aeration basin and clarifier can be expressed as a function of the MLSS concentration. The optimized reactor MLSS is one for which the total cost is a minimum. Ekama et al. (1997) point out that this minimum cost increases for (1) unsettled wastewater, (2) higher influent wastewater strengths, and (3) longer SRTs. In addition, these operating conditions increase the size of the biological reactor relative to that of the clarifier and decrease the reactor size for higher wet weather flow peaking factors and poorer sludge settleabilities.

The data of Keinath et al. (1977) suggest that the total annual cost of an activated sludge/clarifier system is particularly sensitive to low HRTs (fewer than 6 hours), primarily because of clarifier and recycle pumping costs. Tantoolavest et al. (1980) concluded that least-cost activated sludge designs should call for low MLSS concentrations.

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# Chapter 5

# Tertiary Clarifier Design Concepts and Considerations


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# INTRODUCTION

Tertiary clarification is a unit process that can be used after conventional biological treatment to provide effluent water quality that is better than secondary standards. Common applications for tertiary clarification are enhanced removal of phosphorus, suspended solids, metals, and pathogens. Information is presented in this chapter on the scientific basis for tertiary clarification processes, including characterization of suspended solids, settling velocities and overflow rates, chemical coagulation, precipitation of metals, and chemical phosphorus removal. A final section presents information on selected examples of existing facilities using tertiary clarifiers. Material from a large number of sources has been summarized and referenced to provide the practicing engineer detailed information to support the design of tertiary clarification processes on a rational basis.

**HISTORICAL BACKGROUND.** Where or when the first tertiary clarifier was designed and constructed is not known with great certainty; however, it is reasonable to suspect that this occurred during the beginning of the 1960s. During this period, the first steps were taken to limit the input of nutrients to surface waters to control eutrophication. Initially, tertiary clarifiers took one of two forms. The first was the construction of tertiary clarification facilities after the activated sludge process to provide for the chemical precipitation of phosphorus. The second was in the construction of three-sludge processes, wherein denitrification was provided in a third activated sludge process following one dedicated to carbon oxidation and one for nitrification. Clarifiers for two- and three-sludge processes are considered to be a form of secondary sedimentation associated with an activated sludge process and will not be considered further in this chapter.

**CURRENT AND FUTURE USES.** Tertiary clarification is not widely used, but it does have a place in certain advanced wastewater treatment applications, including phosphorus removal, metals removal, pathogen (bacteria and virus) inactivation and removal, and membrane pretreatment.

***Phosphorus Removal.*** In areas of the country with phosphorus-limited surface water bodies, the trend in permit limits for effluent phosphorus concentrations has been decidedly downwards. Examples of this include Lake Onondaga, New York; the Florida Everglades; and Lake Mead, Nevada, where limits of 0.1 mg/L, 0.01 mg/L, and 0.01 mg/L, respectively, are in place. Such limits are difficult to meet with chemical addition before (preprecipitation) or to an activated sludge process (simultaneous precipitation) as these limits are near or below the nutritional limits required for biomass growth. Thus, it is necessary to take the phosphorus concentration down to very low levels after the biological process (post-precipitation). Tertiary phosphorus precipitation to very low concentrations should also follow any type of biological filter such as denitrification filters or upflow anoxic submerged packed bed reactors for the same reason, unless supplemental phosphorus is added before the denitrification reactor.

***Metals Removal.*** For some receiving waters, regulatory agencies have proposed in-stream water quality standards for selected metals at very low concentrations. In effluent-dominated streams, this can require treatment for the removal of metals. One method for removing many metals is chemical precipitation. Thus, tertiary clarifiers can be an important component of tertiary treatment processes for metals removal.

***Pathogen Removal.*** For indirect potable reuse applications, the concept of multiple barriers is often used to establish the degree of reliability and redundancy provided by the treatment process for the removal of pathogens, particularly bacteria, virus, and protozoan cysts. Because conventional primary and biological treatment only provides limited pathogen removal, most wastewater facilities rely primarily on the disinfection process for the destruction of pathogens. Higher log removals of pathogens can be provided by high pH lime clarification. This application of tertiary clarifiers has been demonstrated to provide approximately 1.3 log removal of virus by coagulation and sedimentation (Dryden et al., 1979). Investigations at the Upper Occoquan Water Reclamation Plant, Centreville, Virginia, demonstrated that lime clarification provides significant removal of all pathogens. Selected data from these two studies are presented in Tables 5.1 and 5.2.

Removal of metals and viruses by high-pH lime coagulation, alum coagulation, and high-pH lime treatment with recarbonation has been evaluated and compared in a pilot-plant study (Esmond et al., 1980). Reported results are summarized in Table 5.3.

***Membrane Pretreatment.*** The first generation of wastewater reclamation plants using reverse osmosis (RO) membranes for tertiary treatment relied on lime clarification and granular media filters for membrane pretreatment. For most applications, clarification and granular media filtration has been replaced by microfiltration or ultrafiltration pretreatment. However, lime clarification still has a few advantages for

**Table 5.1** Pathogen concentrations before and after lime treatment at the Upper Occoquan WRP (6 samples) (Rose et al., 1996).

| Parameter* | Secondary effluent | Tertiary lime clarification |
|---|---|---|
| Clostridium (CFU/100 mL) | | |
| Maximum | 1400 | 7.0 |
| Average | 640 | 2.6 |
| Enterococci (CFU/100 mL) | | |
| Maximum | 2700 | 48 |
| Average | 830 | 15 |
| Fecal coliforms (CFU/100mL) | | |
| Maximum | 12 000 | 10 |
| Average | 9900 | 5.2 |
| Coliphage (PFU/100 mL) | | |
| Maximum | 54 000 | 25 |
| Average | 12 000 | 1.6 |

*CFU = colony forming unit and PFU = plaque forming unit.

**Table 5.2** Average pathogen and indicator concentrations before and after lime treatment (12 samples) (Riley et al., 1996).

| Parameter | Secondary effluent | Tertiary lime clarification |
|---|---|---|
| Enterococci (No./100 mL) | 2200 | 15.2 |
| Total Coliforms (No./100 mL) | 110,000 | 43.9 |
| Fecal Coliforms (No./100 mL) | 8,000 | 9.2 |
| Clostridium (No./100 mL) | 4,900 | 4.2 |
| Coliphage—direct (No./100 mL) | 1,800 | 18.2 |
| Coliphage—indirect (No./100 mL) | 114.7 | 0.7 |
| Enterovirus (No./100 mL) | 24.1 | 0.2 |
| Cryptosporidium (No./100 mL) | <310.5 | 9.1 |
| Giardia (No./100 mL) | 1,700 | 32.1 |

high-pressure membrane pretreatment in specific applications. These include situations that can benefit from the following capabilities of lime clarification:

- Ability to remove significant amounts of silica and some sparingly soluble salts that will scale RO membranes in high recovery applications.
- Provision of an additional pathogen barrier.
- Removal of metals. Although most metals will be adequately removed by RO membranes, the metals are concentrated in the brine stream where they may still be objectionable. Lime clarification, however, precipitates the metals as insoluble compounds where they may be permanently bound.

## BASICS—THE SCIENCE OF DESIGN

Biological treatment processes are relied on for the removal of the majority of soluble pollutants in wastewater, including biodegradable carbon, ammonia, and nitrate. In contrast, tertiary clarification processes are primarily intended to remove pollutants and pathogens associated with suspended particles. This implies that the size and

**Table 5.3** Metals and coliform removal—mean parameter concentrations after treatment with lime and alum (Esmond et al., 1980).

| | High-pH lime coagulation | | High-pH lime with recarbonation | | Alum coagulation | |
|---|---|---|---|---|---|---|
| **Parameter*** | **Secondary effluent** | **Lime effluent** | **Secondary effluent** | **Recarbonation effluent** | **Secondary effluent** | **Alum effluent** |
| COD, mg/L | 91 | 27 | 65 | 27 | 57 | 28 |
| $BOD_5$, mg/L | 28 | 4 | 38 | 5 | 28 | 8 |
| TSS, mg/L | 27 | 46 | 44 | 33 | 28 | 17 |
| Organic nitrogen, mg/L | 4.9 | 2.3 | 4.43 | 2.21 | 3.85 | 2.39 |
| Total phosphorus, mg/L | 9.5 | 0.99 | 7.4 | 1.0 | 7.8 | 3.0 |
| Standard plate count, per 100 mL | $2.0 \times 10^5$ | 13 | $9.3 \times 10^3$ | $2.3 \times 10^3$ | $5.1 \times 10^4$ | $4.0 \times 10^3$ |
| Total coliforms per 100 mL | $1.4 \times 10^6$ | 3 | $3.2 \times 10^5$ | $2.3 \times 10^2$ | $4.2 \times 10^5$ | $2.3 \times 10^4$ |
| Fecal coliforms, per 100 mL | $8.0 \times 104$ | 2 | $2.5 \times 10^4$ | $1.7 \times 10^2$ | $3.1 \times 10^4$ | $1.8 \times 10^3$ |
| pH | 7.1 | 11.5 | 7.1 | 7.0 | 7.2 | 7.6 |
| Al, mg/L | 0.38 | 0.94 | 0.47 | 0.18 | 0.42 | 2.13 |
| As, μg/L | 19.5 | 10.8 | 20.6 | 5.6 | 10.1 | 6.8 |
| Ca, mg/L | 42.5 | 156.1 | 64.9 | 105.4 | 55.8 | 69.4 |
| Cd, μg/L | 7.6 | 7.6 | 7.3 | 3.6 | 5.6 | 4.1 |
| Cr, mg/L | 0.085 | 0.012 | 0.052 | 0.014 | 0.066 | 0.023 |
| Cu, mg/L | 0.080 | 0.080 | 0.077 | 0.050 | 0.054 | 0.034 |
| Fe, mg/L | 0.34 | 0.34 | 1.45 | 0.97 | 0.33 | 0.16 |
| Ni, mg/L | 0.082 | 0.082 | 0.106 | 0.087 | 0.079 | 0.068 |
| Pb, mg/L | 0.044 | 0.044 | 0.050 | 0.029 | 0.054 | 0.042 |
| Zn, mg/L | 0.123 | 0.123 | 0.133 | 0.084 | 0.156 | 0.109 |

*$BOD_5$ = 5-day biochemical oxygen demand; COD = carbonaceous oxygen demand; and TSS = total suspended solids.

other physical characteristics of the particles or floc in secondary effluent (such as electrokinetic surface potential, specific gravity, shape) are important variables in selecting a treatment process that will be effective in achieving desired water-quality objectives. This chapter is about sedimentation processes. Sedimentation is claimed to be an efficient process for solids–liquid separation when the suspended-solids mass concentration is large (> 50 mg/L) and when the particle size exceeds approximately 30 to 100 μm (Kavanaugh et al., 1980). Even though this is often not true with secondary effluents, sedimentation processes have still proved practical for the applications listed previously.

**PARTICLE CHARACTERIZATION.** Secondary effluent particles are composed of organic macromolecules, including humic substances, proteins, viruses, bacteria, and algae (Adin, 1998).

Humic substances have been reported to constitute a significant fraction of the residual organics in secondary effluents (Manka and Rebhun, 1982; Rebhun and Manka, 1971). A majority of the soluble organic matter in effluent is of microbial origin and is produced by microorganisms as they degrade the organic substrate in the influent (Grady et al., 1999). There are two sources for this material in secondary effluents—biodegradation of substrate and non-growth-related formation from the release of soluble cellular constituents through lysis and the solubilization of particulate cellular components (Grady et al., 1999).

There have been significant advances in the laboratory instruments used to measure particle sizes and other particle characteristics within the last decade. Patry and Takacs (1992) noted that, in 1992, there were very few studies (Roth, 1981; Roth and Pinow, 1981; Tchobanoglous and Eliassen, 1970) that attempted to measure particle size distribution (PSD) in secondary effluent. Analytical methods now exist to characterize wastewater effluent in much greater detail. Analytical techniques for measuring particle sizes include light blockage, light scattering, change of electrical resistance, microscopic analysis, membrane filtration, and field flow fractionation (Neis and Tiehm, 1997). In the past ten years, there has been significant research on particle size characterization as a result of the availability of these analytical methods. A summary of available methods for measuring particle sizes is given in the Table 5.4.

The original particle size analyzers were based on the Coulter principle, which measures the change in electrical signal produced by nonconductive particles in an electrolyte passing through a small orifice or aperture. The light obscuration method measures the time that it takes for a particle to pass through a laser beam. Light-scattering instruments are more exactly referred to by the description of the measurement they make. Multiangle light scattering (MALS) analyzers are based on the principle that the variation of scattered light intensity with angle may be used to derive an estimate of root mean square particle size. Quasi-elastic light scattering measurements are based on measurement of the particles' diffusion coefficients. From such determinations, instruments are able to derive the so-called hydrodynamic radius of the particles. The accessible range goes down to 1.0 nm and (depending on the experimental conditions) up to approximately a micrometer. Particle size distributions are derived by so-called regularization procedures to extract this information from the measured sample. Light-scattering sensors provide the ability to measure smaller sizes and provide narrower ranges compared to light-obscuration sensors. Multiangle light scattering instruments can measure molar mass, molecular radii, and diffusion coefficients in the range of 8 to 10 nm to larger than 1 μm. Light-blocking instruments have a lower limit of approximately 1.0 μm. Results from different types of instruments can not be directly compared. It is sometimes necessary to dilute effluent to get the particle count within an acceptable analytical range for a specific instrument. Sample water must be used as the dilution water as distilled water will alter the PSD (Adin et al., 1989).

Using MALS to determine particle size *distributions* (rather than average particle size) requires the initial separation of the particles in a sample by field flow fractionation or, if the particles are smaller than 100 nm, by using size-exclusion chromatography. For particles in the submicrometer range, the most accurate determination of size distributions may be achieved in this manner.

Particle size distributions in secondary effluent can typically be mathematically described by the following power law function (Adin, 1999; Adin et al., 1989; Alon and Adin, 1994):

$$\frac{dN}{d(d_p)} = A(d_p)^{-\beta} \tag{5.1}$$

Where

$N$ = number of particles in size interval,
$d_p$ = average particle size of interval (μm),
$A$ = empirical constant, and
$\beta$ = empirical constant.

**TABLE 5.4** Techniques for measuring aqueous PSD (Kavanaugh et al., 1980).

| Measuring principle | Equivalent size measured | Size limits: Minimum (μm) | Size limits: $R^a$ | Typical volume per scan (mL) | Sample handling | Time required for typical PSD in natural waters[b] |
|---|---|---|---|---|---|---|
| Electron microscopy | Statistical length | 0.001 | 50–200 | | Evaporation, carbon coating | 8–16 h |
| Optical microscopy | Statistical length | 0.3 | 40 | $1000\times-10^{-5}$<br>$100\times-10^{-2}$ | Concentration may be required | 4–8 h |
| Electrical sensing zone method (Coulter principle) | Volume diameter | 1 | 20 | 0.05–2 | Sample must contain 0.9% salt solution; scalping may be necessary | 45 min |
| Light scattering: forward-scattering laser light source | Root mean square radius | 0.008 | 10–50 | 10–1000 | Dilution to reduce coincidence errors | 15 min |
| Quasi elastic light scattering | Hydrodynamic radius | 0.001 | 50–1000 | 1 | Filtration to remove large outlying contributors | 5 min |
| Light obscuration | Cross section diameter | 1 | 50–60 | 5–100 | Dilution to reduce coincidence errors | 10 min |

[a]$R$ = ratio of maximum to minimum size for single sensing element, or single magnification.

[b]Speed depends on spread of distribution, number of size intervals required to characterize suspension; assumes PSD with range from 2 to 100 μm; includes sample preparation and instrument time.

Figure 5.1 contains graphs of the power law for β values from 1 to 5. In the power law distribution function, the exponent provides an estimate of particulate contribution by size to the total particulate number, surface area, volume, and light-scattering coefficient. For $\beta > 3$ the smaller size fractions dominate and, for most particulates in natural water and wastewater, the power law coefficient is typically greater than 3. For $\beta = 3$, the surface area concentration is equally distributed among all size intervals, the largest number of particles are in the small sizes (< 3 μm), and the largest volume of particles is in size intervals larger than approximately 10 μm (*5*). If $\beta > 3$ the total surface area concentration of the particulate fraction is predominantly in the finer particles and treatment should be directed at removing these fine particles. If $\beta < 3$, then removal of the larger particles might be sufficient to obtain a significant reduction in the mass concentration of solids, implying that tertiary sedimentation will be most effective for secondary effluents with these PSDs. Reported power law coefficients for secondary effluent particles are provided in Table 5.5.

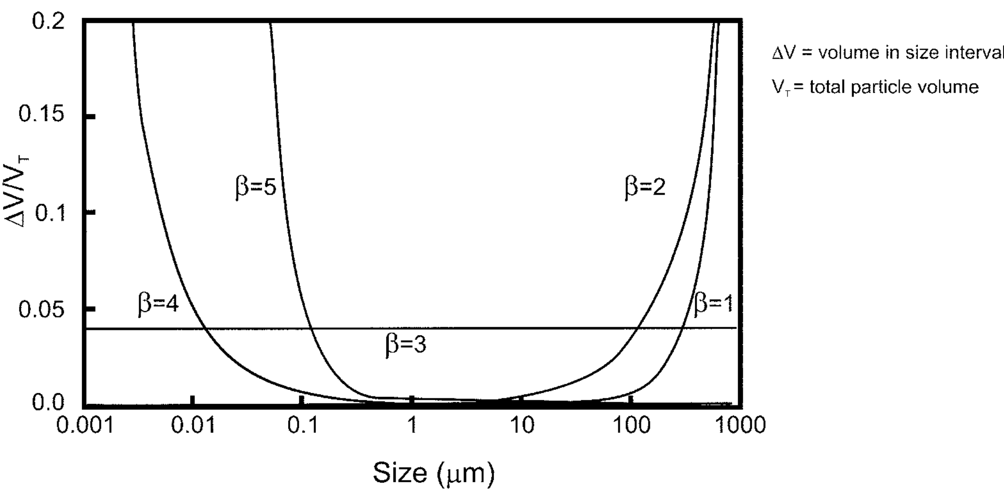


**FIGURE 5.1** Relative contribution of particle size classes to total surface area for a power law frequency distribution (Reprinted with permission from Kavanaugh, M. C.; Tate, C. H.; Trussel, A. R.; Treweek, G. [1980] Use of Particle Size Distribution Measurement for Selection and Control of Solid/Liquid Separation Processes. Particles in Water. In Advances in *Water: Characterization, Fate, Effects, and Removal;* Kavanaugh, M. C.; Luckie, J. O. Eds.; Copyright 1980 American Chemican Society).

**TABLE 5.5** Reported values of the power law exponent for secondary effluent particles.

| Source | Method[c] | β | Correlation coefficient | Reference |
|---|---|---|---|---|
| Activated sludge effluent | Microscope | 3.3 | 0.99 | (Parker, 1970) |
| Effluent, lime treated wastewater | HIAC | 3.1 | 0.97 | (Kavanaugh et al., 1978) |
| Haifa biological filters | HIAC/ROYCO PC 320 | 2.21–2.81 | 0.83–1.0 | (Adin et al., 1989) |
| Aerated lagoon effluent | HIAC/ROYCO PC 320 | 1.33–2.9 | 0.91–0.93 | (Adin et al., 1989) |
| Activated sludge effluent (Ein-Kerem, Jerusalem) | HIAC/ROYCO | 1.6 | 0.99 | Adin and Eli melech, 1989) |
| Jerusalem activated sludge effluent | HIAC/ROYCO PC 320 | 1.73, 2.93 | 0.93, 0.95 | Adin et al., 1989) |
| Hamburg-Stellingen secondary effluent; conventional with chemical phosphorus removal | Serial membrane filtration | 1.0 | 0.91 | (Neis and Tiehm, 1997) |
| Seevetal secondary effluent BNR with chemical phosphorus removal | Serial membrane filtration | 0.8 | 0.96 | (Neis and Tiehm, 1997) |
| Bad Bramstedt secondary effluent; BNR with chemical phosphorus removal | Serial membrane filtration | 1.0 | 0.99 | (Neis and Tiehm, 1997) |
| Secondary effluent (Ede WWTP, Netherlands, Biodenipho) | Met One PCX | 2.75 | — | van der Graaf et al., 2001) |
| Flocculated effluent ($FeCl_3$[a], $AlCl_3$[b]) (Ede WWTP, Netherlands, Biodenipho) | Met One PCX | 3.04 | — | (van der Graaf et al., 2001) |

[a]$FeCl_3$ = ferric chloride.

[b]$AlCl_3$ = aluminum chloride.

[c]HIAC, HIAC/ROYCO, and Met One are particle counting equipment brand names now marketed by HACH Ultra Analytics, Vésenaz, Switzerland; PC 320 and PCX are particle counting equipment model numbers.

Particles in wastewater effluent are mostly colloidal in nature and negatively charged. Typical floc sizes are in the range of 10 to 70 μm, with floc densities of 1.015 to 1.034 g/cm$^3$ (Andreadakis, 1993). Most particles have electrostatic charges in the range of −10 to −18 mV and tend to be stable (Adin and Elimelech, 1989). Adin et al. (1989), by scanning electron microscope analysis of activated sludge effluent particles, showed that the surface of effluent particles is dominated by silica, chloride, and calcium.

A typical PSD for secondary effluent is 20 000, 14 000, and 1 000 particles/1 mL for particle size ranges of 5 to 10, 10 to 15, and 30 to 40 μm, respectively. Roth (1981) found particle concentrations between 5.5 x $10^4$ (6.0 mg/L total suspended solids [TSS]) to 2.3 x $10^6$/mL (93.3 mg/L TSS) in secondary effluent with mean particle sizes between 3 and 50 μm, with few particles greater than 300 μm. Figure 5.2 contains three secondary effluent cumulative PSDs from one study (Roth, 1981), whereas

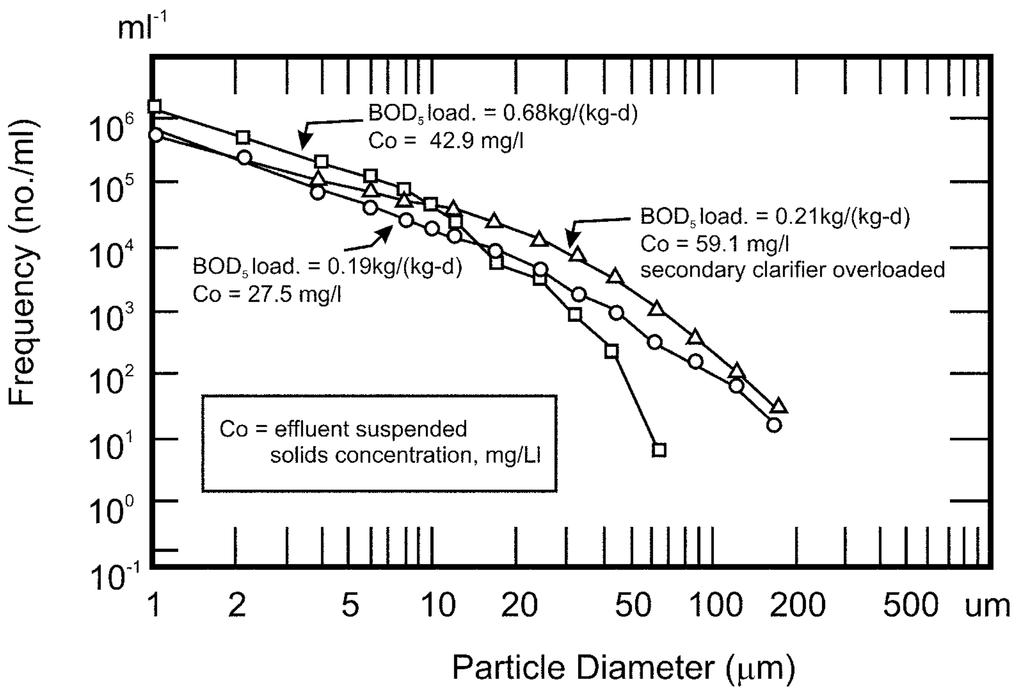


**FIGURE 5.2** Particle size distribution in secondary effluent (BOD$_5$ = 5-day biochemical oxygen demand) (from Roth, 1981).

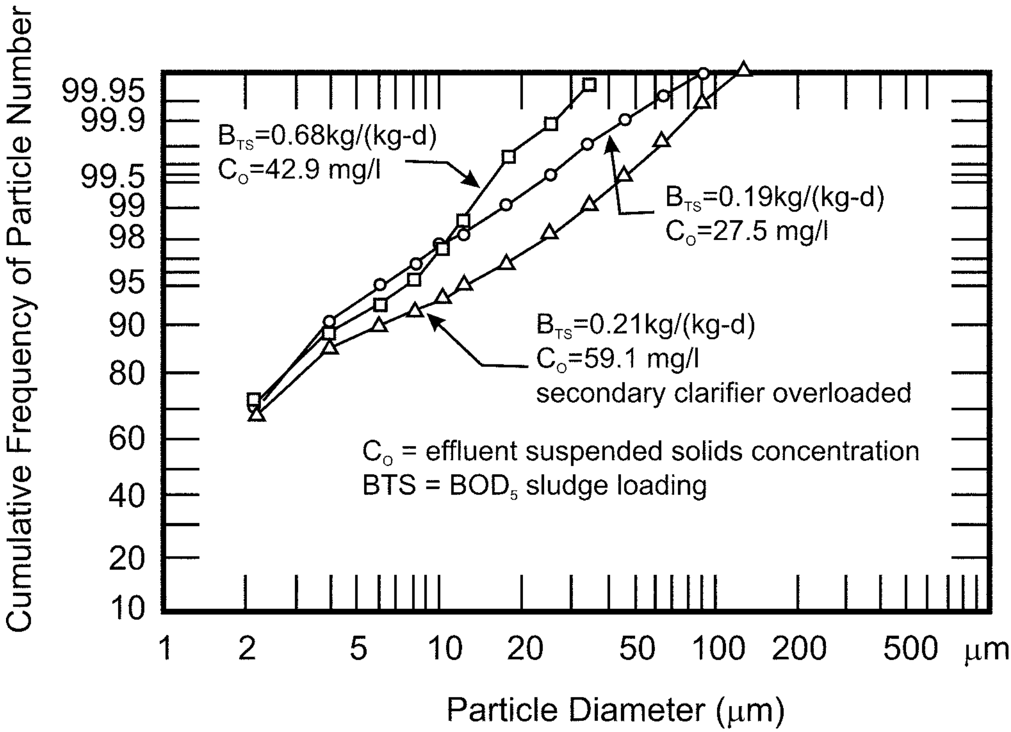


**FIGURE 5.3** Cumulative frequency distribution of particle numbers in secondary effluent (from Roth, 1981).

Figure 5.3 contains the cumulative particle number distribution. Tiehm et al. (1999) found that only 9% of particle sizes at one plant exceeded 32 μm but 55% were larger than this at a second plant and 30% were larger than this at a third plant. Figure 5.4 contains plots of the particle mass distribution versus the particle size in the effluent from four different plants (Tiehm et al., 1999). In contrast, another study (Schubert and Gunthert, 2001) reported that more than 90% of effluent particles were smaller than 32 μm at a trickling filter plant. Tchobanoglous and Eliassen (1970) reported (using optical microscopy) a bimodal frequency distribution, with the majority of particles smaller than approximately 20 μm and another smaller group between 50 and 100 μm. The frequency plot for the particle sizes is reproduced in Figure 5.5. Most researchers, however, have found continuous PSDs as shown in Figure 5.6 (Patry and Takacs, 1992). Parker (1983) attributed bimodal PSDs to excessive shear on

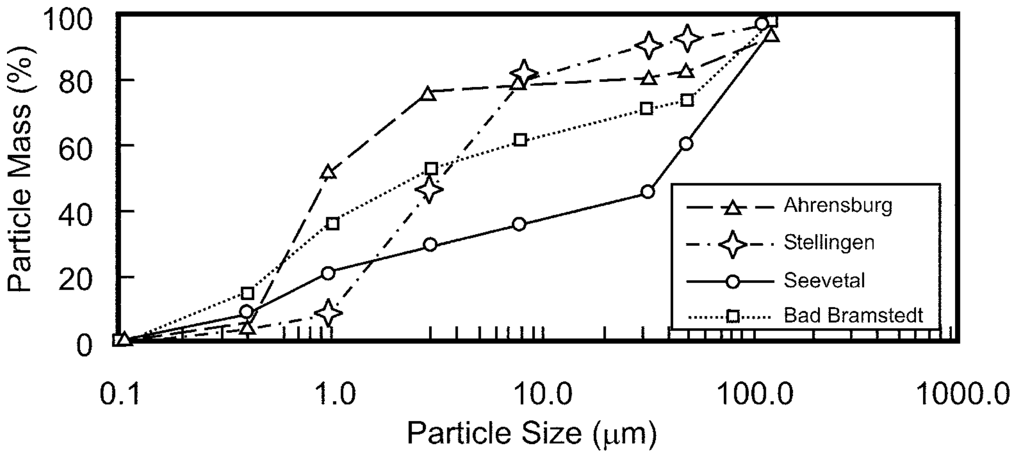


**FIGURE 5.4** Particle size distribution in secondary effluent (reprinted from *Water Science & Technology*, **39,** 99–106, with permission from the copyright holder, IWA).

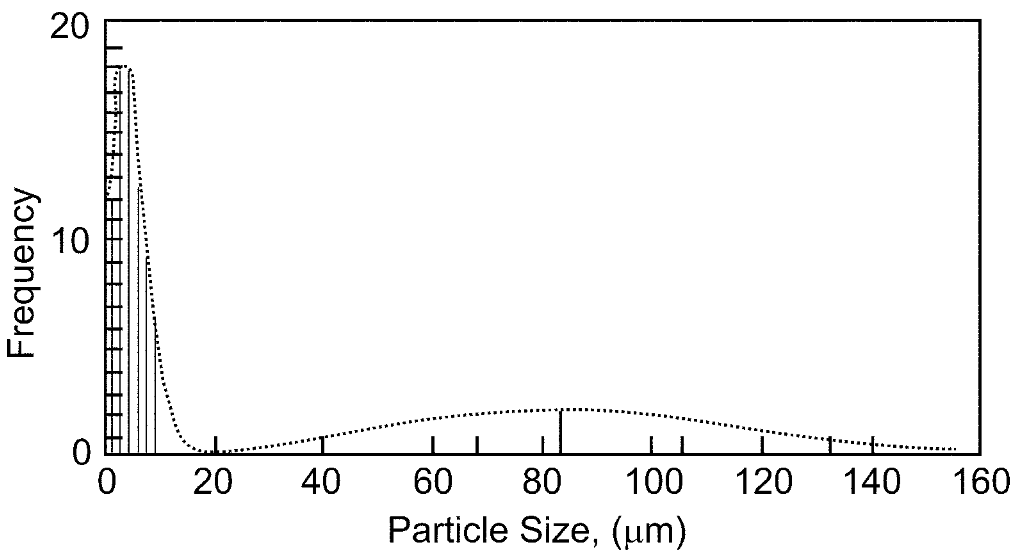


**FIGURE 5.5** Effluent particle size distribution in secondary effluent (reprinted from Tchobanoglous, G; Eliassen, R. [1970] Filtration of Treated Sewage Effluent. *J. Sanit. Eng. Div., Am. Soc. Civ Eng.*, **96,** Journal Paper, 7210, 243, with permission from the publisher, ASCE).

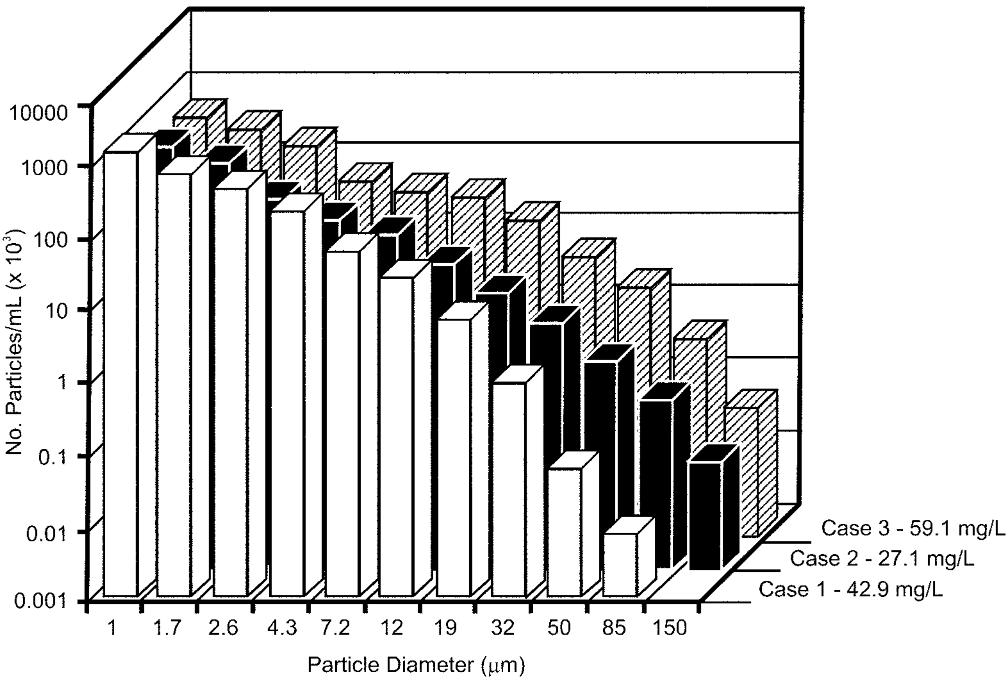


**FIGURE 5.6** Effluent particle size distribution in secondary effluent (reprinted from *Water Research,* **25,** 1263–1271, with permission from the copyright holder, IWA).

the floc. A plot of particle number versus particle size from Parker's work is reproduced in Figure 5.7. A pilot study conducted to compare the effect of mixing intensity on physical characteristics of biological floc also found a bimodal frequency distribution (Galil et al., 1991). This study used optical microscopy to measure floc size and summarized available correlations between mixing intensity and floc size. This summary is reproduced in Table 5.6.

**SETTLING VELOCITIES AND OVERFLOW RATES.** From a rational perspective, the sizing of tertiary clarifiers should be done on the same basis as clarifiers for other purposes. As proposed in Chapter 1, tertiary clarifiers should also be based on the settling regime expected and the range of expected settling velocities for the appropriate type of solids with an allowance for deviations from ideal conditions. Expected settling regimes and settling are discussed for the two major types of solids

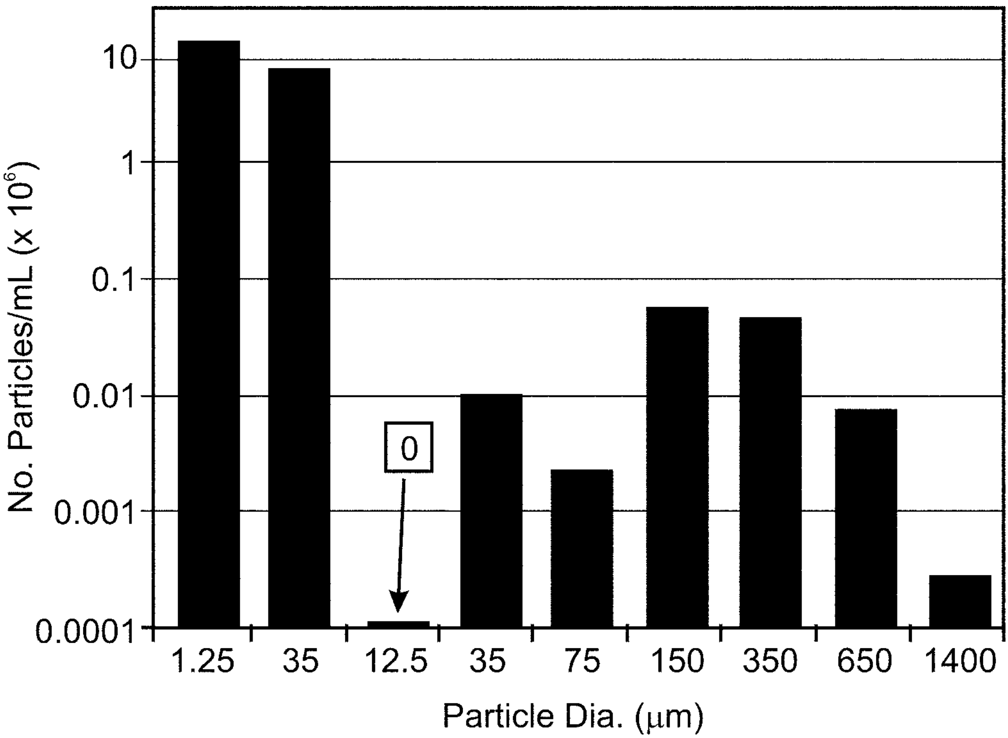


FIGURE 5.7 Particle size distribution in the mixed liquor of an activated sludge process (from Parker et. al., 1971).

TABLE 5.6 Relationship between floc size and velocity gradient—comparison of experimental results (Galil et al., 1991).

| | **Mathematical relationship for mean floc size** | **Floc length for $G^* = 70\ sec^{-1}$** |
|---|---|---|
| Galil et al. (1991) | | |
| Large size | $23\,840/G^{0.71}$ | 1167 |
| Small size | $23\,840/G^{0.71}$ | 267 |
| Argaman and Weddle (1974) | $23\,840/G^{0.71}$ | 185 |
| Parker et al. (1971) | | |
| (for largest floc size not mean) | $23\,840/G^{0.71}$ | 1198 |
| Lentvaar and Rebhun (1991) | $23\,840/G^{0.71}$ | 263 |

$^*G$ = velocity gradient ($sec^{-1}$).

experienced in tertiary clarifier applications—effluent biological suspended solids that have escaped from secondary sedimentation tanks and post secondary treatment chemical precipitates.

Relatively few data are available on the settling properties of either secondary effluent suspended solids or chemical precipitates, especially when compared with the significant body of work on the settling properties of activated sludge mixed liquor. Settling velocities can be inferred from available particle size information and published design clarifier overflow rates; however, direct measurement of settling velocities is preferred and recommended to the extent practical. Theoretically, a settling tank designed with an overflow rate of 1.0 m/h should remove all particles larger than approximately 40 μm, assuming Stokes' law, a water temperature of 20°C, spherical particles, and a particle specific gravity of 1.30. Particles smaller than 100 μm in diameter will require hydraulic detention times greater than that typically provided (Schubert and Gunthert, 2001). Particles smaller than 25 μm should not be expected to be removed by settling under any realistic loading (Parker, 1983).

***Dispersed Activated Sludge Effluent Suspended Solids.*** Secondary effluent suspended solids are those solids not removed in the activated sludge sedimentation tanks because the particles are too small to settle under extant conditions or because hydraulic currents exist that are capable of transporting the solids to effluent weirs. If a tertiary clarifier is provided in series with a secondary clarifier, the settling will occur in conformance with type 2 flocculent settling unless the particles are sufficiently dispersed so that little opportunity exists for flocculation. Then settling will occur as type 1 discrete nonflocculent settling (Ekama et al., 1997). Without the addition of coagulant, most tertiary clarifiers in series will likely experience type 1 settling with the low solids concentrations typical of a well-operated activated sludge process. Tertiary settling tank performance will be a function of influent suspended-solids characteristics, suspended-solids concentration, and hydraulic conditions in the clarifier. Tertiary clarifiers used for suspended-solids removal should have a significantly different hydraulic regime from a typical secondary clarifier because density currents caused by high concentrations of solids should be absent. Under some conditions, tertiary clarifiers, particularly with chemical coagulation, could be an economical alternative to filters, secondary clarifier modifications, or additional secondary clarifiers for meeting low effluent suspended-solids concentrations (< 5 mg/L).

A survey of 21 plants found that the average equilibrium concentration of suspended solids (after batch flocculation and settling) was 6.8 mg/L, which is significantly less than the effluent suspended solids of most municipal activated sludge plants (Ekama et al., 1997; Wahlberg et al., 1994). Flocculation can be improved by eliminating excessive shear, addition of flocculation facilities, or feeding chemicals such as metal salts or polymers. Clarifier testing using the procedures detailed in *WERF/CRTC Protocols for Evaluating Secondary Clarifier Performance* (Wahlberg, 2001) can be used to determine the feasibility of improving effluent suspended solids from secondary clarifiers by improving flocculation or modifying clarifier hydraulics. Then the construction of tertiary clarification facilities can be developed as an alternative concept on a rational basis and compared with other treatment options (see Table 5.18 later in this chapter for a listing of treatment plants using tertiary clarifiers for removal of suspended solids).

***Chemical Precipitates.*** Chemical precipitates in tertiary clarification applications will exhibit either type 2 (flocculent settling) or type 3 (hindered settling), depending on the quantity of floc produced. Prediction of settling velocities for type 2 and type 3 settling must consider the suspended solids concentration. Settling tests are the preferred method for determining design velocities for tertiary applications as compared to the use of textbook settling velocities. Lime clarification applications will almost certainly experience hindered settling. For hindered settling, the settling velocity can be modeled with the Vesilind formula (Vesilind, 1968) and the solids loading rate and hydraulic loading rate can be predicted with a solids flux/state point type analysis. In water treatment, alum and iron floc are expected to have settling velocities on the order of 0.60 to 0.90 m/h, whereas lime softening floc will have velocities of 0.90 to 3.4 m/h (depending on the amount of magnesium hydroxide precipitated) (Mazyck, accessed 2003). Particle settling velocities for common suspensions from Weber (1972) are shown in Table 5.7. Higher velocities are indicated for aluminum and iron hydroxide floc and a lower velocity is indicated for lime floc.

The settling velocity of suspensions ($v_s$) has been related to the settling velocity of the individual particles ($v_p$) according to the following equation (*Water Treatment Plant Design*, 1979).

$$v_s = v_p\left(1 - f\phi^{2/3}\right) \tag{5.2}$$

Where

$f$ = shape factor (2.78 for alum and ferric floc),
$\varphi$ = floc volume fraction, and
$\varphi$ = 0.06 to 0.10 alum and lime floc (0.076 suggested for design).

**Table 5.7** Settling velocities for common suspensions (Weber, 1972).

| Nature of solids | Specific gravity | Settling velocity (m/h*) |
|---|---|---|
| Sand, silt, clay | 2.650 | 0.25 |
| Aluminum and iron floc | 1.002 | 2.99 |
| Calcium carbonate precipitates | 1.200 | 1.51 |
| Primary waste organics | 1.001 | 1.51 |
| Activated sludge organics | 1.005 | 7.20 |

*m/h × 589 = gpd/sq ft.

Settling velocities calculated by this equation are presented in Table 5.8. These values are similar for alum floc but significantly higher for lime floc when compared with those of Weber (1972). For lime floc, a distinction should be made between floc composed mostly of calcium carbonate precipitates and those with a significant fraction of magnesium hydroxide.

Typical recommended design hydraulic overflow rates for tertiary clarifiers are presented in Tables 5.9 and 5.10.

**Table 5.8** Settling velocities for floc particles (Water Treatment Plant Design, 1979).

| Kind of floc | Range of settling velocities (m/h*) |
|---|---|
| Fragile floc, color removal with alum | 2.2–4.4 |
| Medium floc, algae removal with alum | 3.6–5.5 |
| Strong floc, turbidity removal with alum | 4.4–6.4 |
| Strong floc, lime–soda softening | 4.4–6.4 |
| Crystalline floc, calcium carbonate granules | 7.3–12.1 |

*m/h × 589 = gpd/sq ft.

**Table 5.9** Published design overflow rates for tertiary clarifiers.

| Application | Design overflow rates (m/h*) | Source |
|---|---|---|
| Alum and iron | | |
| Alum coagulation—turbidity removal | 2.0 | (Water Treatment Plant Design, 1989) |
| Alum coagulation—color removal | 1.5–1.7 | (Water Treatment Plant Design, 1989) |
| Iron and alum—average | 0.9–1.7 | (U.S. EPA, 1975) |
| Aluminum and iron floc | 2.99 | (Weber, 1972) |
| Coagulation/flocculation and sedimentation | 1.4–1.7 | (Wastewater Reclamation and Reuse, 1998) |
| Sedimentation | 0.85–3.4 | (Culp et al., 1978) |
| Lime | | |
| Lime softening—low magnesium | 3.4 | (Water Treatment Plant Design, 1989) |
| Lime softening—high magnesium | 2.7 | (Water Treatment Plant Design, 1989) |
| Calcium carbonate precipitates | 1.5 | (Weber, 1972) |
| Lime—solids contact, average | 2.0–3.0 | (U.S. EPA, 1975) |
| Lime treatment, peak | 2.5–3.4 | (Wastewater Reclamation and Reuse, 1998) |

*m/h × 589 = gpd/sq ft.

**COAGULATION AND FLOCCULATION.** Examination of available particle size data suggests that the most efficient use of tertiary clarification is provided when sedimentation tanks are preceded by coagulation and flocculation to agglomerate the individual floc into large particles with higher settling velocities. In this manner, coagulants are used to enhance the removal of organic matter, increase settling velocity, remove phosphorus, remove metals, and improve granular media filtration.

**TABLE 5.10** Recommended settling tank loading rates (WPCF, 1983[a]).

| Type of treatment | Conventional clarifier design loading rate (m/h[b]) | | Sludge blanket clarifier design loading rate (m/h) | |
|---|---|---|---|---|
| | Without polymer | With polymer | Without polymer | With polymer |
| Primary settling of lime treated raw wastewater, pH 9.0–10.0 | 1.44 | 2.88 | 2.16 | 3.60 |
| Primary settling of $Fe^{3+}$ or $Al^{3+}$ treated raw wastewater | 1.08 | 2.16 | 1.80 | 2.88 |
| Secondary settling of activated sludge treated with $Fe^{+3}$ or $Al^{+3}$ in aeration basin | 1.08 | — | — | — |
| Tertiary treatment and settling | | | | |
| With lime to pH 10.0 | 1.44 | 2.88 | 3.60 | 5.40 |
| With $Fe^{3+}$ or $Al^{3+}$ | 1.08 | 2.16 | 2.16 | 2.88 |

[a]WPCF (1983) also suggested a minimum *SWD* of 3.7 m for Fe and Al for large settlers and $SWD = 1.6\sqrt{D}$ (SWD = side water depth); feed wells 30–40% of the tank diameter; feed well depths 60–70% of tank depth; and a maximum bottom slope of 1:12.

[b]m/h × 589 = gpd/sq ft.

A range of chemicals and process configurations are available to optimize performance and minimize sludge production and costs.

Several studies have shown that the use of rapid mix and flocculation before tertiary clarifiers and the use of all three before filtration provides better removal of turbidity, solids, pathogens, and phosphorus than chemical addition alone prior to filtration (Ghosh et al., 1994; Kirkpatrick and Asano, 1986; Schimmoller et al., 2000). Sedimentation tank geometry has also been shown to play a role in solids-removal efficiency (Mihopulos and Hahn, 1992). Figure 5.8 summarizes the results from one study that evaluated the effect of geometry. Several facilities using solids-contact basins for lime clarification have reported operating problems (Culp et al., 1978). Wesner (1997) recommended that the operation of solids-contact facilities with lime can be improved significantly by the addition of external rapid mix and flocculation facilities.

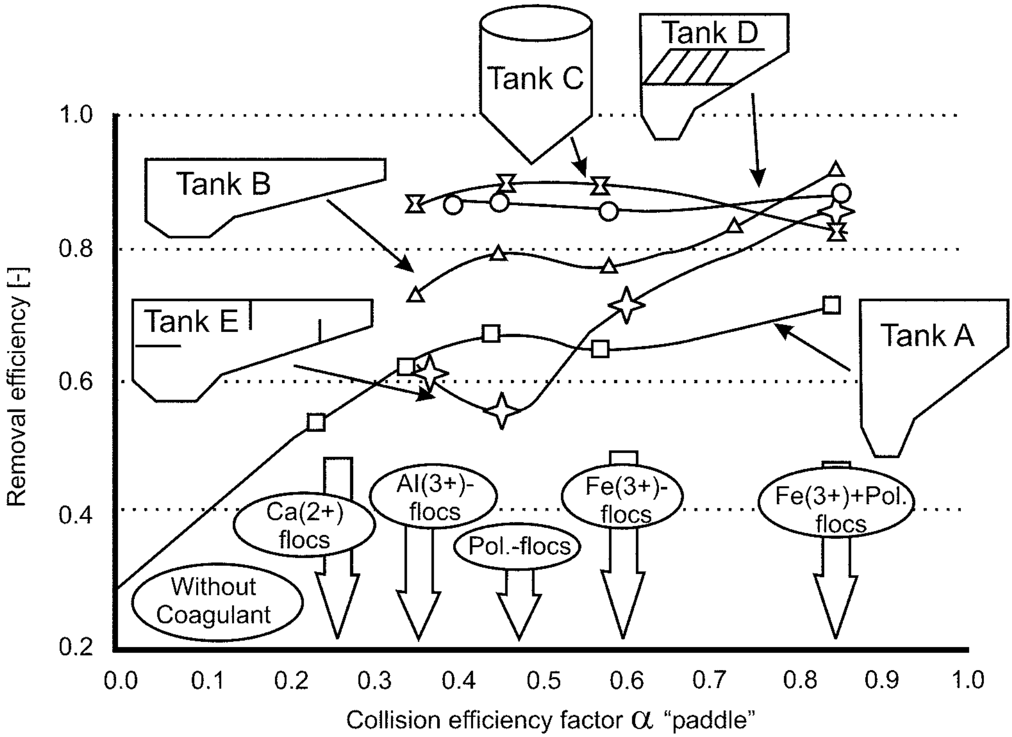


**Figure 5.8** Floc-removal efficiency in different sedimentation tanks as a function of floc formation (described by collision efficiency factor; overflow rate = 0.58 m/h) (from Mihopulos, J.; Hahn H. H. [1992] Effectivity of Liquid–Solids Separation as a function of Apparatus Characteristics and Wastewater Quality. In *Chemical Water and Wastewater Treatment II-Proceedings of the 5th Gothenburg Symposium;* Klur, R., Hahn, H. H. Eds; with kind permission of Springer Science and Business Media).

Aggregation of colloidal particles includes two separate and distinct steps (Weber, 1972).

1. Particle transport to cause interparticle contact and
2. Particle destabilization to permit attachment when contact occurs.

Coagulation has been defined as the overall process of particle aggregation, including both particle destabilization and particle transport, with the term flocculation referring only to particle transport for interparticle contact (Letterman et al., 1990; Weber, 1972) . More practical definitions for these functions follow (Culp et al., 1978):

*"Coagulation: The process whereby chemicals are added to a wastewater resulting in a reduction of the forces tending to keep suspended particles apart. This process physically occurs in a rapid mix or flash mix basin.*

*Flocculation: The agglomeration of suspended material to form particles that will settle by gravity.*

*Sedimentation: The separation of suspended solids from wastewater by gravity.*

*Coagulants: Chemicals, such as alum, iron salts or lime, added, in relatively large concentrations, to reduce the forces tending to keep suspended particles apart.*

*Coagulation–flocculation aids: Materials used in relatively small concentrations which are added either to the coagulation and/or flocculation basins and may be classified as (1) oxidants, such as chlorine or ozone; (2) weighting agents, such as bentonite clay; (3) activated silica; and (4) polyelectrolytes. Polyelectrolytes dissolved in water may ionize to have a positive, a negative, or no charge and are, therefore, referred to respectively as cationic, anionic and non-ionic."*

Coagulation is used to destabilize small particles in wastewater so that they will more readily coalesce into larger size particles. Current theory on coagulation and flocculation (Adin, 1998; Letterman et al., 1990; Odegaard, 1979; Weber, 1972) states that there are four basic mechanisms by which particles are destabilized and agglomerated. These are

- Double layer compression,
- Surface charge neutralization,
- Adsorption and interparticle bridging, and
- Sweep coagulation.

Of these mechanisms for coagulation, it is believed that only two—charge neutralization and sweep coagulation—play an important role in wastewater coagulation. Double layer compression does not apply because the coagulants used are not indifferent electrolytes and undergo other reactions, and the ionic strength of wastewater is too low to have a significant effect. Figure 5.9 contains pH/solubility diagrams for aluminum and iron hydroxides. Bench-scale testing of secondary effluent can be used with such diagrams to identify optimum design conditions for destabilization of particles in secondary effluent (Johnson and Amirtharajah, 1983). An alum coagulation diagram for a secondary effluent is reproduced in Figure 5.10 (Johnson and Amirtharajah, 1983).

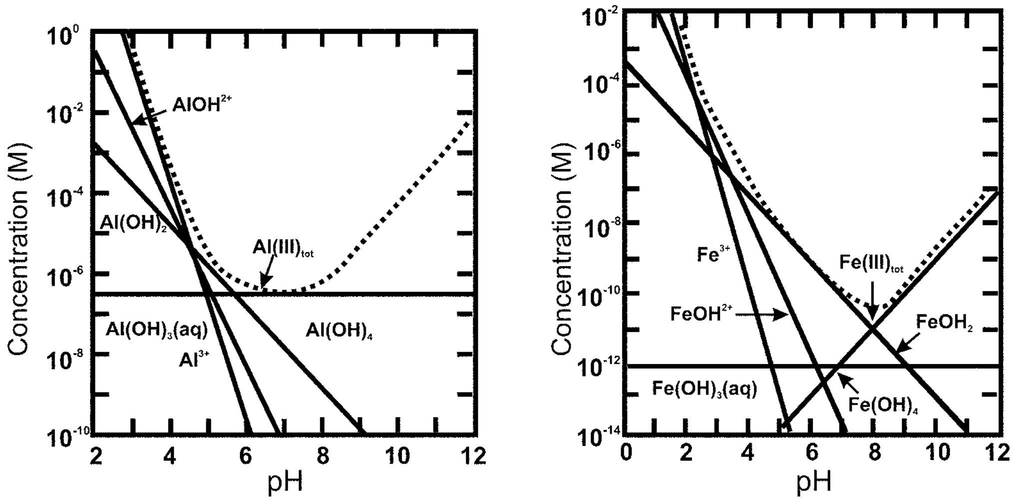


**FIGURE 5.9** Solubility of aluminum and iron hydroxides (from Stumm and Morgan, 1996).

Flocculation fosters the particle transport needed for the growth of the floc created by coagulation into larger particles of settleable size. There are two types of flocculation:

- Perikinetic flocculation, which occurs from thermal agitation (Brownian movement) and is a natural and random process, and
- Orthokinetic flocculation, which occurs from induced velocity gradients in the water.

Orthokinetic flocculation has traditionally been considered the dominant mechanism in wastewater treatment. However, recent studies show that orthokinetic flocculation is ineffective for particles between 1 and 10 μm because the size of the fluid eddies typically are greater than 10 μm and will not affect smaller particles (Metcalf and Eddy, 2003).

Floc formation initially occurs in the coagulation step and results in particles with sizes between 0.5 and 5 μm (Odegaard, 1979). During flocculation, the particle size increases into the range of 100 to 5000 μm. Coagulation and flocculation are always associated with some type of solids–liquid separation process, and optimization of the process will depend on the type of separation process being used (Tambo, 1990).

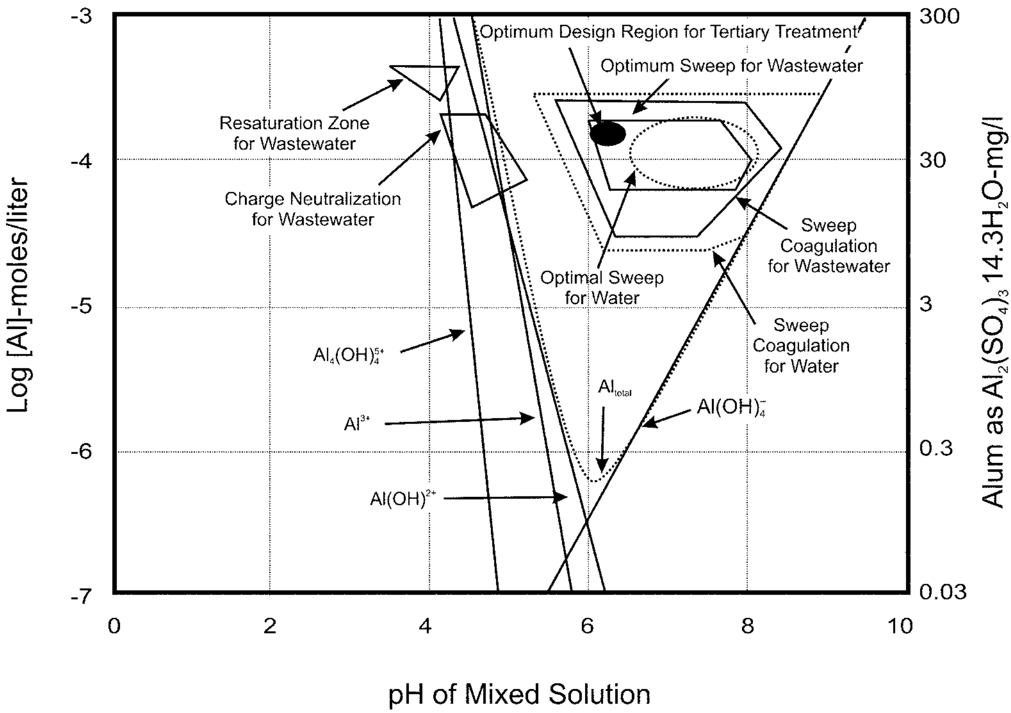


**FIGURE 5.10** Design and operation diagram for alum coagulation (Adapted from *Journal AWWA,* Vol. 75, No. 5 [May 1983], by permisison. Copyright © 1983, American Water Works Association).

Rapid mix and flocculation facilities are traditionally sized using the concepts of mean velocity gradient ($G$) and detention time ($t$). The traditional equation for $G$ is

$$G = \sqrt{\frac{1{,}000P}{\left(\frac{\mu}{1000}\right)V}} = 425\sqrt{\frac{P_w}{t}} \tag{5. 3}$$

Where

$G$ = root-mean-square velocity gradient (the rate of change of velocity) (m/s/m),
$P$ = power input (kW),
$P_w$ = water horsepower per unit flowrate (hp/mg),
$V$ = reactor volume ($m^3$),
$\mu$ = absolute viscosity of the water, N-s/$m^2$ or cP (see Table 5.11), and
$t$ = theoretical detention time, minutes.

**Table 5.11.** Water viscosity and water temperature (AWWA and ASCE, 1997).

| Temperature (°C) | Temperature (°F) | $\mu$ (cP[a]) | $\mu$ (lb-s/sq ft[b]) |
|---|---|---|---|
| 0 | 32 | 1.792 | $3.75 \times 10^{-5}$ |
| 5 | 41 | 1.520 | $3.17 \times 10^{-5}$ |
| 10 | 50 | 1.310 | $2.74 \times 10^{-5}$ |
| 15 | 59 | 1.145 | $2.39 \times 10^{-5}$ |
| 20 | 68 | 1.009 | $2.10 \times 10^{-5}$ |
| 25 | 77 | 0.895 | $1.87 \times 10^{-5}$ |
| 30 | 86 | 0.800 | $1.67 \times 10^{-5}$ |

[a]cP × 0.001 = Pa·s.
[b]lb-s/sq ft × 4.788 026*E* + 01 = Pa · s.

Design parameters for rapid mixing facilities include velocity gradient, detention time, reactor configuration, and mixing device. The time required for particle destabilization is fast—approximately $10^{-10}$ to 1.0 seconds. Jar testing is recommended to establish the design hydraulic detention time and velocity gradients, as the traditional rapid mix detention time of 30 to 60 seconds may not be optimum and excessive shear will break up the floc. Coagulation for water with PSDs that show a significant fraction smaller than 10 μm must be based on perikinetic flocculation and not the velocity gradient concept (Metcalf and Eddy, 2003).

Typical *G* values for rapid mixing are 1500 to 5000 $sec^{-1}$ for inorganic coagulants and 400 to 800 $sec^{-1}$ for organic polymers with corresponding mixing times of 30 to 60 seconds (Young and Edwards, 2000). *G* values in conventional flocculation tanks are between 10 and 75 $sec^{-1}$ and hydraulic retention times are in the range of 10 to 30 minutes, which gives *Gt* values of $10^4$ to $10^5$. In general, the higher the velocity gradient and the higher the coagulant dose, the lower is the required rapid mixing period. For plug-flow rapid mix devices, optimum velocity gradients should be approximately 1200 to 2500 $sec^{-1}$ (Bratby, 1980).

Principle design parameters for flocculation are the velocity gradient, reactor configuration, and detention time. Coagulation reactions are typically first order so tanks in series provides better performance than a single complete-mix reactor. Traditional rapid mix and flocculation design standards, as summarized in Table 5.12, do not distinguish *G* values according to the type of coagulant used. Caution should

**TABLE 5.12** Typical rapid mix and flocculation design parameters.

| Rapid mix | | Flocculation | | |
|---|---|---|---|---|
| **$G$ value ($s^{-1}$)** | **T** | **$G$ value ($s^{-1}$)** | **T** | **Source** |
| ≥300 | 15–60 s | 10–200 | 10 000–100 000 (GT) | (Culp et al., 1978) |
| ~300 complete-mix<br>1200–2500 plug flow | 10–60 s | — | — | (Wiechers et al., 1987) |
| 300–1000 | 20–60 s<br>30 metal salts | 50–80 | <15 min | (U.S. EPA, 1987a) |
| — | 1–5 min | — | 15–30 min | (U.S. EPA, 1987b) |
| 300–1500 | 0.5–2.0 min | — | — | (WEF, 1992) |
| — | 0.5–5 s | 20–100 | 10–30 min | (Richard, 1998) |
| 500–1500 | 5–30 s | 50–100 | 30–60 min | (Metcalf and Eddy, 2003) |
| 1500–6000 | < 1 s | 25–150 | 2–10 min | (Metcalf and Eddy, 2003) |

be exercised when design parameters developed for potable water applications are transposed to wastewater applications.

Coagulation theory is not sufficiently advanced to enable the designer to predict particle-removal efficiencies from basic parameters, and the designer and operator must rely on jar testing and full-scale operations to establish the optimum coagulant, dose, pH, coagulation time, and mixing intensity. An example of the effect of chemical dose on the removal of turbidity, total plate count, and zeta potential is shown in Figure 5.11 (Adin, 1999), whereas Figure 5.12 provides an example of the effect of chemical dose on the effluent PSD (Ghosh et al., 1994). Jar testing procedures are provided in (Ghosh et al., 1994; Hudson, 1981; Wagner, accessed 2003; Weichers et al., 1987). Aluminum, iron, and calcium ions tend to react first with soluble phosphate, and sufficient metal salt must be added to first precipitate the phosphate before forming metal hydroxide precipitants (Weber, 1972).

The efficiency of coagulation is affected by four primary parameters (Weber, 1972):

- Coagulant dose,
- pH (and alkalinity),

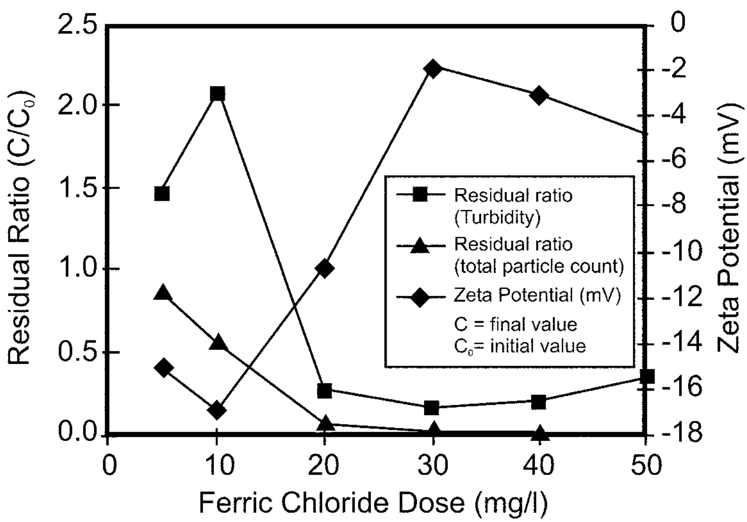


**Figure 5.11** Variation in residual ratio of turbidity, total plate count (TPC), and zeta potential as a function of ferric chloride dose at pH 5.05 (reprinted from *Water Science & Technology;* **40,** 67–74, with permission from the copyright holder, IWA).

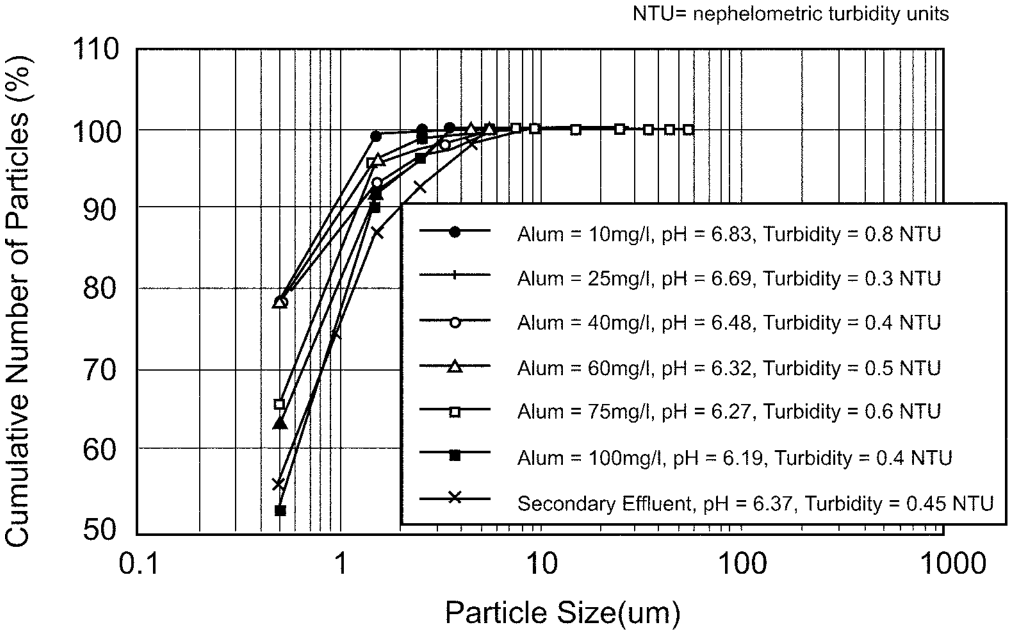


**Figure 5.12** Particle size distribution in untreated and coagulated–settled secondary effluent in the sweep coagulation (high pH) region (NTU = nephelometric turbidity units) (reprinted from *Water Science & Technology;* **30,** 209–218, with permission from the copyright holder, IWA).

- Colloid concentration, and
- Phosphorus concentration (for phosphorus-removal applications).

**COAGULANTS.** Most particles in wastewater carry a negative charge, and the commonly used coagulants provide positively charged ions. Ferric chloride and aluminum sulfate (alum) are the most commonly used coagulants but there are many others that can and have been used. These include lime, ferric sulfate, ferrous sulfate, polyaluminum chloride, and organic polymers. Hydroxide floc formed by the addition of metal salts to water will enmesh and adsorb smaller particles. These mechanisms allow colloidal particles to agglomerate or aggregate into larger particles that will settle by gravity in a reasonable time. Increased numbers of particles and increased particle size improve sedimentation performance by increasing settling velocity and the rate of flocculation. Common properties of commercial coagulants are summarized in Table 5.13.

**TABLE 5.13** Availability of commercial coagulants (Faust and Aly, 1998).

| Common name | Formula | Equivalent weight | pH at 1% | Availability (%) |
|---|---|---|---|---|
| Alum | $Al_2(SO_4)_3 \cdot 14H_2O$ | 114 | 3.4 | Lump—17 $Al_2O_3$ |
| | | | | Liquid—8.5 $Al_2O_3$ |
| Lime | $Ca(OH)_2$ | 40 | 12 | Lump—as CaO |
| | | | | Powder—93–95 |
| | | | | Slurry—15–20 |
| Ferric chloride | $FeCl_3 \cdot 6H_2O$ | 91 | 3–4 | Lump—20 Fe |
| | | | | Liquid—20 Fe |
| Ferric sulfate | $Fe_2(SO_4)_3 \cdot 3H_2O$ | 51.5 | 3–4 | Granular—18.5 Fe |
| Copperas | $Fe_2SO_4 \cdot 7H_2O$ | 139 | 3–4 | Granular—20 Fe |
| Sodium aluminate | $Na_2Al_2O_4$ | 100 | 11–12 | Flake—46 $Al_2O_3$ |
| | | | | Liquid—2.6 $Al_2O_3$ |
| Aluminum chloride | $AlCl_3$ | 44 | — | Liquid |

While alum and ferric chloride are the most commonly used coagulants, there are a number of others available. Jar testing has been used to evaluate the effectiveness of varying degrees of prehydrolysis on the removal of turbidity, phosphorus, carbonaceous oxygen demand (COD) and UV absorbance at 254 nm (Diamadopoulos and Vlachos, 1996). Prehydrolysis creates polymeric species such as $\left[Al_2\,(OH)_2\right]^{4+}$ and $\left[Al_{13}O_4\,(OH)_{24}\right]^{7+}$. The degree of prehydrolysis was defined as $B = \left[OH^-\right]/\left[Me^{+3}\right]$. Potential benefits of prehydrolyzed coagulants are stated to be (Diamadopoulos and Vlachos, 1996):

- Lower coagulant doses,
- Wider range of optimum pH,
- Smaller residual metal concentrations, and
- Less sensitivity to lower temperatures.

Consistent with other studies, phosphorus-removal efficiency decreased with increasing degrees of prehydrolysis as a result of competition between phosphate and hydroxyl ions to occupy sites on the aluminum complex. Prehydrolysis had no measurable effect on COD removal as measured by UV absorbance. In all cases, higher coagulant doses provided greater COD removal. Polyaluminum chloride with $B$ equal to 0.5 and 1.0 provided the optimal removal of turbidity in comparison with alum and polyferric chloride with $B$ equal to 1.8.

Polymers are commonly added in chemical coagulation processes to enhance flocculation and settling. Guidelines have been proposed for the type of polymer to use under different pH conditions (Wiechers et al., 1987).

- pH $<$ 6.5 anionic polyelectrolyte,
- pH $>$ 6.5 cationic polyelectrolyte, and
- Neutral polyelectrolyte where charge neutralization is not a factor.

**METAL PRECIPITATION.** Metals are commonly removed from wastewater by precipitation reactions, and the separation of metal precipitates from water can be accomplished by tertiary clarifiers. Most commonly, metal precipitation is accomplished by the addition of lime and sometimes caustic soda to create metal hydroxides. Caustic soda produces less sludge than lime under some conditions. Aluminum salts, iron salts, lime, sulfides, and synthetic thiopolymers can also be used to precipitate metals. Iron salts are often preferred because aluminum addition can result in an increase in soluble aluminum in the effluent and iron hydroxide will adsorb and co-precipitate metals. In general, iron hydroxides remove an equal or greater amount of a metal (Cu, Cd, Zn, Cr(III), Ni, and Pb) than lime or caustic soda at all pH values (Droste, 1997). Sludge production with alum and ferric chloride is typically much less than with lime precipitation. Metal sulfides are typically several orders of magnitude less soluble than metal hydroxides and sulfides are better at achieving very low effluent metal concentrations.

The solubility product ($K_{SP}$) is defined by following:

$$K_{SP} = \left[Me^{2+}\right]\left[OH^{-}\right]^{2} \tag{5.4}$$

Solubility products for common metal salt precipitates are listed in Table 5.14. Figure 5.13 contains pH/concentration plots for a number of metal oxides and hydroxides.

Though sulfide precipitation can produce lower effluent metal concentrations, odor problems and effluent toxicity can be associated with the use of sulfides. Synthetic thiopolymers will precipitate metals without odor and toxicity problems but are more expensive. Combinations of chemicals can be used to obtain the desired effluent metal concentration at a lower cost. Examples of this would be iron salts/lime, iron salts/thiopolymer, and lime/thiopolymer. The lowest metal concentration achievable with precipitation processes is approximately 0.05 mg/L. A number of challenges are associated with precipitation reactions for the removal of metals. Metals form many complexes, and in addition to increasing the overall solubility of the metal, individual complexes can have significantly higher solubility than the other compounds. A common example of this is the formation of metal ammonia compounds. Different metal salts have minimum solubility at different pH values, thereby complicating attempts to removal of multiple metals in one process.

As with all chemical reactions involving the formation of a solid phase, the performance of metal-removal processes is ultimately connected with the ability to separate the particles from the water. Iron and alum floc should have settling characteristics similar to those experienced in potable water treatment processes and other tertiary clarification applications with the same coagulants; however, this should be verified by testing. In comparison to Table 5.9, a maximum overflow rate of 0.85 m/h has been recommended for precipitation of metals (Eckenfelder, 1980). Reported removals of metals with lime precipitation are listed in Table 5.15.

**TABLE 5.14** Solubility product constants for metal carbonates, hydroxides, and sulfides at 25°C (Metcalf and Eddy, 2003; Sawyer et al., 1994).

| Salt | Equilibrium equation | Solubility product, $pK_s$ |
|---|---|---|
| AgCl | $AgCl \Leftrightarrow Ag^{+} + Cl^{-}$ | 9.52 |
| $CaF_2$ | $CaF_2 \Leftrightarrow Ca^{+2} + 2F^{-}$ | 10.52 |
| $CaCO_3$ | $CaCO_3 \Leftrightarrow Ca^{+2} + CO_3^{2-}$ | 8.30 |
| $MgCO_3$ | $MgCO_3 \Leftrightarrow Mg^{+2} + CO_3^{2-}$ | 4.40 |
| $BaSO_4$ | $BaSO_4 \Leftrightarrow Ba^{+2} + SO_4^{2+}$ | 10.00 |
| $CaSO_4$ | $CaSO_4 \Leftrightarrow Ca^{+2} + SO_4^{2-}$ | 4.70 |
| $Al(OH)_3$ | $Al(OH)_3 \Leftrightarrow Al^{+3} + 3OH^{-}$ | 32.00 |
| $Ca(OH)_2$ | $Ca(OH)_2 \Leftrightarrow Ca^{+2} + 2OH^{-}$ | 5.10 |
| $Cd(OH)_2$ | $Cd(OH)_2 \Leftrightarrow Cd^{+2} + 2OH^{-}$ | 13.70 |
| $Cr(OH)_3$ | $Cr(OH)_3 \Leftrightarrow Cr^{+3} + 3OH^{-}$ | 30.22 |
| $Cu(OH)_2$ | $Cu(OH)_2 \Leftrightarrow Ca^{+2} + 2OH^{-}$ | 18.70 |
| $Fe(OH)_3$ | $Fe(OH)_3 \Leftrightarrow Fe^{+3} + 3OH^{-}$ | 37.22 |
| $Fe(OH)_2$ | $Fe(OH)_2 \Leftrightarrow Fe^{+2} + 2OH^{-}$ | 14.30 |
| $Mg(OH)_2$ | $Mg(OH)_2 \Leftrightarrow Mg^{+2} + 2OH^{-}$ | 11.05 |
| $Zn(OH)_2$ | $Zn(OH)_2 \Leftrightarrow Zn^{+2} + 2OH^{-}$ | 17.10 |
| $Ni(OH)_2$ | $Ni(OH)_2 \Leftrightarrow Ni^{+2} + 2OH^{-}$ | 15.70 |
| $Pb(OH)_2$ | $Pb(OH)_2 \Leftrightarrow Pb^{+2} + 2OH^{-}$ | 15.60 |
| $CaHPO_4$ | $CaHPO_4 \Leftrightarrow Ca^{+2} + HPO_4^{2-}$ | 6.52 |
| $Ca_3(PO_4)_2$ | $Ca_3(PO_4)_2 \Leftrightarrow Ca^{+2} + PO_4^{3-}$ | 27.00 |
| CdS | $CdS \Leftrightarrow Cd^{+2} + S^{2-}$ | 28.00 |
| CuS | $CuS \Leftrightarrow Cu^{+2} + S^{2-}$ | 35.21 |
| FeS | $FeS \Leftrightarrow Fe^{+2} + S^{2-}$ | 17.20 |
| PbS | $PbS \Leftrightarrow Pb^{+2} + S^{2-}$ | 27.15 |
| HgS | $HgS \Leftrightarrow Hg^{+2} + S^{2-}$ | 52.00 |
| NiS | $NiS \Leftrightarrow Ni^{+2} + S^{2-}$ | 24.00 |
| $(Ag)_2S$ | $(Ag)_2S \Leftrightarrow 2Ag^{+} + S^{2-}$ | 28.15 |
| ZnS | $ZnS \Leftrightarrow Zn^{+2} + S^{2-}$ | 22.80 |

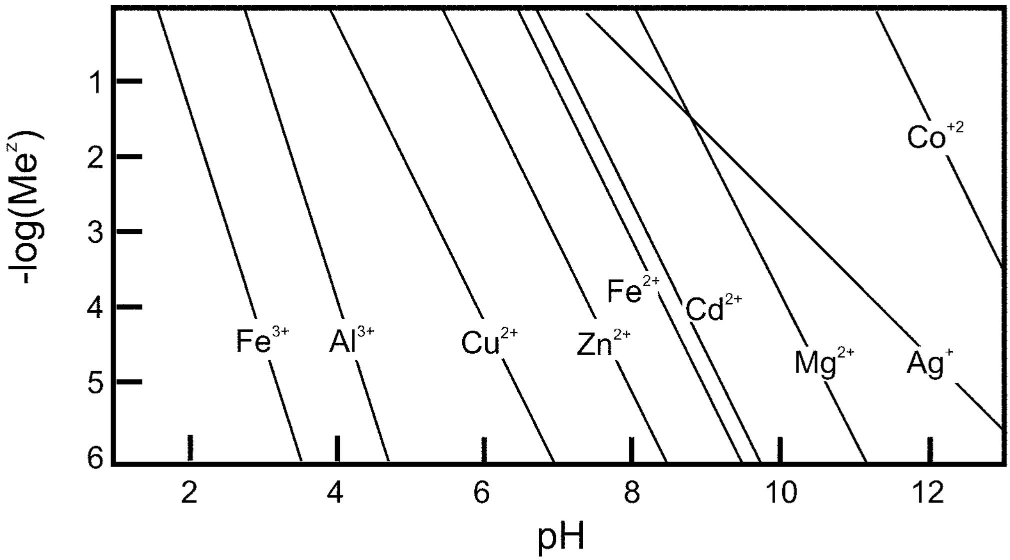


**FIGURE 5.13** Solubility of metal oxides and hydroxides (from Stumm and Morgan, 1996).

**TABLE 5.15** Removal of heavy metals by lime, coagulation, settling, and recarbonation (Argaman and Weddle, 1974; Eckenfelder, 1980).

| | Concentration (mg/L) | | |
|---|---|---|---|
| **Metal** | **Influent** | **Effluent** | **Removal (%)** |
| Ag | 0.24–1.51 | 0.01–0.02 | 96–99 |
| As | 7.00–8.40 | 0.20–0.30 | 96–97 |
| Ba | 0.36–1.08 | 0.04–0.14 | 87–89 |
| Cd | 0.54–5.78 | 0.01–0.19 | 95–99 |
| Co | 0.42–1.29 | 0.04–0.09 | 90–96 |
| $Cr^{+6}$ | 0.45–1.40 | 0.30–1.25 | 11–33 |
| Cu | 0.60–1.47 | 0.04–0.23 | 84–93 |
| Hg | 3.26–4.45 | 0.29–0.61 | 86–91 |
| Mn | 1.37–2.26 | 0.01–0.02 | 99 |
| Ni | 0.75–1.36 | 0.11–0.20 | 85 |
| Pb | 0.41–1.21 | 0.04–0.05 | 90–96 |
| Zn | 7.34–9.61 | 0.12–0.18 | 97–99 |

**CHEMICAL PHOSPHORUS-REMOVAL PROCESSES.** Chemical phosphorus removal converts soluble orthophosphorus to a solid precipitate that can then be removed from the liquid stream. Aluminum, iron, and calcium salts can be used to precipitate phosphates. Standard solubility curves for a number of metal phosphates are shown in Figure 5.14. The minimum phosphorus concentration attainable by precipitation reactions is complicated by the formation of multiple competing species and complexes. Phosphate-precipitation chemistry is described in detail elsewhere and will only be briefly mentioned in this section as it relates to estimating chemical requirements and sludge production. The engineering design of facilities to provide chemical phosphorus removal requires that the designer establish the desired effluent phosphorus concentration; select the type of chemical to use; and estimate chemical doses, alkalinity consumption, and sludge production. Factors to consider with chemical phosphorus removal are the choice of precipitant, initial mixing conditions, the surface charge of precipitated particles, flocculation

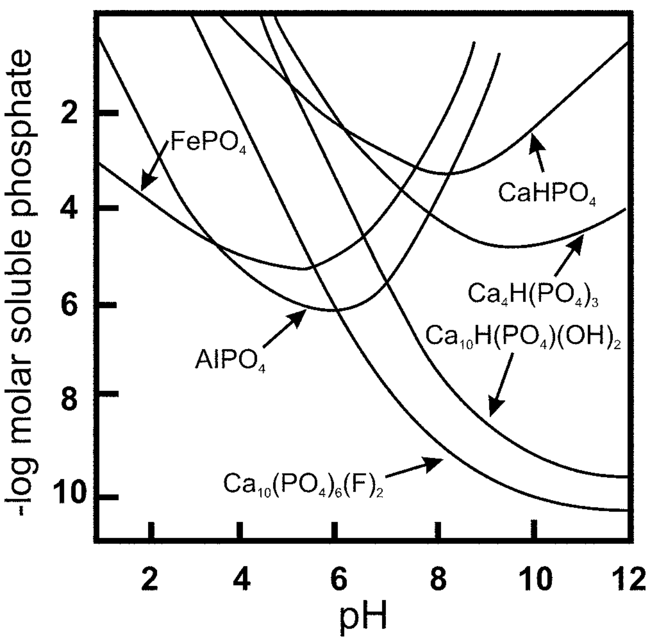


**Figure 5.14** Solubility diagram for solid phosphate phases (from Stumm and Morgan, 1996).

conditions, and the method of solids separation. Tertiary clarification for the precipitation of phosphorus, also known as postprecipitation, is addressed here.

Tertiary phosphorus-control methods are common in Sweden (Balmer and Hultman, 1988). A 1982 survey identified 554 plants out of 760 that used postprecipitation for phosphorus removal, and the plants with postprecipitation had an average effluent total phosphorus concentration of 0.53 mg/L. This study concluded that effluent total phosphorus concentrations below 0.5 mg/L were possible by precipitation alone but that filtration was required to lower the effluent phosphorus below 0.2 mg/L. Postprecipitation was able to provide soluble phosphorus concentrations less than 0.1 mg/L with good operation.

A number of studies have evaluated the ability to meet effluent total phosphorus limits less than 0.2 mg/L using various types of clarifiers and clarification in combination with filtration (Clark et al., 1999; Holtz, 1999; Hunt et al., 2000; Karsen and Brown, 2002; Maldonado, 2002; Mueller et al., 1999; Ross et al., 1994; Sydney Water Corporation, 1998; Wiseman et al., 1999). Based on this work, the lowest phosphorus concentration attainable by chemical precipitation is in the range of 0.05 to 0.07 mg/L. Pilot testing and cost evaluations have indicated that the combination of rapid mix, flocculation, and sedimentation before filtration is more economical for phosphorus removal than rapid mixing or rapid mixing and flocculation prior to filtration (Schimmoller et al., 2000). A combination of pilot- and full-scale testing at Henderson, Nevada, demonstrated that inclined plate gravity settling and continuous backwash upflow filters required less coagulant than solids-contact clarifiers and downflow filters to produce an effluent total phosphorus concentration less than 0.10 mg/L because the solids-contact clarifiers required a minimum coagulant dose for stable operation (Hunt et al., 2000). Testing at the Rock Creek advanced wastewater treatment plant showed that a combination of solids-contact clarifiers and filters produced an effluent total phosphorus of 0.05 mg/L, whereas testing with direct filtration was only able to achieve an effluent total phosphorus concentration of 0.15 mg/L (Mueller et al., 1999).

Work by the Sydney Water Corporation on their South Creek wastewater treatment plants has shown that the use of tertiary clarifiers and filters is allowing them to produce effluent with very low total phosphorus concentrations (Karsen and Brown, 2002). Effluent with a total phosphorus concentration between 0.01 and 0.04 mg/L is being produced by the Quakers Hill wastewater treatment plant. Before the addition of tertiary clarifiers, Sydney Water Corporation was able to lower the effluent total phosphorus at the Quakers Hill and St Marys wastewater treatment plants from typical values of 0.25 mg/L and 0.68 mg/L, respectively, to approximately 0.06 mg/L by optimizing operating procedures and refurbishing existing tertiary filters (Karsen and Brown, 2002).

Gwinnett County, Georgia, was able lower the effluent total phosphorus at their Beaver Ruin and Jackson Creek water reclamation facilities from less than 1.0 mg/L to less than 0.25 mg/L by increasing alum and polymer addition and adding multiple feed points (Muckerman, 2001). A study evaluated methods to upgrade phosphorus removal at both plants to meet a decrease in the effluent total phosphorus limit from 1.0 to 0.3 mg/L and recommended that tertiary clarifiers be constructed (Muckerman, 2001).

Typical operating data from a two-stage lime clarification process shows that a two-stage lime process can reach lower effluent total phosphorus concentrations than single-stage lime clarification (Sydney Water Corporation, 1998). First-stage effluent total phosphorus of approximately 0.16 mg/L and second-stage concentrations of approximately 0.07 mg/L have been reported; however, performance varies with the lime dose and pH achieved.

**DESIGN METHODS.** In estimating the requirements for chemical phosphorus removal, consideration must be given to the different species of phosphorus that exist in wastewater. Phosphorus species found in secondary effluent include soluble phosphorus compounds, colloidal particles of phosphorus precipitants, and organic phosphorus bound in the effluent volatile solids as represented by the following equation:

$$C_{T,\,eff} = S_{P,\,eff} + X_{P,\,eff} + X_{VSS} m_{P,\,vss} \tag{5.5}$$

Where

$C_{T,\,eff}$ = total effluent phosphorus concentration (mg/L),
$S_{P,\,eff}$ = soluble effluent phosphorus compounds (mg/L) (see Table 5.16),
$X_{P,\,eff}$ = particulate effluent inorganic phosphorus concentration (mg/L),
$X_{vss}$ = effluent volatile suspended solids concentration (mg/L), and
$m_{P,\,VSS}$ = fraction phosphorus in effluent volatile suspended solids (unitless).

The effluent particulate inorganic phosphorus concentration is typically assumed to be zero, although for applications that require very low effluent phosphorus concentrations this must be carefully evaluated.

Formulas for calculating chemical dose, chemical feed rates, alkalinity consumption, and sludge production are given below:

$$D_{Me} = \left(\frac{Me}{P}\right)\left(\frac{MW_{Me}}{MW_P}\right)\left(C_{P,in} - \Delta C_{P,proc} - C_{P,res}\right) \tag{5.6}$$

Where

$D_{Me}$ = metal ion dose required (mg/L Me ion),
Me/P = practical molecular ratio of Me ion to P ion required to reach desired $C_{P,\ res}$,
$MW_{Me}$ = molecular weight Me ion (mg/mM),
$MW_P$ = molecular weight P ion (mg/mM),
$C_{P,in}$ = total influent phosphorus concentration (mg/L),
$\Delta C_{P,\ proc}$ = phosphorus removed in treatment process (mg/L), and
$C_{P,\ res}$ = total residual soluble phosphorus concentration (mg/L) = $[H_3PO_4] + [H_2PO_4^-] + [HPO_4^{-2}] + [PO_4^{-3}] + [MeH_2PO_4^{-2}] + [MeHPO_4^{+}]$

Phosphorus compounds formed by chemical precipitation are summarized in Table 5.16.

Graphs of theoretical and measured values for $C_{P,\ res}$ can be found in *Biological and Chemical Systems for Nutrient Removal* (WEF, 1998). For the purposes of estimating chemical dose, the total residual soluble phosphorus should be calculated from eq 5.5. Conservatively, phosphorus removed in the treatment process by conventional mechanisms, $\Delta C_{P,\ proc}$, can be assumed to be zero. Otherwise, typical phosphorus removal in conventional primary clarifiers may be estimated as 1% of the volatile suspended solids (VSS) removed; however, there is little published information on this. In conventional biological treatment processes, phosphorus removal can be estimated as 2% of the VSS produced, according to the following equation:

$$C_{P,\ proc} = m_{P,VSS}\ Y_b\ \Delta S \tag{5.7}$$

Where

$Y_b$ = net biomass yield, mass biomass produced per mass five-day biochemical oxygen demand ($BOD_5$) removed (mg VSS/m $BOD_5$); and

$m_{P,\ VSS}$ = 0.020 to 0.025 (unitless).

**Table 5.16** Phosphorus precipitates (adapted from Jenkins and Hermanowicz, 1991).

| Metal ion | Possible precipitate | Formula | Molecular weight | Approximately pH of minimum solubility |
|---|---|---|---|---|
| Ca (II) | β-Tricalcium phosphate | $Ca_3(PO_4)_2$ | 310 | > 11 |
| | Hydroxyapatite | $Ca_5(OH)(PO_4)_3$ | 502 | > 11 |
| | Dicalcium phosphate | $CaHPO_4$ | 88.1 | > 11 |
| | Calcium carbonate (calcite) | $CaCO_3$ | 100.1 | > 11 |
| Fe(II) | Ferrous phosphate (vivianite) | $Fe_3(PO_4)_2$ | 357 | ~8 |
| | Ferrous hydroxide | $Fe(OH)_2$ | 89.9 | |
| Fe(III) | Ferric hydroxide | $Fe(OH)_3$ | 106.9 | |
| | Ferric phosphate | $Fe_X(OH)_Y(PO_4)_Z$ | 186 ($r$ = 1) | 4.5–5.5 |
| | (strengite) | $Fe_r(H_2PO_4)(OH)_{3r\text{-}1}$ | 294 ($r$ = 2) | |
| | | | 400 ($r$ = 3) | |
| Al(III) | Aluminum phosphate | $Al_X(OH)_Y(PO_4)_Z$ | 158 ($r$ = 1) | |
| | (variscite) | $Al_r(H_2PO_4)(OH)_{3r\text{-}1}$ | 236 ($r$ = 2)<br>314 ($r$ = 3) | 5.5–6.5 |
| | Aluminum hydroxide | $Al(OH)_3$ | 78 | |
| Mg(II) | Magnesium ammonium phosphate (struvite) | $MgNH_4PO_4$ | 137.3 | ~10.7 |

### *Chemical Quantities*

$$\mathrm{M_{chem}} = \frac{Q_o}{1000} D_{Me} \frac{MW_{chem}}{MW_{Me}} \bullet \frac{1}{f_{chem}} \tag{5.8}$$

Where

$M_{\text{chem}}$ = mass dry chemical required (kg/d),
$Q_{\text{o}}$ = plant flow ($m^3$/d),
$D_{\text{Me}}$ = metal ion dose (mg/L),
$\text{MW}_{\text{chem}}$ = molecular weight of commercial chemical (Da) (see Table 5.17), and
$f_{\text{chem}}$ = purity of commercial chemical (see Table 5.17).

$$\mathrm{V_{chem}} = \frac{M_{chem}}{s_v} \tag{5.9}$$

Where

$V_{\text{chem}}$ = volume liquid chemical required ($m^3$/d) and
$s_{\text{v}}$ = specific volume of liquid commercial chemical (kg/$m^3$).

### *Sludge Production*

$$\mathrm{M_{TS}} = \mathrm{M_{XTSS}} + \mathrm{M_{MePO_4}} + \mathrm{M_{MeOH}} \tag{5.10}$$

Where

$M_{\text{TS}}$ = total additional sludge (mg/L),
$M_{\text{XTSS}}$ = weight additional suspended-solids removal that will result from the addition of the metal salt (mg/L),
$M_{\text{MePO4}}$ = weight metal phosphate sludge generated (mg/L), and
$M_{\text{MeOH}}$ = weight metal hydroxide sludge generated from excess chemical addition (mg/L).

$$\mathrm{M_{MePO_4}} = \Delta C_P \left( \frac{MW_{MePO_4}}{MW_P} \right) \tag{5.11}$$

$$\mathrm{M_{MeOH}} = \left[ \left( \frac{D_{Me}}{MW_{Me}} \right) - \left( \frac{Me}{P} \right) \left( \frac{\Delta C_P}{MW_P} \right) \right] MW_{MeOH} \tag{5.12}$$

**TABLE 5.17** Properties of chemical coagulants used for precipitation of phosphorus.

| Common name | Formula | Molecular weight | Typical purity (wt %) | Weight ratio (chemical: phosphorus) | pH of aqueous solution | Alkalinity* as $CaCO_3$ | Bulk density ($kg/m^3$) |
|---|---|---|---|---|---|---|---|
| Quick lime (dry) | $CaO$ | 56.0 | 90 $CaO$ | $1.5 \times A^b$ | 12 | 1.79 | 880–960 |
| Hydrated lime (dry) | $Ca(OH)_2$ | 74.1 | 93 $Ca(OH)_2$ | $1.5 \times A^b$ | 12 | 1.35 | 400–540 |
| Aluminum sulfate (dry) | $Al_2(SO_4)_3 \cdot 14H_2O$ | 594.3 | 17 $Al_2O_2$ | 9.6:1 | 3.0–3.5 | -0.45 | 600–1100 |
| | $Al_2(SO_4)_3 \cdot 18H_2O$ | 666.4 | 17 $Al_2O_3$ | 10.8:1 | 3.0–3.5 | -0.45 | 600–1100 |
| Aluminum sulfate (liquid) | $Al_2(SO_4)_3 \cdot 14H_2O$ | 594.3 | 8.3 $Al_2O_3$ 49 alum | 9.6:1 | 3.0–3.5 | -0.45 | 1330 |
| | $Al_2(SO_4)_3 \cdot 18H_2O$ | 666.4 | 8.3 $Al_2O_3$ | 10.8:1 | 3.0–3.5 | -0.45 | 1330 |
| Sodium aluminate (dry) | $Na_2Al_2O_4$ | 164.0 | 41-46 $Al_2O_3$ | 3.6:1 | 11–12 | 0.54 | 640–800 |
| Sodium aluminate (liquid) | $Na_2Al_2O_4$ | 164.0 | 4.9-26.7 $Al_2O_3$ | 3.6:1 | 11–12 | 0.54 | 1400–1500 |
| Ferric chloride (liquid) | $FeCl_3$ | 162.1 | 35-45 $FeCl_3$ | 5.2:1 | 3–4 | -1.85 | 1340–1490 |
| Ferric sulfate (dry) | $Fe_2(SO_4)_3$ | 400.0 | 70-90 $Fe_2(SO_4)_3$ | 6.4:1 | 3–4 | -0.75 | 960–1120 |
| | $Fe_2(SO_4)_3 \cdot 3H_2O$ | 454.0 | | 7.3:1 | 3–4 | -0.75 | |
| Ferrous sulfate (dry) | $FeSO_4$ | 151.9 | 55-58 $FeSO_4$ | 3.2 ($Fe^{+2}$):1 | 3–4 | -0.36 | 700–1200 |
| | $FeSO_4 \cdot 7H_2O$ | 278.0 | | 3.2 ($Fe^{+2}$):1 | 3–4 | -0.36 | 700–1200 |
| Ferrous sulfate (liquid) | $FeSO_4$ | 151.9 | 5 $Fe^{+2}$ | 3.2 ($Fe^{+2}$):1 | 3–4 | -0.36 | 1150 |
| | $FeSO_4 \cdot 7H_2O$ | 278.0 | 5 $Fe^{+2}$ | 3.2 ($Fe^{+2}$):1 | 3–4 | -0.36 | 1150 |
| Ferrous chloride (liquid) | $FeCl_2$ | 126.8 | 20-25% $FeCl_2$ | 3.2 ($Fe^{+2}$):1 | 3–4 | -2.37 | 1190–1250 |

*mg alkalinity as $CaCO_3$ added (+), or removed (−) per milligram of chemical added.

Where

$D_{Me}$ = dose of metal ion (mg/L);
$MW_{MePO4}$ = molecular weight of metal phosphate (Da);
$MW_P$ = molecular weight of phosphours (Da);
$\Delta C_P$ = total phosphorus chemically removed (mg/L); and
Me/P = theoretical dose metal salt (mol Me/mol phosphorus).

Properties of chemical coagulants used for phosphorus precipitation are contained in Table 5.17 and their stoichiometry for phosphorus precipitation, dose, sludge production, and alkalinity consumption are discussed in the following paragraphs.

## ALUMINUM

$Al^{+3} + H_nPO_4^{n-3} \Leftrightarrow AlPO_4(s) + nH^+$

$Al^{+3} + PO_4^{-3} \Leftrightarrow AlPO_4(s)$

Theoretically, 1 mol aluminum will precipitate 1 mol phosphorus. The stoichiometric weight ratio of aluminum to phosphorus is 0.87:1. One mol of alum reacts with 2 mol (190 g) of phosphate containing 62 g phosphorus to form 2 mol (244 g) of aluminum orthophosphate ($AlPO_4$). The stoichiometric weight ratio of aluminum sulfate ($Al_2(SO_4)_3$).$18H_2O$ to phosphorus is 666/62 or 10.8:1. The stoichiometric weight ratio of $Al_2(SO_4)_3.14H_2O$ to phosphorus is 594/62 or 9.6:1. Typically, a dosage of 1.5 to 3.0 mol of aluminum per mol phosphorus is required. For a dosage of 1.5:1, the precipitation of 0.4 kg/d (1 lb/d) phosphorus requires 11.8 L (3.13 gal) of 48% alum solution. The typical optimum pH range for phosphorus removal using aluminum salts is 6.0 to 6.5.

### *Alum Dose*

In metric units:
Weight alum (aluminum sulfate) per unit volume commercial alum

$$= (0.48)(1330 \text{ kg/m}^3) = 638 \text{ kg alum/m}^3$$

In U.S. customary units:
Weight alum (aluminum sulfate) per unit volume commercial alum

$$= (0.48)\ (11.1 \text{ lb/gal}) = 5.33 \text{ lb alum/gal}$$

In metric units:

Weight aluminum per unit volume commercial alum

$$= f_{chem}\,(sv)\,(\text{no. mol Me/mol compound}) \cdot (MW_{Me}/MW_{cmpd})$$
$$= (0.48)\,(1330)\,(2)\,(26.98/666.7)$$
$$= 51.7 \text{ kg Al/m}^3$$

In U.S. customary units:

Weight aluminum per unit volume commercial alum

$$= f_{chem}\,(sv)\,(\text{no. mol Me/mol compound}) \cdot (MW_{Me}/MW_{cmpd})$$
$$= (0.48)\,(11.1)\,(2)\,(26.98/666.7)$$
$$= 0.431 \text{ lb Al/gal}$$

In metric units:

Theoretical aluminum dose per unit mass phosphorus

$$= 1.0 \text{ kg P } (MW_{Al}/MW_{P})\,(Al/P)$$
$$= (1.0)\,(26.98/30.97)\,(1.0)$$
$$= 0.87 \text{ kg Al/kg P}$$

In U.S. customary units:

Theoretical aluminum dose per unit mass phosphorus

$$= 1.0 \text{ lb P } (MW_{Al}/MW_{P})\,(Al/P)$$
$$= (1.0)\,(26.98/30.97)\,(1.0)$$
$$= 1.8 \text{ lb Al/lb P}$$

In metric units:

Actual commercial alum dose per unit mass phosphorus

$$= (Al/P)\,(M_{Al}/M_{P}) \,/\, (sv_{Al})$$
$$= 1.5\,(0.87)/(51.7)$$

$= 0.025$ m$^3$ alum solution/kg P (note specific volume of 1282 kg/m$^3$ gives a value of 0.026)

In U.S. customary units:

Actual commercial alum dose per unit mass phosphorus

$$= (Al/P)\,(M_{Al}/M_{P}) \,/\, (M_{Al}/\text{gal})$$
$$= 1.5\,(1.8) \,/\, (0.431)$$

$= 3.03$ gal alum solution/lb P (note specific volume of 10.7 lb/gal gives a value of 3.13)

$f_{chem}$ = fraction metal salt in commercial chemical;
sv = specific volume of commercial chemical (kg/m$^3$);
(Al/P) = molar ratio of aluminum to phosphorus;
$MW_{Al}$ = molecular weight of aluminum = 26.98 Da;

$MW_{AlPO4}$ = molecular weight of aluminum phosphate = 121.9 Da;
$MW_P$ = molecular weight of phosphorus = 30.97 Da;
$MW_{Al\,(OH)3}$ = molecular weight of aluminum hydroxide = 78 Da;
$MW_{Me}$ = molecular weight of metal (Da);
$MW_{cmpd}$ = molecular weight of metal salt (Da);
$M_{Al}/M_P$ = mass aluminum per mass phosphorus;
$M_{AlPO4}$ = aluminum phosphate sludge (mg/L);
$M_{Al\,(OH)3}$ = aluminum hydroxide sludge (mg/L); and
$\Delta C_P$ = phosphorus removed chemically (mg/L).

***Sludge Quantities***

$$M_{AlPO4} = \Delta C_P\,(MW_{AlPO4}/\,MW_P) = 1\text{ mg P/L }(121.9/30.97) = 3.94\text{ mg }AlPO_4/l$$

$$M_{Al\,(OH)\,3} = [(D_{Al}\,/\,MW_{Al}) - (Al/P)\,(\Delta C_P\,/\,MW_P)]\,MW_{Al\,(OH)\,3}$$
$$= [(1.31/26.98) - (1)\,(1/30.97)]\,78$$
$$= (0.048 - 0.032)\,78$$
$$= 1.25\text{ mg }Al(OH)_3\,/L$$

$M_{TS}$ = 3.94 + 1.25 = 5.19 mg/L additional sludge per mg/L phosphorus removed.

***Alkalinity Reduction***

$$Al_2(SO_4)_3{\cdot}18H_2O + 3Ca(HCO_3)_2 \Rightarrow 2Al(OH)_3(s) + 3CaSO_4 + 6CO_2(g) + 18H_2O$$

$$Al_2(SO_4)_3{\cdot}14H_2O + 3Ca(HCO_3)_2 \Rightarrow 2Al(OH)_3(s) + 3CaSO_4 + 6CO_2(g) + 14H_2O$$

One mole of alum reacts with 3 mol of alkalinity. Therefore, 1 mg/l of alum reacts with 1/(666.7)·(3)(100) = 0.45 mg/l alkalinity as calcium carbonate ($CaCO_3$).

The reaction between sodium aluminate and phosphorus is as follows:

$$Na_2O{\cdot}Al_2O_3 + 2PO_4^{-3} \Leftrightarrow 2AlPO_4(s) + 2NaOH + OH^-$$

The molar ratio of aluminum to phosphorus is 1:1, the weight ratio is 0.87 to 1.00; and the weight ratio of sodium aluminate to phosphorus is approximately 3.6:1.

## IRON

$$Fe^{+3} + H_nPO_4^{n-3} \Leftrightarrow FePO_4 + nH^+$$

$$Fe^{+3} + PO_4^{-3} \Leftrightarrow FePO_4(s)$$

Theoretically, 1 mol iron will precipitate 1 mol phosphorus. The stoichiometric weight ratio of iron to phosphorus: is 1.8:1. For phosphorus removal, 162.3 g of ferric chloride ($FeCl_3$) reacts with 95 g orthophosphate ($PO_4$) to form 150.8 g ferric phosphate ($FePO_4$), and the weight ratio of $FeCl_3$ to phosphorus is 5.2:1. Typical iron doses are 1.1 to 2.0 mol iron/mol phosphorus as P. The optimum wastewater pH to obtain minimum phosphorus solubility is approximately 5.0.

The molar stoichiometry of iron to phosphorus in ferrous phosphate is 1.5 to 1, whereas in ferric phosphate it is 1 to 1. Thus the amount of phosphorus removed per mole of iron added for the stoichiometric region of dosing (down to soluble phosphorus concentrations of approximately 0.5 mg/L) is more favorable for the ferric than for the ferrous salt.

### *Ferric Chloride Dose*

In metric units:

Weight ferric chloride per unit volume commercial ferric chloride

$$= (0.30)\ (1342\ \text{kg/m}^3)$$
$$= 402\ \text{kg ferric chloride/m}^3$$

In U.S. customary units:

Weight ferric chloride per unit volume commercial ferric chloride

$$= (0.30)\ (11.2\ \text{lb/gal})$$
$$= 3.36\ \text{lb ferric chloride/gal}$$

In metric units:

Weight $FeCl_3$ per unit volume commercial solution

$$= f_{\text{chem}}\ (\text{sv})\ (\text{no. mol Me/mol compound}) \cdot (\text{MW}_{\text{Me}}/\text{MW}_{\text{cmpd}})$$
$$= (0.30)\ (1342)\ (1)\ (55.847/162.2)$$
$$= 138.6\ \text{kg Fe/m}^3$$

In U.S. customary units:

Weight $FeCl_3$ per unit volume commercial solution

$$= f_{\text{chem}}\ (\text{sv})\ (\text{no. mol Me/mol compound}) \cdot (\text{MW}_{\text{Me}}/\text{MW}_{\text{cmpd}})$$
$$= (0.30)\ (11.2)\ (1)\ (55.847/162.2)$$
$$= 1.16\ \text{lb Fe/gal}$$

In metric units:

Theoretical iron dose per unit mass phosphorus

$$= 1.0\ \text{kg P}\ (\text{MW}_{\text{Fe}}/\text{MW}_{\text{P}})\ (\text{Fe/P})$$
$$= (1.0)\ (55.847/30.97)\ (1)$$
$$= 1.8\ \text{kg Fe/kg P}$$

In U.S. customary units:

Theoretical iron dose per unit mass phosphorus

$$= 1.0\ \text{lb P}\ (\text{MW}_{\text{Fe}}/\text{MW}_{\text{P}})\ (\text{Fe/P})$$
$$= (1.0)\ (55.847/30.97)\ (1)$$
$$= 1.8\ \text{lb Fe/lb P}$$

In metric units:

Actual commercial ferric chloride dose per unit mass phosphorus

$$= (\text{Fe/P})\,(M_{\text{Fe}}/M_{\text{P}})/(\text{sv}_{\text{Fe}})$$
$$= (2.0)\,(1.8)/(138.6)$$
$$= 0.026 \text{ m}^3 \text{ ferric chloride solution/kg P}$$

In U.S. customary units:

Actual commercial ferric chloride dose per unit mass phosphorus

$$= (\text{Fe/P})(M_{\text{Fe}}/M_{\text{P}})/(M_{\text{Fe}}/\text{gal})$$
$$= (2.0)\,(1.8)\,/\,(1.16)$$
$$= 3.1 \text{ gal ferric chloride solution/lb P}$$

$MW_{Fe}$ = molecular weight of iron = 55.847 Da;
(Fe/P) = molar ratio of iron to phosphorus; and
$sv_{Fe}$ = specific volume of iron ($kg/m^3$).

### *Sludge Quantities*

$$M_{\text{FePO4}} = \Delta C_{\text{P}}\,(\text{MW}_{\text{FePO4}}/\,\text{MW}_{\text{P}}) = 1 \text{ mg P/L}\,(150.82/30.97)$$
$$= 4.87 \text{ mg FePO}_4\text{/L}$$

$$M_{\text{Fe(OH) 3}} = [(D_{\text{Fe}}\,/\,\text{MW}_{\text{Fe}}) - (\text{Fe/P})\,(\Delta C_{\text{P}}\,/\,\text{MW}_{\text{P}})]\,\text{MW}_{\text{Fe (OH) 3}}$$
$$= [(3.61/55.847) - (1)\,(1/30.97)]\,106.9$$
$$= (0.0646 - 0.0323)\,106.9$$
$$= 3.45 \text{ mg Fe(OH)}_3\text{ /L}$$

$M_{TS}$ = 4.87 + 3.45 = 8.32 mg/L additional sludge per mg/L phosphorus removed.

### *Alkalinity Reduction*

$$FeCl_3 + 3H_2O \Rightarrow Fe(OH)_3(s) + 3H^+ + 3Cl^-$$

$$3H^+ + 3HCO_3^- \Rightarrow 3H_2CO_3$$

One mole of ferric chloride reacts with 3 mol alkalinity. Therefore, 1 mg/L of ferric chloride reacts with 1/(162.2)·(3)(100) = 1.85 mg/L alkalinity as $CaCO_3$. The alkalinity required for 1 mg/L of ferrous sulfate is 0.36 mg/L; the lime required is 0.40 mg/L; and the oxygen required is 0.029 mg/L.

## LIME

$$Ca(OH)_2 + H_2CO_3 \Rightarrow CaCO_3 + 2H_2O$$

$$Ca(OH)_2 + Ca(HCO_3)_2 \Leftrightarrow 2CaCO_3 + 2\,H_2O$$

$$10Ca^{+2} + 6PO4^{-3} + 2OH^- \Leftrightarrow Ca_{10}(PO_4)_6(OH)_2$$

The quantity of lime required to precipitate phosphorus is typically 1.4 to 1.5 times the total alkalinity. Between pH 9.0 and 10.5, precipitation of calcite and apatite compete. Precipitation of phosphorus can be modeled as an equilibrium reaction between calcite and hydroxyapatite.

$$10CaCO_3(s) + 2H^+ + 6HPO_4^{-2} + 2H_2O \Leftrightarrow Ca_{10}(PO_4)_6(OH)_2(s) + 10HCO_3^-$$

$$K_{phos} = 10^{+32} = [HCO_3^-]^{10} / [H^+]^2[HPO_4^{-2}]^6$$

if $A$ = alkalinity = $[HCO_3^-]$ and $[HPO_4^{-2}] = C_p$

$$\log\ C_P = +\frac{5}{3}\log\ A + \frac{1}{3}pH - 5.33$$

where $K_{phos}$ is the equilibrium constant for precipitation of phosphorus from water as a calcium phosphate.

Particulate phosphorus remains in suspension for hours or days at pH 9 to 10 at concentrations of several milligrams per liter (Butler, 1991). Flocculation of particulate phosphorus rather than precipitation of dissolved phosphorus is the key mechanism for good phosphorus removal. Good flocculation will not occur until the pH is increased to at least 11.5, and this may increase the equilibrium phosphorus concentration substantially. Small concentrations of magnesium will increase the rate of flocculation at lower pH values. This can be provided by adding a small percentage (<15%) of sea water (Butler, 1991).

## TYPES OF TERTIARY CLARIFIERS

**EXISTING FACILITIES.** Existing tertiary clarifier installations were identified from literature searches, manufacturers' reference lists, Internet searches, and personal experience. Table 5.18 contains a summary of facility information about selected installations that were identified and for which such data were available. Tables 5.19 and 5.20 list existing facilities that use tertiary clarification with lime and with high-rate clarification. Like the majority of existing tertiary clarifiers, the newer tertiary clarifier facilities use high-rate clarification to provide phosphorus removal.

**LIME CLARIFICATION.** Lime clarification is an established and proven tertiary clarification process. While more modern technologies have effectively replaced lime clarification for many applications, the ability of lime precipitation to remove specific inorganic pollutants can make it a viable tertiary treatment alternative in special circumstances. Up until 1995, tertiary lime clarification was a key unit process in nearly all water reclamation facilities producing reclaimed water for high-end uses such as industrial process water and indirect potable reuse. Lime treatment's popularity was due to its ability to remove phosphates, sulfates, organic matter, magnesium and calcium hardness, iron and manganese, and heavy metals and to destroy or remove pathogens such as bacteria and viruses. In the case of membrane treatment, process recovery can be limited by the presence of sparingly soluble salts of calcium, barium, strontium, and silica that are not removed by primary and secondary treatment of wastewater. Lime clarification, used as a pretreatment process prior to reverse osmosis membranes, removes such scale-forming compounds from the feed water to the membrane processes. Sludge production generated by lime can be minimized by stripping carbon dioxide and using acid. Wastewater composition plays a significant role in the overall efficiency of the lime clarification process.

Regarding design aspects for membrane treatment, the threshold concentration for influent silica that will not result in reverse osmosis membrane scaling for a certain recovery can be calculated as

$$SiO_{2C} = SiO_{2f} \times \frac{1}{\left(1-Y\right)} \tag{5.13}$$

Where

$SiO_{2c}$ = silica concentration in concentrate (mg/L),
$SiO_{2f}$ = silica concentration in influent (mg/L), and
$Y$ = recovery of the reverse osmosis system, expressed as a decimal.

Use of commercial antiscalants or threshold inhibitors can increase the solubility of silica, thereby increasing recovery in the reverse osmosis system.

***One Stage Versus Two Stage.*** To achieve maximum removal of a majority of the sparingly soluble constituents, excess lime is typically added to raise the pH of the feed water to between 11.0 and 12.0. Literature data suggest that, at this high pH, most of the phosphates, magnesium, silica, and heavy metals are precipitated. High pH (> 11) is also sufficient to result in extremely low calcium levels, provided an

**Table 5.18** Existing wastewater treatment facilities with tertiary clarification.*

| Facility | Location | Purpose | Chemical | Design flow (permit/ maximum ) ($m^3/d$) | No. | Size (m) | SWD (m) | Design (m/h) | Actual (m/h) | Performance data |
|---|---|---|---|---|---|---|---|---|---|---|
| Upper Occoquan | Centreville, Virginia | Phosphorus removal | Lime at 150 mg/L | 204 400 | 14 | 38 | 3.7 | 1.5 | | |
| Rock Creek | Hillsboro, Oregon | Phosphorus removal | Alum | 75 700/132 500 | 6 | | | | | |
| Durham | Durham, Oregon | Phosphorus removal | Alum | 75 700/170 300 | 3 | | | | | |
| McMinnville WWTP | McMinnville, Oregon | Phosphorus removal | Alum (20 mg/L)<br>AlClH (5 mg/L)<br>Polymer (0.3 mg/L) | 22 700 | 2 | 21 | 6.1 | 1.2 | 0.4 | TSS = 2<br>Orthophosphorus = 0.04<br>Total phosphorus = 0.2 |
| Quaker's Hill | Sydney, Australia | Phosphorus removal | Alum | 82 900 | 4 | 1230 $m^2$ total | | | 2.8 | <5 mg/L |
| Rouse Hill | Sydney, Australia | Phosphorus removal | | | | | | | | |
| St. Mary's | Sydney, Australia | Phosphorus removal | Alum | | | | | | 2.4 | |
| Tahoe-Truckee | Truckee, California | | Lime | 18 300/33 800 | | 19.8 | 3 | 1.4/2.7 | 2.9/5.4 | |
| Blue River | Silverthorne, Colorado | Phosphorus removal | Alum | ADF = 6000<br>Peak = 6800<br>Design = 15 100 | 2 | 15.2 | 3 | | 1.4 | |
| Marlborough Easterly | Marlborough, Massachusetts | Phosphorus removal | Alum | 20 800/62 500 | 4 | 24.4 | 3.7 | 0.5 | 1.4 | BOD = 7.1<br>Total phosphorus = 0.44<br>TSS = 6.5 |

| | | | | | | | | |
|---|---|---|---|---|---|---|---|---|
| Buchanan | Buchanan, Michigan | Phosphorus removal | Ferric chloride | 5700 | 1 | 18.3 | | |
| Laughlin WRF | Clark County, Nevada | Phosphorus removal | Alum Polymer | 11 400 | 1 | 35.0 | | 1.8 |
| South Bend | South Bend, Indiana | Phosphorus removal | | 45 800 | 2 | 61 | | |
| Libertyville | Libertyville, Illinois | | | 7600 | 2 | 16.8 | | |
| Marshall | Marshall, Minnesota | | | 3800 | 90 | 27.4 | | |
| Napa | Napa, California | Lagoon effluent | | 56 800 | 4 | 24.4 | | |
| Alexandria | Alexandria, Minnesota | | | 26 500 | 1 | 22.9 | | |
| Chisholm Creek | Oklahoma City, Oklahoma | | | 18 900 | 1 | 27.4 | | |
| West Camden | Sydney, Australia | Phosphorus removal | | 10 800 | | | | |
| Big Sister Creek | Angola, New York | Phosphorus removal | Ferric chloride Anionic polymer | —/37 800 | 2 | 22.9 | 4.6 | 1.9 |
| Black River | Lorain, Ohio | Phosphorus removal | | —/56 800 | 2 | 38.1 | 3.7 | 1.0 |
| Grove City | Grove City, Pennsylvania | Phosphorus removal | | —/11 700 | 2 | 12.2 × 12.2 | 4.6 | 2.1 |
| F. Wayne Hill WRF | Gwinnett County, Georgia | Phosphorus removal | Lime | —/126 000 | 4 | 32.0 | | 1.6 |
| Northwest STP | El Paso, Texas | Phosphorus removal | Lime | —/5900 | 1 | 12.2 × 12.2 | | 2.1 |
| League City WWTF | League City, Texas | Phosphorus removal | Lime | —/42 600 | 1 | 33.5 | | 2.0 |

*SWD = side water depth; ADF = average daily flow; OP = orthophosphate; TP = total phosphorus; WRF = water reclamation facility; STP = sewage treatment plant; WWTF = wastewater treatment facility; $m^3/d$ ÷ 3785 = mgd; m × 3.281= ft; and m/h × 589 = gpd/sq ft.

**TABLE 5.19** Full-scale wastewater reclamation plants with lime clarification.*

| Plant | Year started | Design flow ($m^3$/d) | Unit processes |
|---|---|---|---|
| Windhoek, Namibia | 1969 | | Dissolved air flotation, lime clarification, recarbonation, filtration, chlorination |
| OCWD Water Factory 21 | 1975 | 19 000 | Lime clarification, recarbonation, mixed media gravity filters. |
| Upper Occoquan Sewage Authority | 1978 | 204 000 | 2-stage lime clarification, filtration, granular activated carbon, ion exchange, chlorination |
| Tahoe-Truckee | 1978 | 28 000 | 2-stage lime clarification, filtration, GAC, ion exchange, chlorination |
| Saudi Arabia Petromin | 1980 | 12 500 | 2-stage lime, cooling towers, filtration, granular activated carbon , $Cl_2$, dechlorination |
| El Paso Fred Hervey Reclamation Plant | 1985 | 37 800 | PACT, two-stage lime clarification, filtration, ozonation, granular activated carbon, chlorination |
| Vero Beach Municipal Power Plant | 1992 | 760 | Chemical clarification, dual media filtration, granular activated carbon, reverse osmosis |
| Madras Fertilizers Limited | 1993 | 12 000 | Activated sludge, 2-stage lime, air stripping, 2-stage filters, $Cl_2$, dechlorination, reverse osmosis |
| Madras Refineries Limited | 1993 | 9500 | Activated sludge, 2-stage lime, air stripping, 2-stage filters, $Cl_2$, dechlorination, reverse osmosis |
| West Basin MWD— | 1994 | 19 000 | Lime treatment, filtration |
| El Segundo Plant | 1997 | 9500 | Microfiltration, reverse osmosis |

*GAC = granular activated carbon; PACT = powdered activated carbon; OCWD = Orange County Water District (California); MWD = Municipal Water District; $m^3/d \div 3785 =$ mgd.

**Table 5.20** Tertiary phosphorus removal facilities using high-rate clarification.

| Plant | Flow ($m^3/d$) | Influent (mg/L) | | Effluent (mg/L) | | Removal (%) | |
|---|---|---|---|---|---|---|---|
| | | Range | Average | Range | Average | Range | Average |
| Herford, Germany | 2 × 15 100 | 0.26–1.23 | 0.62 | 0.03–0.36 | 0.13 | 58–94 | 81 |
| Colombier, Switzerland | 1 × 56 800 | 0.2–3.3 | 1.45 | 0.09–0.76 | 0.28 | 75–95 | 85 |
| Lindsay, Ontario, Canada | 2 × 15 100 | 0.1–0.2 | | 0.03–0.09 | | | 80 |
| Acheres, Paris, France | 5 × 390 000 | 1.6–4.6 | 2.7 | 0.03–2.1 | 0.62 | 69–90 | 77 |
| Marin County, California | 7600 | | | | | | |
| Breckenridge, Colorado | 7600 | | | | | | |
| Leucadia, California | 7600 | | | | | | |
| San Rafael, California | 7600 | | | | | | |
| Beauvais, France | 39 400 | | | | | | |
| Golbey, Epinal, France | 53 000 | | | | | | |
| Breckenridge, Colorado | Pilot study | 1.08–1.59 | | 0.06–0.11 | | 91–95 | |
| West Palm Beach, Florida | Pilot study | 0.27–0.4 | | 0.022–0.07 | | 85–93 | |
| Everglades, Florida | Pilot study | 0.147–0.167 | | 0.008–0.011 | | 93–95 | |
| Onondaga County, New York | Pilot study | 0.38–1.06 | 0.7 | 0.04–0.16 | 0.1 | 77–92 | 86 |
| Cincinnati, Ohio | Pilot study | 1.2–5.2 | | 0.03–0.09 | | 50–99 | |
| Fort Worth, Texas | Pilot study | 0.1–10 | | 0.05–2.3 | | 27–98 | |
| Port Clinton, Ohio | Pilot study | 1.6–5.8 | | 0.36–<0.04 | | 80–99 | |
| Midwest, Ohio | Pilot study | 0.8–7.8 | | 0.09–0.6 | | 80–98 | |

appreciable carbonic species concentration remains after $CaCO_3$ precipitation. Feed waters high in noncarbonate hardness, however, may require the addition of carbonic species (e.g., soda ash) to remove excess calcium. The majority of the full-scale wastewater reclamation plants include two-stage lime clarification where excess lime is added to raise the pH to 11.0 to 12.0 in the first stage, thereby precipitating most scale-forming constituents. Effluent from the first stage then passes to a second clarifier where carbon dioxide and soda ash (if required) is added to stabilize the pH of the first stage effluent between 9.5 and 10.5 and precipitate additional calcium. The second stage provides additional process stability and also aids in recovery of lime after recalcination because sludge produced in this stage is free from most of the impurities.

On the other hand, one-stage lime clarification includes a single conventional clarifier or solids-contact clarifier wherein all the precipitation is done. Lowering the pH by addition of carbon dioxide further stabilizes the effluent before subsequent treatment. A significant advantage of one-stage over the two-stage process is the cost savings from the elimination of a second clarifier and additional sludge handling. However, the most commonly encountered problem with one-stage lime clarification is the control of precipitation following recarbonation. This can be overcome by addition of a sufficient quantity of carbonic species (e.g., soda ash) to precipitate most of the calcium and thereby maintain extremely low effluent concentrations. Also by substituting carbon dioxide with sulfuric acid, the pH of the effluent can be stabilized without further precipitation. A typical secondary wastewater was used to calculate residual calcium concentrations using basic equilibrium concepts and elucidate the mechanism of two-stage versus one-stage lime precipitation. The resulting graph for residual calcium versus pH for the two processes is presented in Figure 5.15. The graph indicates that addition of sufficient soda ash in the one-stage process results in extremely low effluent calcium concentration and further addition of carbon dioxide or acid will not cause significant precipitation after pH adjustment.

Wastewater composition plays a significant role in the overall efficiency of the lime clarification process. Particularly, alkalinity and ammonia affect both the lime dose and the precipitation process. Lime treatment of nitrified effluent has been reported to produce a more consistent and higher quality effluent than a nonnitrified effluent (Kluesener et al., 1975). The results suggest that some constituents in nonnitrified wastewater interfere with chemical precipitation reactions. In addition, the lime–ammonia reaction is counterproductive in producing the desired water quality because of the buffering capability of the ammonium/ammonia pair. Literature data

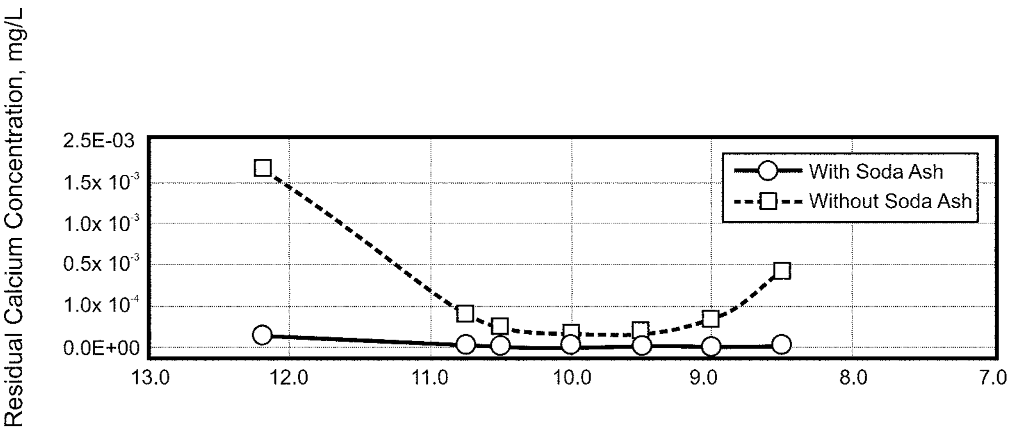


**FIGURE 5.15** Residual calcium concentration (after lime clarification and stabilization by addition of carbon dioxide).

suggest that an increase in the alkalinity of wastewater increases the lime dose proportionally.

***Metal Removal.*** In addition to precipitating a majority of scale-forming constituents from feed water, lime clarification has been traditionally used to remove heavy metals as metal hydroxide precipitates, especially from industrial effluent. Literature data suggest that none of the pH values for maximum precipitation of all metals coincides, hence an optimum pH range must be found, typically ranging from a pH of 7 to 10.5. However, in the case of excess lime treatment, most of the metals are removed to desirable levels for further treatment with reverse osmosis membranes. In some instances, hydroxides that may remain in colloidal state can be settled by addition of polymers. In cases where metal chelates are present in the feed water (compounds such as polyphosphates and synthetic organic polymers, which may not be removed from secondary treatment, may complex with metals), use of a strong oxidizing agent as a preliminary step to break the chelating agents may be an option before lime clarification.

***Silica.*** High silica concentrations are unacceptable in many cooling water applications, as they cause a silica scale precipitate on heat exchanger surfaces. One of the major advantages of lime clarification is its ability to remove silica and dissolved metals from secondary effluent. Several methods are available for removal of silica, including precipitation with aluminum and iron hydroxides, zinc chloride, magnesium oxide, and ultrahigh lime clarification. Alumina and ferric hydroxide have been found to be very effective in removal of silica in the pH range of 8 to 9 by means of adsorption of silica onto hydroxide precipitates. However, the chemical dosages required are in excess of those required by stoichiometry and the settling characteristics of the floc are poor, resulting in poor removal. The use of magnesium oxide results in significant silica removal at a pH of approximately 10.2, with removal increasing with increases in water temperature. A magnesium-to-silica dose of 7: 1 is suggested in the literature. However, the overall silica-removal efficiency with metal ions depends on the initial silica concentrations and increases with higher initial concentrations. Ultrahigh lime treatment (pH of 11 to 12 and high calcium concentrations) was shown to be effective in precipitating silica (Batchelor and McDevitt, 1984). Both high pH and a high calcium concentration were found to be necessary for silica removal. As the lime dose is increased, pH increases; therefore, concentration of the di-negative silicate ion ($H_2SiO_4^{2-}$) increases. This results in supersaturation with respect to solid calcium silicate and an increase in the rate of silica removal. Hence, lime clarification can be used to remove silica in addition to other scale-forming constituents for the feed water to the reverse osmosis process.

Silica removal by lime treatment is limited by the magnesium concentration of the water and the site-specific water chemistry. As a result, there is only a relatively small range of influent silica concentrations within which lime treatment has a decided advantage over microfiltration. And if a significant fraction of the silica is colloidal silica that can be removed by microfiltration or ultrafiltration, lime treatment has no advantage with regard to silica.

This point is illustrated by Figure 5.16. The maximum solubility of silica is approximately 130 to 150 mg/L $SiO_2$ at approximately 25 °C. At 75% recovery, the factor of concentration is 1/ (1-0.75) or 4.0. Thus, reverse osmosis influent silica concentrations greater than approximately 32 mg/L could result in scaling of the reverse osmosis membranes. The addition of a high-quality antiscalant or threshold inhibitor can increase the silica solubility to approximately 240 mg/L. Thus, with an inhibitor the reverse osmosis influent silica concentration can be as high as 60 mg/L. Table 5.19 contains examples of full-scale wastewater reclamation plants with lime clarification.

**HIGH-RATE CLARIFICATION.** High-rate clarification, which is described in more detail in Chapter 3, has also been successfully used for tertiary phosphorus removal. This technology has demonstrated that it produces a lower phosphorus

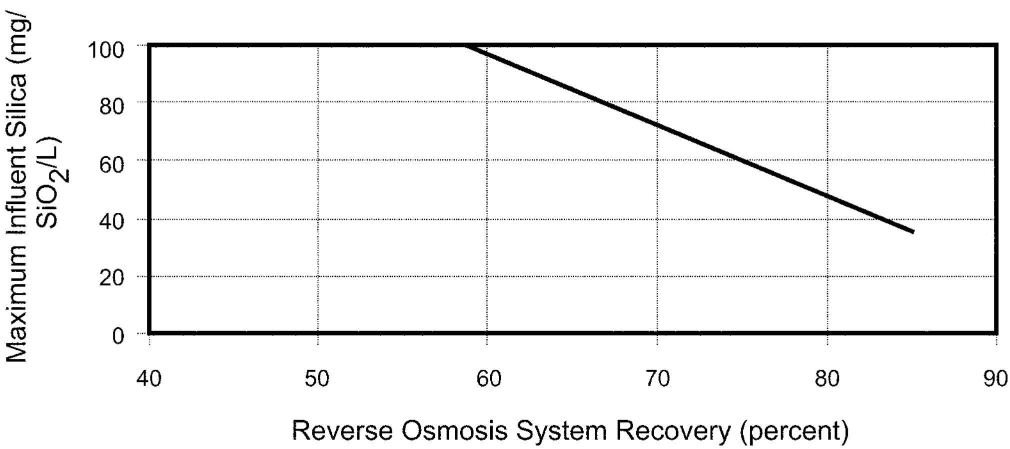


**FIGURE 5.16** Maximum allowable reverse osmosis feedwater silica concentration as a function of system recovery.

concentration in some situations than other competing clarification methods (Holtz, 1999). Full-scale plants using ballasted flocculation for phosphorus removal in tertiary applications are listed in Table 5.20. By constructing high-rate clarification at the end of the treatment process, high-rate clarification can be used to treat secondary effluent for phosphorus removal during dry weather and can be available for the treatment of infrequent peak storm flows diverted around the biological treatment process. Finally, Table 5.21 depicts the operating data summary of tertiary operations from a plant in Acheres, France.

**CLARIFIERS IN SERIES.** Table 5.22 contains a list of some existing full-scale wastewater treatment plants that operate with secondary clarifiers in series. Most are small and appear to have been modified from the original designs to allow operation in this mode. Virtually no data are available on the performance of clarifiers in series other than final effluent $BOD_5$ and suspended-solids concentrations. Although the concept seems unusual at first, anecdotal information suggests that clarifiers in series will result in suspended-solids reductions that are only slightly worse than many tertiary filters. Where the objective of tertiary treatment is incremental reductions in effluent suspended solids (e.g., suspended solids less than 5.0 mg/L), tertiary clarification could be a competitive alternative depending on the actual settling velocity distribution of the solids to be removed and the required

**Table 5.21** Acheres, France, operating data summary—tertiary operation.

| Month/ year | Average flow ($m^3/d$*) | No. trains | $FeCl_3$ | Polymer | TSS In | TSS Out | Total phosphorus In | Total phosphorus Out | $BOD_5$ In | $BOD_5$ Out |
|---|---|---|---|---|---|---|---|---|---|---|
| Apr-00 | 358 863 | 2.0 | 60 | 0.5 | 41 | 17 | 3.1 | 1.4 | 26.0 | 13.7 |
| Sep-00 | 317 000 | 1.2 | 37 | 0.5 | 29 | 9 | 3.7 | 1.4 | 24.0 | 10.0 |
| Oct-00 | 521 000 | 1.9 | 40 | 0.5 | 30 | 13 | 3.3 | 1.1 | 23.0 | 9.0 |
| Nov-00 | 444 000 | 1.6 | 40 | 0.5 | 31 | 11 | 3.3 | 0.9 | 23.0 | 9.0 |
| Aug-02 | 422 000 | 1.5 | 40.4 | 0.49 | 49 | 9 | 3.4 | 0.7 | 20.5 | 5.0 |
| Sep-02 | 334 000 | 1.4 | 42.7 | 0.54 | 47 | 13 | 3.8 | 1.4 | 32.0 | 11.0 |
| Oct-02 | 406 000 | 1.56 | 40.7 | 0.53 | 56 | 9 | 4.0 | 1.0 | 36.0 | 10.0 |

*$m^3/d \div 3785 = mgd$.

depth of the clarifiers. Seasonal dry or summertime use of clarifiers in series should be able to maximize performance of existing facilities without compromising facility capacity during cold or wet seasons. While the argument can be made that proper design of secondary clarifiers including adequate flocculation should result in effluent quality that is comparable to that obtained from existing clarifiers in series, full-scale experience suggests that a more detailed evaluation of existing clarifiers in series is warranted.

# CASE STUDIES

**ROCK CREEK ADVANCED WASTEWATER TREATMENT PLANT, HILLSBORO, OREGON.** Rock Creek Advanced Wastewater Treatment Plant (AWTP) is one of two large treatment plants on Oregon's Tualatin River owned and operated by Clean Water Services (formerly known as Unified Sewerage Agency of Washington County). Draining an area of approximately 1840 $k^2$ (710 sq mile) west of Portland, the Tualatin River flows east approximately 130 km (80 mile) from two reservoirs in the Coast Range Mountains to the Willamette River. During the summer, wastewater discharges can be as much as 25% of the river flow.

**Table 5.22** Existing facilities with series clarification.

| Plant | Design flow ($m^3/d$[b]) | Average flow ($m^3/d$) | Diameter (m) | Overflow rate (m/h) | First-stage effluent BOD (mg/L) | First-stage effluent TSS (mg/L) | Second-stage effluent BOD (mg/L) | Second-stage effluent TSS (mg/L) |
|---|---|---|---|---|---|---|---|---|
| Ash Flat, Arkansas | 290 | 160 | 5.5 | 0.51 | — | — | 2.8 | 2.5 |
| Cowan, Tennessee | 1500 | 1900 | 11.0 | 0.66 | — | — | 2.4 | 5.5 |
| Flippin, Arkansas | 660 | 590 | 5.5 | 1.16 | — | — | 1.8 | 3.1 |
| Holdingford, Minnesota | 450 | 450 | | | — | — | 2.2 | 3.8 |
| Davie, Florida | 1900 | 590 | 13.4 | 0.56 | — | — | 2.2 | 3.6 |
| Columbia, Kentucky | 2300 | 3000 | | | — | — | 3.0 | 2.4 |
| Randolph, Wisconsin | 3400 | 2600 | | | — | — | 5.8 | 4.3 |
| Shelton, Washington[a] | 16 200 | 7600 | 22.9 | 1.64 | — | 4.0 | — | 2.0 |
| Old Forge, New York | 600–1300 | 500–900 | 7.9 × 7.9 | 1.92 | — | — | 5–10 | 5–10 |

[a]Occasional series operation in summer time.

[b]$m^3/d \times 3785$ = gpd/sq ft.

Tertiary clarification and filtration was provided to meet the original permit, which required 75% removal of phosphorus. Problems with algal blooms with associated low dissolved oxygen and pH swings and with ammonia toxicity have historically occurred in the lower reaches of the Tualatin River in summer because of low river flows and a very low hydraulic gradient combined with inputs of ammonia and phosphorus from agricultural sources and municipal wastewater facilities. As a result of the water-quality problems and a court decision, the Oregon Department of Environmental Quality established a seasonal total maximum daily load for phosphorus of 0.07 mg/L in 1988. A renewal of the permit in 2004 altered the phosphorus limit from a mass-based limit to a concentration-based limit of 0.10 mg/L based on increased understanding of phosphorus loads from other sources in the watershed. The treatment plant's tertiary season is from approximately May 1 through October 31, which coincides with low river flows in the Pacific Northwest. Table 5.23 summarizes the original and current permit limits for the Rock Creek AWTP.

Influent to the Rock Creek plant includes typical municipal wastewater from residential, commercial, and light industrial sources; a dilute high-volume wastewater received from local industries year-round; and sludge, carrier water, and seasonal influent wastewater transfers from the Forest Grove and Hillsboro–Westside wastewater treatment plants. Current dry weather design capacity is 148 000 $m^3/d$ as compared to an average dry weather flow of 92 400 $m^3/d$ for the five years ending in 2002. Wet weather flows have averaged 148 000 $m^3/d$ over the last two years, with peak days of 500 000 $m^3/d$ (ODEQ, 2004).

**TABLE 5.23** Rock Creek monthly average discharge standards (ODEQ, 2003).

| | Original | | Current | |
|---|---|---|---|---|
| **Parameter** | **Summer (dry)** | **Winter (wet)** | **Summer (dry)** | **Winter (wet)** |
| $CBOD_5$*, mg/L | 20 | 30 | 8 | 20 |
| TSS, mg/L | 20 | 30 | 8 | 20 |
| Ammonia, mg/L as nitrogen | — | — | River flow and calendar based mass limits | |
| Total phosphorus, mg/L as phosphorus | 75% removal | | 0.10 | — |

*$CBOD_5$ = 5-day carbonaceous biochemical oxygen demand.

Originally constructed in 1977 as a 56 800-$m^3$/d tertiary facility, Rock Creek's treatment process included influent pumping, screening, grit removal, primary clarification, high-purity oxygen activated sludge, tertiary clarifiers, filters, and chlorination (Crom, 1977). The plant has been expanded and upgraded several times, including one project completed in 1993 that added a second train of tertiary clarification. An expansion completed in 2003 renovated the influent pumping station, added a new headworks and primary clarifiers, and made improvements to the digester complex and effluent filters. The plant currently has two liquid treatment trains, both using conventional activated sludge with anoxic selectors and fine-pore aeration. Waste activated sludge (WAS) was originally gravity thickened and then anaerobically digested. Primary sludge was thickened in the primary clarifier and then anaerobically digested. Currently, both the WAS and primary sludge are thickened using gravity belt thickeners followed by anaerobic digestion.

Phosphorus is now removed by two-stage alum addition in which approximately 20 mg/L of alum is added to the primary clarifiers followed by another 50 mg/L of alum in the secondary effluent. Tertiary clarifiers are used to settle the alum floc, and sand filters are used to reduce effluent suspended solids in the tertiary effluent. Alum sludge from the tertiary clarifiers is returned to the primaries to use any leftover alum. The chemical sludge is then processed with the primary sludge.

Tertiary clarification in the original 1970s facility was provided by two 33.5-m-diam conventional clarifiers with a side water depth of 3.75 m, rapid sludge removal mechanisms, and inboard double-weir launders. A typical section for the conventional tertiary clarifiers at Rock Creek is presented in Figure 5.17. The original tertiary clarifiers are still in service and are known as the West clarifiers. Separate rapid mix and flocculation precede the conventional tertiary clarifiers. The conventional tertiary clarifiers are nearly identical to the secondary clarifiers and, during wet weather, the tertiary clarifiers function as additional secondary clarifiers to provide a peak wet weather capacity of 170 000 $m^3$/d.

When the tertiary clarification process was expanded in the 1990s, four 18.3-m-diam upflow, "ClariCone" solids-contact clarifiers were constructed. A typical section for the solids-contact clarifiers is given in Figure 5.18. Flow is pumped from the secondary clarifiers to the solids-contact clarifiers where rapid mix and flocculation occur inside the vessel as the flow proceeds up through the clarifier. One of the primary drivers for selection of the ClariCone was limited site space. Less land was required using solids-contact clarifiers than more conventionally designed chemical clarifiers. Screw-induced flow centrifugal pumps were installed to lift the secondary effluent into the solids-contact clarifiers.

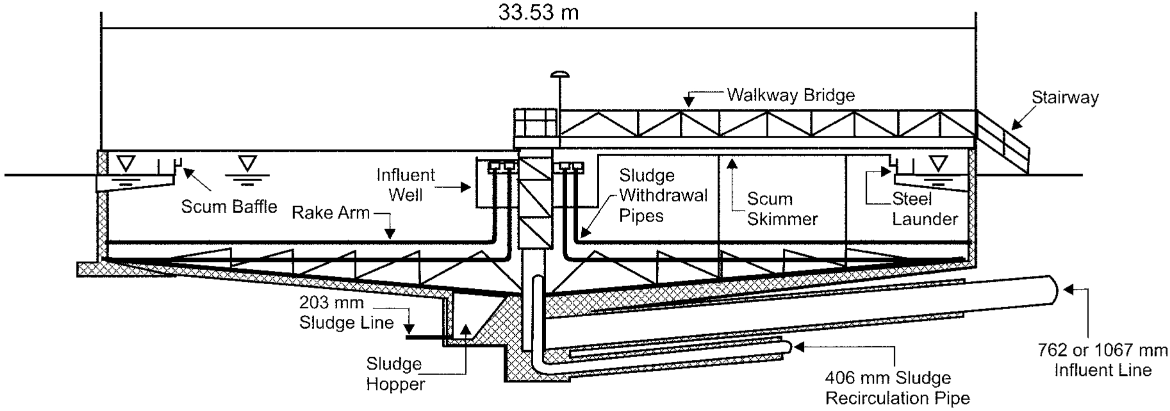


**FIGURE 5.17** Composite section of a conventional tertiary clarifier at Rock Creek AWTP.

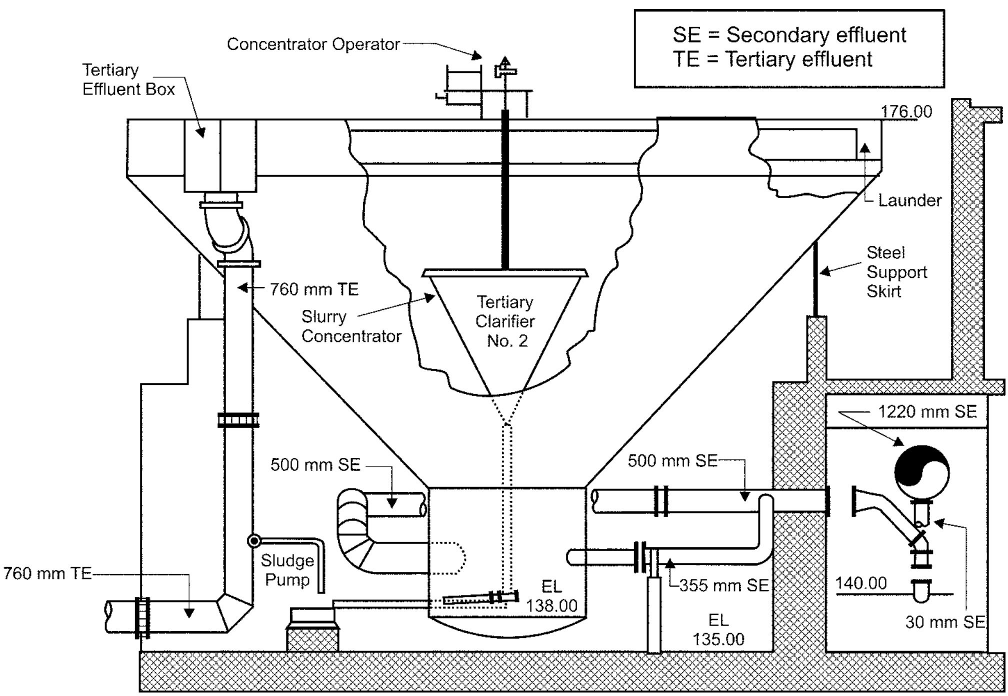


**FIGURE 5.18** Section of a solids-contact tertiary clarifier at Rock Creek AWTP.

Alum was retained as the primary coagulant after 1990 based on engineering studies that concluded that the effluent phosphorus could not be lowered significantly by switching to lime. It was also decided that alum was preferable to ferric chloride primarily based on lower cost, even though the use of ferric chloride would allow higher design hydraulic overflow rates. However, the aesthetic and safety concerns associated with handling ferric chloride were also a factor in the choice of coagulant. Multipoint chemical addition was implemented in the 1990 expansion. Piping was provided to allow alum feed to the primary clarifiers, activated sludge process, or secondary effluent before the tertiary clarifiers.

Design criteria are summarized in Table 5.24 for the original 1977 facility and the 1991 expansion project. Another expansion of tertiary treatment is now (2004) under

**TABLE 5.24** Tertiary clarifier design criteria for the Rock Creek AWTP.

| Item | Units | Conventional clarifiers (west) | Solids-contact clarifiers (east) |
|---|---|---|---|
| Design flows | | | |
| Average dry weather | $m^3/d$ | 65 900 | 60 560 |
| Maximum month dry weather | $m^3/d$ | 72 300 | 75 700 |
| Maximum month wet weather | $m^3/d$ | NA* | NA |
| Peak day | $m^3/d$ | NA | NA |
| Number units | | 2 | 4 |
| Diameter | m | 33.5 | 18.3 |
| Side water depth | m | 3.75 | 8.5 |
| Overflow rate (at indicated flow) | m/h ($m^3/d$) | 1.00 (21 600 $m^3/d$)<br>1.25 (26 500 $m^3/d$) | 2.4 (15 100 $m^3/d$) |
| Alum dose | mg/L | 50 | 50 |
| Polymer dose | mg/L | 0.5 | 0.5 |
| Rapid mix | | | |
| Detention time (average day dry weather) | s | 45 | — |
| Velocity gradient | $s^{-1}$ | 425 | — |
| Flocculation | | | |
| Detention time (average day dry weather) | min | 20 | — |
| Velocity gradient | $s^{-1}$ | 60 | — |

*NA = not available.

construction. After the 2004 expansion, the existing tertiary clarifiers will be base-loaded and direct filtration following alum addition to the secondary clarifier effluent will be used to treat the balance of flow.

Results of a six-month performance evaluation conducted after startup of the solids-contact tertiary clarifiers are summarized in Table 5.25. Within four months of startup, the solids-contact units were close to meeting the effluent total phosphorus limit of 0.07 mg/L. Performance has improved over time and both the solids-contact and conventional clarifiers consistently produce less than 0.07 mg/L after filtration.

Under the current operating strategy, the solids-contact clarifiers are base-loaded at a peak flow of approximately 75 700 $m^3$ (20 mgd) (68 130 $m^3$ average [18 mgd average]) to facilitate operations and maximize phosphorus removal. Experience at Rock Creek has shown that the solids-contact clarifiers typically produce marginally lower effluent total phosphorus concentrations but operate best over a limited range of flows. At more than approximately 83 300 $m^3$ (22 mgd) and less than 19 000 $m^3$ (5 mgd), maintaining the sludge blanket is problematic. All flow more than 75 700 $m^3$ (20 mgd) is directed to the conventional tertiary clarifiers.

Monthly average performance data for 2002 and 2003 are summarized in Tables 5.26 and 5.27. All tertiary clarifiers are in operation during the summer season for phosphorus, and approximately 1.0 to 1.5 mg/L of polymer is added to

**TABLE 5.25** Summary of Rock Creek AWTP phosphorus-removal demonstration (Daigger and Butz, 1992).

| | | Concentration (mg/L) | | | | | | |
|---|---|---|---|---|---|---|---|---|
| Date | Influent flow ($m^3$/d) | Influent TP* | Primary alum dose | Primary effluent TP | Secondary effluent TP | Tertiary alum dose | Final effluent TP |
| May 1990 | 59 000 | 9.3 | 124 | 3.9 | 1.52 | 0 | 1.21 |
| June 1990 | 57 000 | 9.9 | 122 | 3.0 | 0.01 | 0 | 0.80 |
| July 1990 | 55 000 | 9.9 | 113 | 3.0 | 0.73 | 18 | 0.55 |
| August 1990 | 52 000 | 10.3 | 120 | 2.8 | 0.61 | 48 | 0.09 |
| September 1990 | 48 000 | 9.7 | 127 | 1.9 | 0.41 | 60 | 0.07 |
| October 1990 | 52 000 | 9.1 | 109 | 2.3 | 0.44 | 53 | 0.09 |
| Average | 54 000 | 9.7 | 119 | 2.8 | 0.75 | 30 | 0.47 |

*TP = total phosphorus.

**Table 5.26** Monthly average performance for the Rock Creek solids-contact tertiary clarifiers (ClariCone®) for 2002 and 2003.

| Month/ Year | Flow ($m^3$/d) | Overflow rate (m/h) | Alum dose (mg/L) | Influent TP[a] (mg/L) | Influent OP[b] (mg/L) | Influent TSS (mg/L) | Clarifier effluent TP (mg/L) | Clarifier effluent OP (mg/L) | Clarifier effluent TSS (mg/L) | Filter effluent TP (mg/L) | Filter effluent OP (mg/L) | Filter effluent TSS (mg/L) |
|---|---|---|---|---|---|---|---|---|---|---|---|---|
| **2002** | | | | | | | | | | | | |
| May | 68 800 | 2.7 | | 0.62 | 0.28 | 5.4 | 0.15 | 0.01 | 9.3 | 0.04 | 0.01 | 1.1 |
| Jun | 64 800 | 2.6 | 49 | 0.38 | 0.22 | 4.7 | 0.08 | 0.01 | 5.6 | 0.03 | 0.02 | 0.5 |
| Jul | 67 800 | 2.7 | 45 | 0.90 | 0.73 | 6.1 | 0.15 | 0.02 | 4.6 | 0.07 | 0.02 | 0.8 |
| Aug | 68 500 | 2.7 | 47 | 1.25 | 0.96 | 7.7 | 0.20 | 0.03 | 5.4 | 0.10 | 0.03 | 1.4 |
| Sep | 68 300 | 2.7 | 40 | 0.42 | 0.28 | 4.3 | 0.10 | 0.03 | 5.1 | 0.04 | 0.02 | 0.7 |
| Oct | 67 100 | 2.7 | 39 | 0.65 | 0.58 | 3.2 | 0.12 | 0.05 | 4.9 | 0.07 | 0.05 | 0.4 |
| Min | 64 800 | 2.6 | 39 | 0.38 | 0.22 | 3.2 | 0.08 | 0.01 | 4.6 | 0.03 | 0.01 | 0.4 |
| Max | 68 800 | 2.7 | 49 | 1.25 | 0.96 | 7.7 | 0.20 | 0.05 | 9.3 | 0.10 | 0.05 | 1.4 |
| Ave | 67 600 | 2.7 | 44 | 0.70 | 0.51 | 5.2 | 0.13 | 0.03 | 5.8 | 0.06 | 0.03 | 0.8 |
| **2003** | | | | | | | | | | | | |
| May | 72 500 | 2.9 | 45 | 0.57 | 0.46 | 4 | 0.18 | 0.02 | 7.4 | 0.06 | 0.03 | 3.1 |
| Jun | 71 000 | 2.8 | 41 | 0.34 | 0.23 | 4 | 0.12 | 0.02 | 6.2 | 0.05 | 0.02 | 1.1 |
| Jul | 72 100 | 2.9 | 44 | 0.31 | 0.16 | 7 | 0.07 | 0.02 | 4.4 | 0.04 | 0.01 | 1.0 |
| Aug | 71 100 | 2.8 | 41 | 0.62 | 0.43 | 5 | 0.12 | 0.02 | 4.5 | 0.07 | 0.02 | 1.2 |
| Sep | 70 400 | 2.8 | 45 | 0.34 | 0.22 | 3 | 0.06 | 0.01 | 4.0 | 0.03 | 0.01 | 0.9 |
| Oct | 71 600 | 2.8 | 41 | 0.48 | 0.34 | 4 | 0.09 | 0.01 | 3.9 | 0.04 | 0.01 | 1.0 |
| Min | 70 400 | 2.8 | 41 | 0.31 | 0.16 | 3 | 0.06 | 0.01 | 3.9 | 0.03 | 0.01 | 0.9 |
| Max | 72 500 | 2.9 | 45 | 0.62 | 0.46 | 7 | 0.18 | 0.02 | 7.4 | 0.07 | 0.03 | 3.1 |
| Ave | 71 500 | 2.8 | 43 | 0.44 | 0.31 | 5 | 0.11 | 0.02 | 5.0 | 0.05 | 0.02 | 1.4 |

[a]TP = total phosphorus.

[b]OP = orthophosphorus.

**TABLE 5.27** Monthly average performance for the Rock Creek conventional tertiary clarifiers for 2002 and 2003.

| Month/ Year | Flow ($m^3$/d) | Overflow rate (m/h) | Alum dose (mg/L) | Influent TP[a] (mg/L) | Influent OP[b] (mg/L) | Influent TSS (mg/L) | Clarifier effluent TP (mg/L) | Clarifier effluent OP (mg/L) | Clarifier effluent TSS (mg/L) | Filter effluent TP (mg/L) | Filter effluent OP (mg/L) | Filter effluent TSS (mg/L) |
|---|---|---|---|---|---|---|---|---|---|---|---|---|
| **2002** | | | | | | | | | | | | |
| May | 51 300 | 1.2 | 58 | 0.86 | 0.64 | 6.3 | 0.40 | 0.03 | 11.7 | 0.17 | 0.05 | 3.2 |
| Jun | 57 400 | 1.4 | 48 | 0.47 | 0.22 | 10.6 | 0.22 | 0.01 | 11.7 | 0.07 | 0.02 | 3.4 |
| Jul | 45 400 | 1.1 | 43 | 1.14 | 0.86 | 9.2 | 0.37 | 0.04 | 8.4 | 0.13 | 0.06 | 1.9 |
| Aug | 38 200 | 0.9 | 23 | 0.56 | 0.13 | 12.4 | 0.17 | 0.01 | 5.3 | 0.10 | 0.02 | 2.9 |
| Sep | 34 500 | 0.8 | 42 | 0.63 | 0.38 | 9.0 | 0.17 | 0.03 | 6.1 | 0.06 | 0.04 | 1.7 |
| Oct | 32 100 | 0.8 | 43 | 1.31 | 1.18 | 5.9 | 0.24 | 0.08 | 4.8 | 0.10 | 0.08 | 1.4 |
| Min | 32 100 | 0.8 | 23 | 0.47 | 0.13 | 5.9 | 0.17 | 0.01 | 4.8 | 0.06 | 0.02 | 1.4 |
| Max | 57 400 | 1.4 | 58 | 1.31 | 1.18 | 12.4 | 0.40 | 0.08 | 11.7 | 0.17 | 0.08 | 3.4 |
| Ave | 43 100 | 1.0 | 43 | 0.83 | 0.57 | 8.8 | 0.26 | 0.03 | 8.0 | 0.11 | 0.05 | 2.4 |
| **2003** | | | | | | | | | | | | |
| May | 59 000 | 1.4 | 22 | 0.43 | 0.08 | 9 | 0.21 | 0.02 | 9.3 | 0.08 | 0.02 | 2.5 |
| Jun | 43 500 | 1.0 | 33 | 0.39 | 0.13 | 8 | 0.16 | 0.02 | 7.4 | 0.08 | 0.02 | 2.6 |
| Jul | 48 400 | 1.1 | 27 | 0.31 | 0.09 | 7 | 0.16 | 0.02 | 8.1 | 0.07 | 0.03 | 2.1 |
| Aug | 46 300 | 1.1 | 37 | 0.49 | 0.13 | 10 | 0.19 | 0.02 | 8.0 | 0.09 | 0.03 | 1.7 |
| Sep | 46 800 | 1.1 | 35 | 0.28 | 0.03 | 7 | 0.09 | 0.01 | 6.1 | 0.04 | 0.01 | 1.1 |
| Oct | 49 000 | 1.2 | 29 | 0.81 | 0.28 | 15 | 0.26 | 0.06 | 6.0 | 0.13 | 0.06 | 2.0 |
| Min | 43 500 | 1.0 | 22 | 0.28 | 0.03 | 7 | 0.09 | 0.01 | 6.0 | 0.04 | 0.01 | 1.1 |
| Max | 59 000 | 1.4 | 37 | 0.81 | 0.28 | 15 | 0.26 | 0.06 | 9.3 | 0.13 | 0.06 | 2.6 |
| Ave | 48 800 | 1.2 | 31 | 0.45 | 0.12 | 9 | 0.18 | 0.03 | 7.3 | 0.08 | 0.03 | 2.0 |

[a]TP = total phosphorus.
[b]OP = orthophosphorus.

both sets of clarifiers as a settling aid. The solids-contact clarifiers typically perform better than the conventional units in terms of hydraulic overflow rate and total phosphorus. Overflow rates for the solids-contact clarifiers are significantly higher for the conventional units, with an average of approximately 2.7 to 2.8 m/h compared with 1.0 to 1.2 m/h for the conventional clarifiers. Effluent total phosphorus is slightly lower from the solids-contact clarifiers, with an average of approximately 0.11 to 0.13 mg/L. In comparison, the conventional units averaged approximately 0.18 to 0.26 mg/L total phosphorus. Total suspended solids in the effluent from both sets of clarifiers is approximately the same or slightly higher than the secondary effluent TSS. Even for TSS, the solids-contact units more often show slightly better performance and more frequently show a reduction in TSS across the clarifiers. Effluent TSS from the solids-contact clarifiers averaged approximately 5 mg/L and ranged from approximately 4 to 10 mg/L, whereas the TSS from the conventional clarifiers averaged approximately 7 to 8 mg/L and ranged from 5 to 12 mg/L.

**WATER FACTORY 21, FOUNTAIN VALLEY, CALIFORNIA.** Orange County Water District (OCWD) manages the groundwater basin underneath the western one-half of Orange County, California. To protect and sustain the aquifer as a water supply, OCWD has recycled reclaimed water into the aquifer since 1976. A maximum of 56 775 $m^3$ (15 mgd) of treated wastewater from Water Factory 21 blended with 32 550 $m^3$ (8.6 mgd) of deep well water is injected to the Talbert gap to create a hydraulic barrier to seawater intrusion. In 2003, the original Water Factory 21 treatment plant was decommissioned to make way for the new and significantly larger groundwater replenishment system project that was constructed on the same site. Water Factory 21 used a combination of chemical clarification, recarbonation, multimedia filtration, granular activated carbon (GAC), reverse osmosis, and chlorination to treat secondary effluent from the adjacent OCSD Plant No. 1. Lime sludge was thickened, recalcined in a multiple-hearth furnace, and reused. Chemical clarification at Water Factory 21 was a key component of the pretreatment process before GAC and revesre osmosis. Clarified and filtered water provided feedwater to the 34 065-$m^3$ (9-mgd) GAC system and to the 18 925-$m^3$ (5-mgd) reverse osmosis process. Tertiary treatment was used to reduce organic and inorganic constituents in the secondary effluent that foul reverse osmosis membranes. Successful pretreatment maximized the performance (reverse osmosis flux, cleaning frequency, and recovery) and the useful life of the reverse osmosis membranes.

The chemical clarification system comprised separate rapid mix, flocculation, and gravity settling. A lime dose of 375 to 500 mg/L was used to coagulate the secondary effluent and raised the pH to 11.3. Approximately 0.1 mg/L of anionic polymer was added to the third stage of the flocculation tank as a settling aide. Hydraulic detention times were 1 minute in rapid mix, 30 minutes in flocculation, and 85 minutes in settling. The lime clarification process was designed for an overflow rate of 1.6 m/h. In 1975, the capital cost for the reclamation facilities (chemical clarification, recarbonation, filtration, chlorination, and associated systems) was $13,400,000. Operating costs for the advanced water treatment (AWT) plant were approximately $0.27/m$^3$ ($1.01/1000 gal) in 1996. Design criteria for the tertiary clarification facilities at Water Factory 21 are summarized in Table 5.28 and simplified plan and section drawings are shown in Figures 5.19 and 5.20.

During most recent years before the plant was decommissioned in 2004, the Water Factory 21 tertiary treatment process did not operate at capacity. Average monthly flows to the lime clarification process in 1988/1989 were 11 100 to 46 200 m$^3$/d, whereas in 1998 they were only 5700 to 21 200 m$^3$/d. Overflow rates ranged from approximately 0.5 to 2.2 m/h in 1988/1989 and 0.3 to 1.0 m/h in 1998. Typical performance data from two years (June 1988 to July 1989 and 1998) are presented in Table 5.29. Lime clarification at Water Factory 21 typically provided a 2 log removal for both total and fecal coliform and was credited with a 2 to 3 log virus removal. Lime clarification effluent typically contained less than 1 total coliform per 100 mL, and total organic carbon (TOC) removal averaged approximately 26% of the TOC, which resulted in a typical effluent TOC concentration of 7.6 mg/L. In 1998, turbidity was reduced approximately 84% and TOC was reduced approximately 35%. Lime clarification also reduced the concentrations of hardness, alkalinity, silica, and metals. Phosphates were typically reduced to approximately 0.1 mg/L.

Silt density index values of the filtered, chlorinated tertiary effluent typically range from approximately 4.5 to greater than 6.6. Although lime clarification and multimedia filters allowed the downstream reverse osmosis system to operate reasonably well, cleaning intervals were short at approximately three weeks. The clarifier effluent turbidity averaged 1.1 nephelometric turbidity units (NTU). A virus-monitoring program was conducted from 1975 to 1981. No virus was ever detected after lime clarification.

Water Factory 21 was originally designed with two-stage lime clarification for additional total dissolved solids (TDS) reduction. The recarbonation basin was divided into three compartments, with the first and third used for the addition of

**TABLE 5.28** Water Factory 21 tertiary clarifier design criteria (Argo and Moutes, 1979).

| Item | Units | Tertiary treatment |
|---|---|---|
| Design flows | $m^3/d$ | 56 800 |
| Rapid mix | | |
| No. units | | 2 in series |
| Type mixing | | Mechanical |
| Width | m | 3.66 |
| Length | m | 3.66 |
| Side water depth | m | 3.66 |
| Detention time | min | |
| Flocculation | | |
| No. units | | 2 (each 3 cells) |
| Type mixing | | Variable-speed oscillating |
| Width | m | 14.6 |
| Length | m | 12.5 |
| Side water dept | m | 3.4 |
| Detention time | min | 30 (10/cell) |
| Settling basins | | |
| Number units | | 2 |
| Length | m | 36.6 |
| Width | m | 12.2 |
| Side water depth | m | |
| Surface area | $m^2$ | 446 |
| Overflow rate | m/h | 2.6 |
| Lime dose | mg/L (as CaO*) | 350–500 |
| Clarifier pH | | 11.3 |
| Polymer type | | Anionic |
| Polymer dose | mg/L | |
| Recarbonation | | |
| Number units | | 2 (3 cells: recarbonation/settling/recarbonation) |
| Detention time | min | 15 each (1st and 3rd cells) |
| Overflow rate | m/h | 5.1 |

*CaO = calcium oxide.

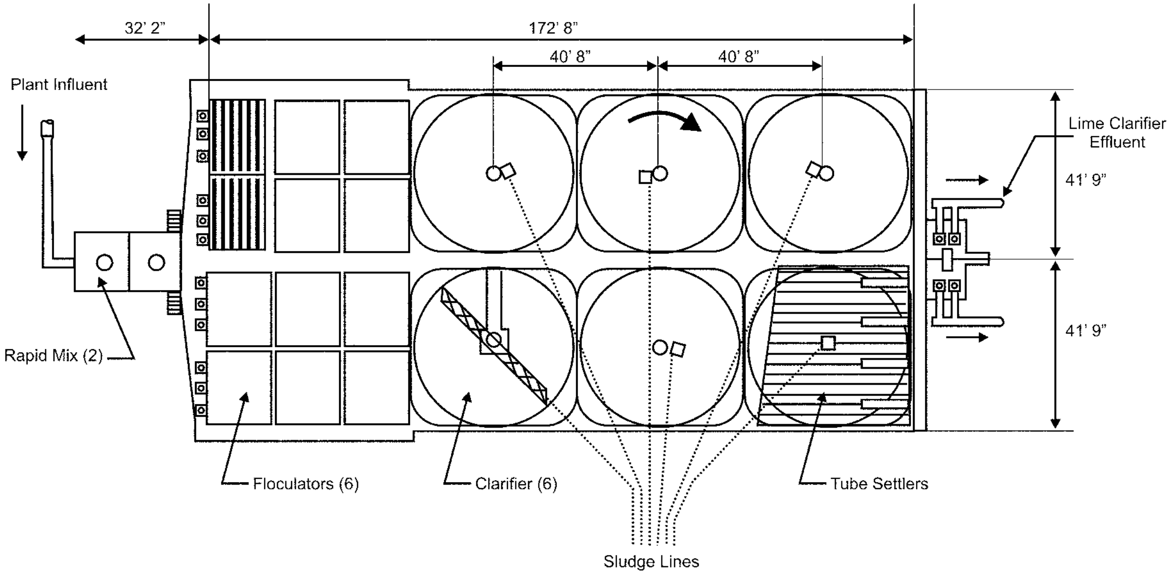


**FIGURE 5.19** Lime rapid mix, flocculation, and clarifier plan view at Water Factory 21 (ft × 0.304 8 = m; in. × 25.4 mm).

carbon dioxide from stack gases from the lime furnace or purchased gas when the furnace was down. Intermediate settling was to occur in the middle basin; however, experience showed that TDS was reduced only approximately 5% and calcium carbonate precipitate carried over to the multimedia filters. As a result, intermediate clarification was abandoned, and the plant was operated as a single-stage process.

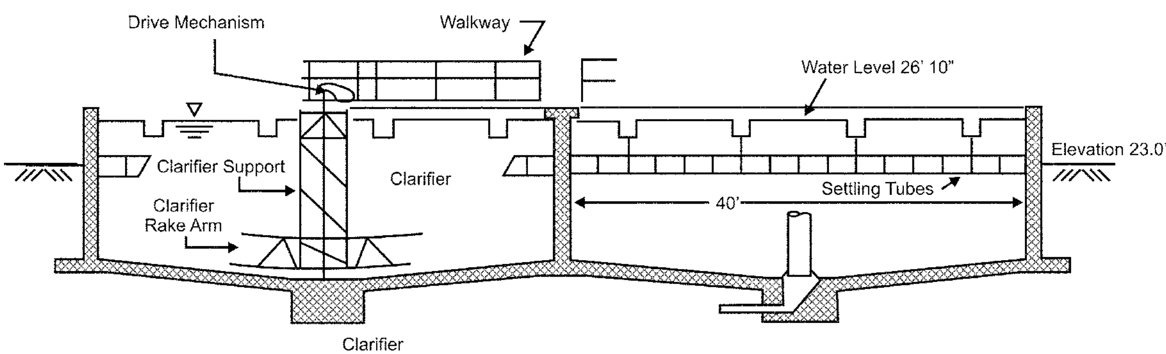


**FIGURE 5.20** Lime clarifier section at Water Factory 21 (ft × 0.304 8 = m; in. × 25.4 = mm).

**TABLE 5.29** Tertiary clarifier performance data at Water Factory 21.

| | | July 1988–June 1989[b] | | 1998 | |
|---|---|---|---|---|---|
| Constituent[a] | Units | Influent | Clarifier effluent | Influent | Clarifier effluent |
| Tertiary flow | $m^3/d$ | 11 100–46 200 | | 5700–21 200 | |
| Conductivity | μmhos/cm | | | 1658 | 1656 |
| TDS | mg/L | 940 | 1010 | 959 | 858 |
| pH | units | | | 7.4 | 11.1 |
| Magnesium | mg/L | 23 | 2 | 25 | 1 |
| Calcium | mg/L | 86 | 90 | 82 | 95 |
| Iron | μg/L | | | 281.4 | 8.7 |
| Manganese | μg/L | 56 | 1.3 | 54.4 | 0.5 |
| Arsenic | μg/L | <5 | <5 | 0.2 | 0.0 |
| Barium | μg/L | 94 | 35 | 35.6 | 19.7 |
| Chromium | μg/L | 33 | 8 | 1.9 | 0.4 |
| Copper | μg/L | 49 | 9 | 14.1 | 5.9 |
| Nickel | μg/L | 77 | 27 | 18.8 | 13.2 |
| Lead | μg/L | 5 | 0.5 | 0.7 | 0.1 |
| Total hardness | mg/L | | | 308 | 241 |
| Total alkalinity | mg/L | 204 | 253 | 256 | 174 |
| Fluoride | mg/L | 1.3 | 0.9 | 1.10 | 0.47 |
| Boron | mg/L | 0.79 | 0.63 | 0.50 | 0.38 |
| Silica | mg/L | | | 21.7 | 11.0 |
| TOC | mg/L | 14 | 11 | 10.95 | 7.06 |
| Color | units | | | 29 | 20 |
| Total coliform | CFU/100 mL | 420 000 | 0.3 | 660 000 | <1 |
| Fecal coliform | CFU/100 mL | 43 000 | 0.03 | 112 000 | <1 |

[a]TDS = total dissolved solids and TOC = total organic carbon.
[b]Metals data is the average for the period January 1980 through March 1981.

**UPPER OCCOQUAN SEWAGE AUTHORITY WATER RECLAMATION PLANT, CENTREVILLE, VIRGINIA.** In response to very rapid urbanization of the Upper Occoquan Watershed in the 1960s, a 56 775-$m^3$ (15-mgd) regional advanced water reclamation plant was constructed to replace eleven small secondary treatment plants. Urbanization of the watershed contributed to a number of severe water-quality problems in the Occoquan reservoir that receives the discharge from the plant. In the 1960s, the problems included frequent algae blooms from excessive nutrients and fish kills from reverse osmosis depletion. Construction of the Upper Occoquan Sewage Authority (UOSA) plant was intended to consolidate wastewater treatment in the watershed and eliminate water-quality problems in the reservoir. Effluent quality criteria for the UOSA plant are summarized in Table 5.30.

The original sequence of treatment processes at the UOSA plant included primary treatment, secondary treatment, high lime flocculation and clarification, recarbonation, filtration, granular activated carbon absorption, ion exchange, chlorination, and discharge to a final effluent reservoir. In the early 1980s, use of clinoptilolite ion-exchange beds for tertiary ammonia removal was discontinued and the plant was operated in a nitrifying mode. Subsequently, the clinoptilolite was removed and replaced with sand filter media to ensure reliable removal of carbon fines and other residual suspended solids.

**TABLE 5.30** Upper Occoquan Sewage Authority plant effluent limits (McEwen, 1998).

| Parameter | Units | Concentration |
|---|---|---|
| COD | mg/L | 10.0 |
| Suspended solids | mg/L | 1.0 |
| Total phosphorus | mg/L | 0.1 |
| Unoxidized nitrogen[a] | mg/L | 1.0 |
| MBAS[b] | mg/L | 0.1 |
| Turbidity | NTU | 0.5 |
| Coliform bacteria | Per 100 mL | < 2 |

[a]Process to be operated to maintain less than 5 mg/L nitrate nitrogen in Occoquan Reservoir.
[b]MBAS = methylene blue active substances.

After a series of intermediate expansions, the plant underwent expansion to 204 390 $m^3$ (54 mgd) (1997–-2004), in which the biological process was expanded and upgraded to provide full-time, improved denitrification facilities; the pressure filters were supplemented with gravity filters; and the single-stage downflow carbon columns were supplemented with two-stage upflow/downflow carbon columns. Expansion of the liquid treatment facilities roughly mirrored the original design except for changes in the number of units, design improvements intended to improve operating efficiency, and the significant changes noted previously. Typical performance for the entire UOSA facility is detailed in Table 5.31.

Tertiary lime clarification of the secondary effluent is provided primarily to meet the effluent phosphorus limit of 0.1 mg/L. This was an exceptionally low limit at the time that the plant was constructed and, even today, there are few plants with limits less than 0.1 mg/L. A traditional two-stage lime clarification sequence is used as

**TABLE 5.31** Typical UOSA WRP performance (McEwen, 1998).

| Location | COD (mg/L) | TSS (mg/L) | TKN[a] (mg/L) | TP[b] (mg/L) | MBAS[c] (mg/L) | Turbitity (NTU) |
|---|---|---|---|---|---|---|
| Plant influent | 400 | 170 | 34 | 9.0 | 5.3 | — |
| Primary effluent | 240 | 65 | 29 | 7.2 | — | — |
| Secondary effluent | 40 | 16 | 2.5 | 5.9 | — | 4.5 |
| Chemical clarifier | 18 | 11 | 1.2 | 0.15 | — | 1.5 |
| Second-stage recarbonation | 17 | 10 | 1.2 | 0.09 | — | 0.85 |
| Filter effluent | 13 | 0.2 | 0.75 | 0.03 | 0.08 | 0.3 |
| Carbon effluent | 11 | 1.0 | 0.6 | 0.03 | 0.05 | 0.6 |
| Postcarbon filtration | — | 0.3 | 0.55 | 0.03 | — | 0.25 |
| Final effluent | 8 | 0.1 | 0.46 | 0.03 | 0.04 | 0.25 |
| Effluent limits | 10 | 1 | 1 | 0.10 | 0.10 | 0.5 |

[a]TKN = total Kjeldahl nitrogen.
[b]TP = total phosphorus.
[c]MBAS = methylene blue active substances.

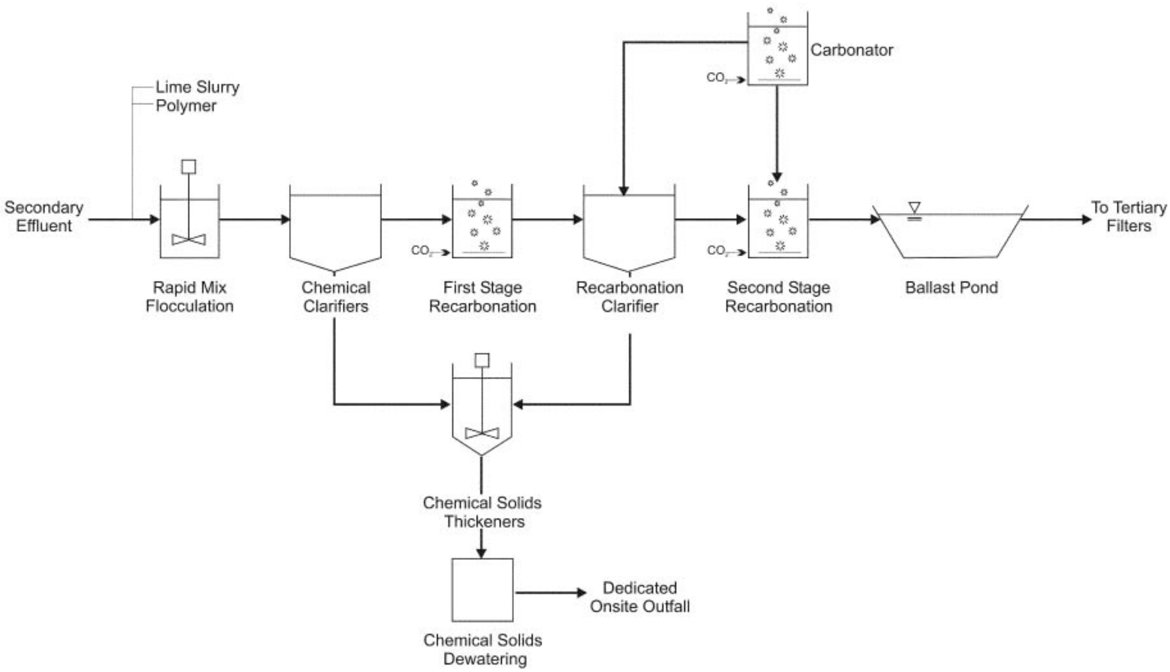


**FIGURE 5.21** Process schematic for lime clarification process at UOSA.

shown in the simplified partial process flow diagram in Figure 5.21. Chemical treatment includes rapid mixing, flocculation, first-stage settling, and two-stage recarbonation with intermediate settling. All the tertiary clarifiers are circular center-feed units with hydraulic overflow rates of approximately 1.5 m/h at design 30-day flows. Detailed design criteria for the lime clarification process are provided in Table 5.32, and simplified plan and section drawings for the tertiary clarifiers are provided in Figures 5.22 and 5.23.

Slaked calcium oxide is added as the primary coagulant and to raise the pH to approximately 11.0 to 11.2. An anionic polyelectrolyte is added in the flocculation basins to enhance settling. After the chemical clarifiers, sufficient carbon dioxide is added to lower the pH to approximately 9.5 to 10 and then flow is directed to the recarbonation clarifiers to precipitate additional sludge. Additional carbon dioxide is added following the second-stage clarifiers to lower the pH to approximately 7.5 to 8.0. Changes in pH and alkalinity through the tertiary clarification process in 2003 are summarized in Table 5.33.

**Table 5.32** Design criteria for the UOSA two-stage lime clarification process.

| Unit process | Units | Original | Expansion |
|---|---|---|---|
| Flows | | | |
| Average | $m^3/d$ | 71 600 | 87 500 |
| Maximum 30 days | $m^3/d$ | 92 000 | 112 500 |
| Maximum 7 days | $m^3/d$ | 127 900 | 156 250 |
| Flow split | | 45 | 55 |
| Rapid mix basins | | | |
| Number | | 2 | 4 |
| Length | m | 6.4 | 4.9 |
| Width | m | 6.4 | 6.1 |
| SWD[a] | m | 4.3 | 3.7 |
| Detention time at QMM[b] | min | 4.9 | 6.1 |
| Mixer motor power | kW | 29.8 | 22.4 |
| Flocculation basins | | | |
| Number | | | |
| Length | m | 14.3 | 14.3 |
| Width | m | 6.1 | 6.1 |
| SWD | m | 4.3 | 4.3 |
| Detention time at QMM | min | 16 | 21 |
| Mixer motor power | kW | 1 at 14.9<br>1 at 3.7 | 1 at 14.9<br>1 at 3.7 |
| Chemical clarifiers | | | |
| Number | | 3 | 4 |
| Diameter | m | 38.1 | 38.1 |
| SWD | m | 3.7 | 3.7 |
| Surface overflow rate | m/h | 1.2 | 0.9 |
| First-stage recarbonation basins | | | |
| Number | | 2 | 4 |
| Length | m | 9.1 | 13.4 |
| Width | m | 4.6 | 6.7 |
| SWD | m | 4.3 | 4.9 |
| Detention time at QMM | min | 5 | 25 |
| Recarbonation clarifiers | | | |
| Number | | 3 | 4 |
| Diameter | m | 38.1 | 38.1 |
| SWD | m | 3.7 | 4.5 |
| Surface overflow rate | m/h | 1.2 | 0.9 |
| Second-stage recarbonation basins | | | |
| Number | | | |
| Length | m | 9.1 | 13.4 |
| Width | m | 4.6 | 6.7 |
| SWD | m | 4.3 | 4.9 |
| Detention time at QMM | min | 5 | 25 |

[a]SWD = side water depth.

[b]QMM = maximum month flow.

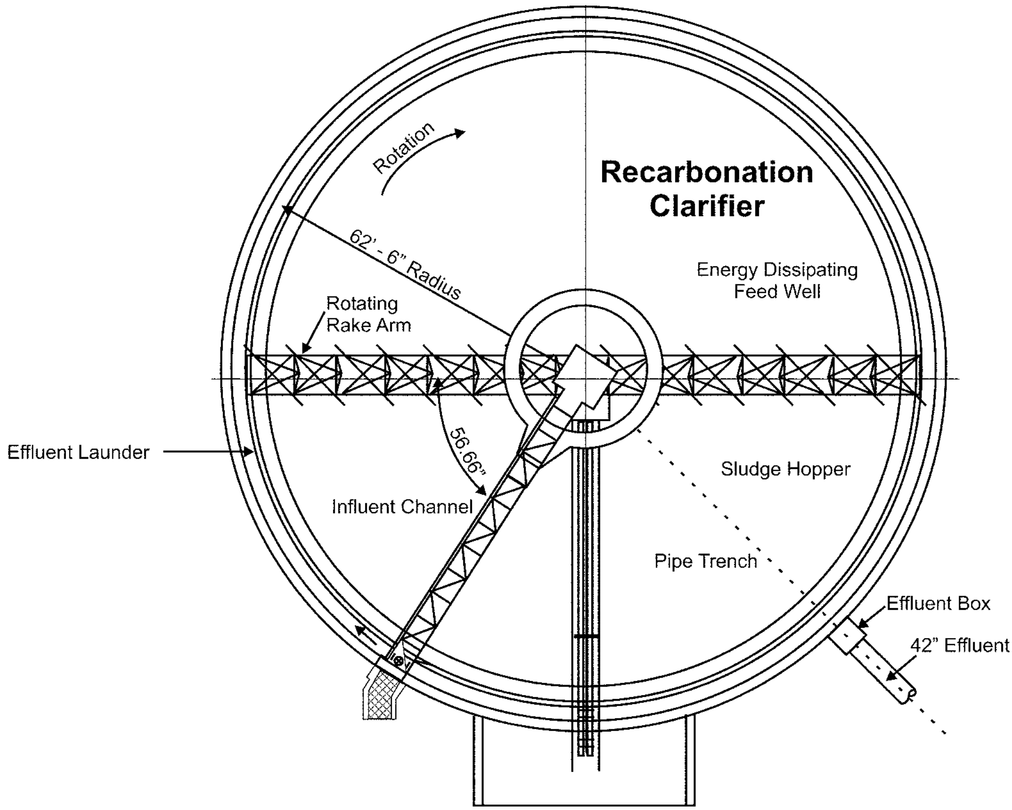


**FIGURE 5.22** Typical plan for recarbonation clarifier at UOSA (ft × 0.304 8 = m; in. × 25.4 = mm).

Initially, plant staff operated the chemical clarifiers at a pH of approximately 11.3 to be certain to meet the total phosphorus limit; however, with experience plant staff have become comfortable operating at slightly lower pH values while still meeting the total phosphorus permit limit. Operating at lower pH values reduces costs for lime, carbon dioxide, and sludge handling, although phosphorus concentrations from the lime clarification process are slightly higher, requiring polishing with alum before final filtration. Carbon dioxide for recarbonation is obtained from two sources—combustion of anaerobic digester gas and commercial liquefied carbon dioxide gas (LCG). During low-flow periods, sufficient carbon dioxide is available from burning digester gas. Under most flow conditions, LCG must be used to supplement the digester gas. Over time, the plant has been able to reduce the amount of LCG from approximately 100 mg/L in 1991 to approximately 40 to 50 mg/L in 2004 by improving the combusted biogas yield and the efficiency of the gas-transfer system. Typical lime doses are approximately 140 to 150 mg/L as $CaCO_3$. Although 0.4 mg/L anionic polymer is added year-round to the flocculation basins as a settling aid, experience has shown that polymer is most necessary during the winter. Following the first-stage chemical clarifiers, the pH is lowered to 9.5 to 10.0 in the first-stage recarbonation process and to 7.0 in the second-stage recarbonation process. Flow equalization is provided after recarbonation and before GAC absorption, filtration, and disinfection. Alum is added before the final sand filters to drop the effluent total phosphorus from approximately 0.2 mg/L, leaving the second-stage recarbonation down to less than 0.10 mg/L in the filter effluent.

Scaling is a significant operational problem throughout the UOSA lime clarification process. Thick, dense scale rapidly accumulates on all surfaces that are in contact with the water following lime addition. Scale in the chemical clarification section

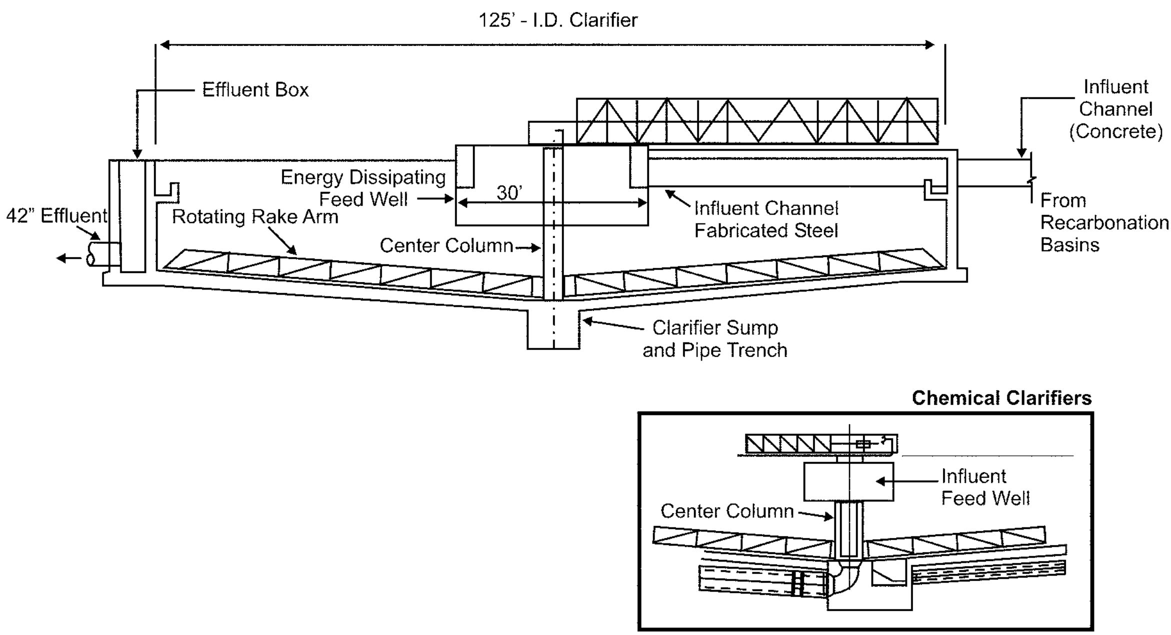


**FIGURE 5.23** Typical section for recarbonation clarifiers at UOSA (ft × 0.304 8 = m; in. × 25.4 = mm).

**Table 5.33** Summary of daily pH and alkalinity values through the UOSA tertiary lime clarification process in 2003.

| | pH | | | Alkalinity (mg/L $CaCO_3$) | | |
|---|---|---|---|---|---|---|
| **Location in process** | **Average** | **10th percentile** | **90th percentile** | **Average** | **10th percentile** | **90th percentile** |
| Secondary effluent | 6.5 | 6.3 | 6.7 | 59 | 45 | 76 |
| Rapid mix | 10.9 | 10.6 | 11.2 | 184 | 121 | 254 |
| First-stage clarifier effluent | 11.1 | 10.8 | 11.4 | 155 | 118 | 200 |
| First-stage recarbonation effluent | 10.3 | 10.0 | 10.5 | | | |
| Recarbonation clarifier effluent | 10.2 | 9.9 | 10.4 | 94 | 71 | 117 |
| Second-stage recarbonation effluent | 8.0 | 7.5 | 8.4 | 96 | 76 | 115 |

is typically denser and more difficult to remove than scale in the recarbonation section. Constant descaling is required to maintain the capacity of pipes, valves, and orifices. Depending on flowrates and the ability to remove units from service, plant staff have dedicated one or two full-time crews whose sole function is to remove accumulated scale. Various methods of descaling have been used, including hand chisels, and acid. Hand chiseling was very labor intensive and not all areas could be cleaned, wherease acid flushes did not provide uniform removal of scale. Currently, high-pressure water jets and vacuum jets are the preferred method for removing scale. Design features were included in the recent expansion to facilitate regular scale removal, including the use of channels instead of pipes whenever possible, use of grooved pipe couplings, and provision of cleanouts to allow access to all pipe runs.

Monthly average values for TSS, orthophosphorus, and total phosphorus for the UOSA lime clarification process are summarized in Table 5.34. Total phosphorus is reduced from an average of 2.3 mg/L in the secondary effluent to 0.17 mg/L in the second-stage recarbonation effluent.

**Table 5.34** Monthly average performance for the UOSA tertiary clarifiers for 2003.

| Month/ year | Secondary effluent TSS (mg/L) | Secondary effluent OP[a] (mg/L) | Secondary effluent TP[b] (mg/L) | First-stage clarifier effluent TSS (mg/L) | First-stage clarifier effluent TP (mg/L) | First-stage clarifier effluent OP (mg/L) | Second-stage recarbonation effluent TSS (mg/L) | Second-stage recarbonation effluent TP (mg/L) |
|---|---|---|---|---|---|---|---|---|
| 2003 | | | | | | | | |
| Jan | 20.7 | 1.67 | 2.3 | 16.5 | 0.23 | 0.28 | 11.0 | 0.21 |
| Feb | 18.9 | 1.45 | 2.3 | 17.3 | 0.26 | 0.33 | 10.0 | 0.28 |
| Mar | 25.6 | 1.14 | 2.1 | 14.7 | 0.22 | 0.26 | 14.2 | 0.22 |
| Apr | 39.3 | 1.39 | 2.6 | 14.7 | 0.24 | 0.28 | 14.9 | 0.24 |
| May | 36.9 | 1.59 | 2.7 | 12.5 | 0.22 | 0.26 | 18.0 | 0.19 |
| Jun | 9.9 | 1.69 | 2.0 | 12.7 | 0.16 | 0.18 | 12.7 | 0.10 |
| Jul | 14.1 | 1.88 | 2.4 | 9.5 | 0.18 | 0.22 | 12.8 | 0.13 |
| Aug | 10.0 | 1.98 | 2.4 | 11.2 | 0.12 | 0.15 | 7.7 | 0.07 |
| Sep | 13.0 | 1.80 | 2.2 | 13.9 | 0.19 | 0.22 | 10.3 | 0.14 |
| Oct | 9.5 | 2.11 | 2.5 | 12.6 | 0.26 | 0.28 | 9.0 | 0.13 |
| Nov | 8.7 | 1.54 | 1.9 | 13.7 | 0.22 | 0.27 | 8.7 | 0.16 |
| Dec | 27.6 | 1.17 | 2.0 | 20.2 | 0.19 | 0.23 | 19.1 | 0.16 |
| Average | 19.5 | 1.62 | 2.3 | 14.1 | 0.21 | 0.25 | 12.3 | 0.17 |

[a]OP = orthophosphorus.
[b]TP = total phosphorus.

**IOWA HILL WATER RECLAMATION FACILITY, BRECKENRIDGE, COLORADO.** The Breckenridge Sanitation District owns and operates five water reclamation facilities (WRFs) in Summit County, Colorado. The Iowa Hill WRF is the newest of the plants and was placed in service in the spring of 2000 with a maximum monthly design flow of 5680 $m^3/d$ and a peak hourly design flow of 11 400 $m^3/d$. Effluent from the plant discharges to the Blue River, which runs into the Dillion Reservoir. The Blue River is used for recreational activities, and the Dillion Reservoir serves as a drinking water source for the city of Denver, Colorado. An overall water-basin-wide approach has been used at the Breckenridge Sanitation District for regulating phosphorus discharge, with the total phosphorus discharged from all five plants limited to 318 kg/a. Iowa Hill WRF was designed with advanced treatment for phosphorus removal, providing extremely low phosphorus concentrations in the effluent. Construction of the new plant has reduced the need for costly capital upgrades for the district's satellite facilities, while maintaining operating costs at their current levels. By allowing the district to manage a combined phosphorus discharge from the five plants, it can meet current and future discharge requirements while maintaining cost-efficient service. Design limits for the Iowa Hill WRF are shown in Table 5.35.

Limited space at a reclaimed mine, with close proximity to schools and a residential neighborhood, was the only available location for the plant. Therefore, the plant design required a compact, completely enclosed treatment system free of any odor,

TABLE 5.35 Iowa Hill Water Reclamation Facility design effluent limits.

| | Design effluent limits (mg/L) (30-day average/7-day average) | | | |
|---|---|---|---|---|
| **Parameters** | **Secondary clarifier** | **Biological aerated filter** | **High-rate clarifier** | **Sand filter** |
| $BOD_5$ | 25/— | 5/<10 | — | — |
| TSS | 25/— | 5/<10 | <5/<5 | 3/5 |
| Ammonia nitrogen | 30/— | 3/4 | — | — |
| Total phosphorus | 2/— | — | 0.1/0.2 | 0.02/0.02 |

noise, or visual nuisances. The treatment plant occupies 4189 $m^2$, and the administration building occupies an additional 669 $m^2$.

The treatment plant consists of the following sequence of unit processes:

- Rotary fine screens and grit removal;
- Activated sludge biological nutrient removal and secondary clarifiers;
- Interstage flow equalization, intermediate pumping, and fine screens;
- Biological aerated filtration for nitrification;
- High-rate solids-contact clarifier for tertiary phosphorus removal;
- Sand filtration; and
- Chlorination and dechlorination.

The activated sludge process consists of two parallel trains, each having two anaerobic zones (71.4 $m^3$ each) in series followed by an aerobic zone (929 $m^3$ each) and secondary clarifier. This process removes the majority of the BOD and TSS and partially nitrifies the wastewater. Approximately 50% of the total phosphorus is removed, yielding an average secondary effluent total phosphorus concentration of 1.74 mg/L.

Biological aerated filters (four filters—each 26 $m^2$) complete nitrification and maintain the ammonia below 1.0 mg/L. A small amount of phosphorus, approximately 0.1 mg/L, is removed by the biological filters through assimilation and filtration of suspended solids.

Tertiary phosphorus removal is achieved in a high-rate dense sludge clarification process. The pH is maintained in the range of 6.4 to 7.0 by injecting sodium hydroxide (25% solution) at the inlet to the dense sludge process. Aluminum sulfate is injected to the rapid mix tank for coagulation at a dose of approximately 80 mg/L, which varies depending on the influent water quality. The coagulation retention time is approximately 3 minutes at the design average flow of 5680 $m^3/d$. The water is then transferred to the reaction zone, where polymer at a dose of 0.5 to 1.0 mg/L is injected to aid in the flocculation of coagulated particles. Recycled solids are also introduced to the reactor to aid in flocculation. The internal recycling of previously formed solids enhances the solids-contact process and increases reaction rates. At

**Table 5.36** Design criteria for the dense sludge process at Breckenridge, Colorado.

| Item | Units | Value |
|---|---|---|
| Design flow | | |
| Average | $m^3/d$ | 5680 |
| Peak | $m^3/d$ | 11 400 |
| Rapid mix | | |
| No. units | | 1 |
| Type mixing | | Mechanical |
| Width | m | 1.68 |
| Length | m | 1.68 |
| Side water depth | m | 4.34 |
| Detention time | min | 3.1/1.5 |
| Motor power | kW | 1.5 |
| Flocculation | | |
| No. units | | 1 |
| Type mixing | | |
| Width | m | 2.74 |
| Length | m | 2.74 |
| Side water depth | m | 4.34 |
| Detention time | min | 8.3/4.1 |
| Motor power | kW | 2.2 |
| Settling basins | | |
| No. units | | 1 |
| Length | m | 4.72 |
| Width | m | 4.72 |
| Side water depth | m | 4.19 |
| Surface area | $m^2$ | 22.3 |
| Overflow rate | m/h | 10.6/21.3 |
| Scraper speed | rpm | 0.1–0.4 |
| Motor power | kW | 0.37 |
| NaOH[a] dose | mg/L | |
| Clarifier pH | s.u.[b] | 6.4–7.0 |
| Polymer dose | mg/L | 0.5–1.0 |
| Alum dose | mg/L | 80 |

[a]NaOH = sodium hydroxide, also known as caustic soda.

[b]s.u. = standard units.

the design flow of 5680 $m^3/d$, the flocculation reactor retention time is approximately 8 minutes. The densely structured precipitate is transitioned from the reactor basin through a piston flocculation zone to the clarification and thickening zone. A part of the thickened sludge inventory is recycled back to the reactor basin (3 to 6% of the treated flow), thereby increasing the solids in the reactor and improving the performance of the process. Additional solids removal is achieved by the use of settling tubes incorporated to the top of the clarification zone. The settling tube surface area is 11.1 $m^2$ and is designed for a peak overflow rate of approximately 21.3 m/h. Design criteria for the tertiary clarification process are summarized in Table 5.36, and a simplified section for the tertiary clarifier is presented in Figure 5.24.

All sludge and backwash wastewater produced at the Iowa Hill WRF are discharged by gravity through a collection pipe to the Farmers Corner WRF, where the sludge is dewatered with centrifuges. The dried class B biosolids are used for soil remediation at a local mine.

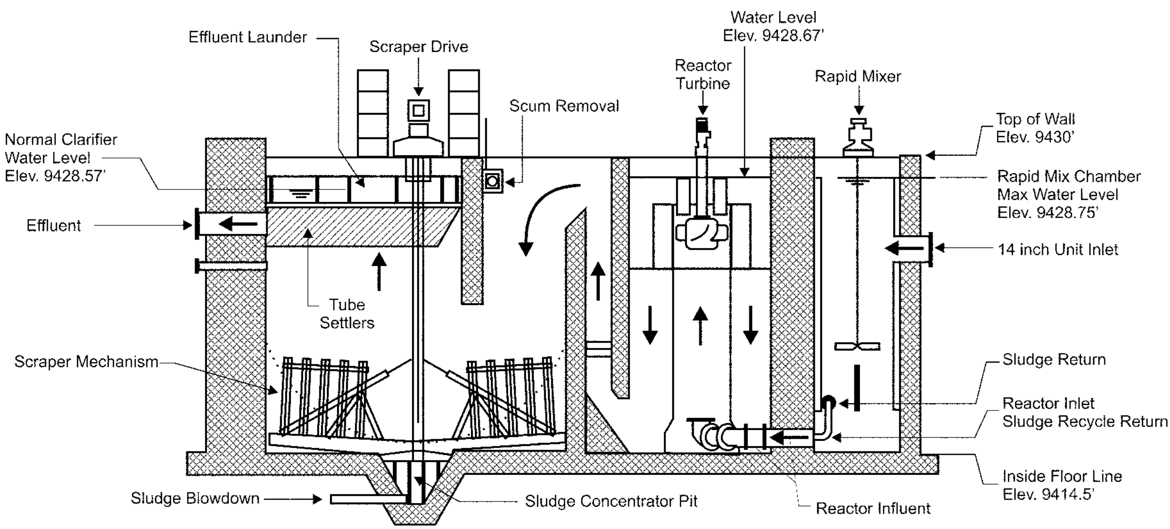


**FIGURE 5.24** Sectional elevation for the dense sludge process at Breckenridge Sanitation District, Breckenridge, Colorado (ft × 0.304 8 = m).

**Table 5.37** Iowa Hill WRF monthly influent and effluent concentrations for 2003.

| Month | Flow (mgd[a]) | BOD influent (mg/L) | BOD effluent (mg/L) | TSS influent (mg/L) | TSS effluent (mg/L) | $NH_3$-N[b] influent (mg/L) | $NH_3$-N effluent (mg/L) |
|---|---|---|---|---|---|---|---|
| January | 0.782 | 137 | 0.76 | 167 | 1.49 | 30.1 | 0.26 |
| February | 0.736 | 159 | 0.94 | 184 | 1.07 | 27.5 | 0.28 |
| March | 0.894 | 165 | 0.75 | 187 | 1.61 | 24.2 | 0.32 |
| April | 0.594 | 184 | 0.89 | 158 | 1.11 | 22.7 | 0.29 |
| May | 0.757 | 123 | 0.95 | 131 | 1.09 | 10.8 | 0.29 |
| June | 0.705 | 169 | 1.03 | 133 | 1.14 | 19.7 | 0.37 |
| July | 0.710 | 178 | 0.75 | 151 | 1.37 | 22.8 | 0.24 |
| August | 0.681 | 151 | 0.72 | 164 | 1.00 | 26.6 | 0.31 |
| September | 0.516 | 142 | 0.73 | 175 | 0.97 | 23.7 | 0.33 |
| October | 0.462 | 135 | 0.56 | 134 | 0.95 | 18.1 | 0.27 |
| November | 0.555 | 155 | 0.78 | 151 | 0.89 | 20.2 | 0.42 |
| December | 0.630 | 166 | 1.3 | 150 | 0.84 | 24.6 | 0.32 |
| Yearly Average | 0.668 | 155 | 0.85 | 157 | 1.13 | 22.6 | 0.31 |

[a]mgd × 3785 = $m^3$/d.
[b]$NH_3$-N = ammonia nitrogen.

A summary of the 2003 monthly plant influent and effluent concentrations for BOD, TSS, and ammonia is shown in Table 5.37. Monthly average phosphorus concentrations in 2003 at the different steps of the plant are provided in Table 5.38. The raw water has an average total phosphorus concentration of 3.6 mg/L, which varies in the range of 1.0 to 8.9 mg/L. High-rate clarification achieves 98% total phosphorus removal and provides an average effluent concentration total phosphorus concentration of 0.032 mg/L.

Final filtration, through four continuous backwash upflow sand filters (37.2 $m^2$ each) removes most the remaining suspended particles and produces a treated effluent with average total phosphorus of 0.009 mg/L.

**Table 5.38** Iowa Hill WRF monthly average total phosphorus data for 2003.

| Month | Plant influent (mg/L) | Secondary effluent (mg/L) | Clarifier effluent (mg/L) | Plant effluent (mg/L) |
|---|---|---|---|---|
| January | 3.80 | 1.32 | 0.033 | 0.014 |
| February | 4.09 | 2.03 | 0.017 | 0.009 |
| March | 3.93 | 1.81 | 0.034 | 0.013 |
| April | 3.37 | 1.19 | 0.116 | 0.012 |
| May | 1.75 | 0.99 | 0.020 | 0.006 |
| June | 2.99 | 1.77 | 0.024 | 0.007 |
| July | 3.74 | 1.22 | 0.012 | 0.005 |
| August | 3.35 | 1.64 | 0.011 | 0.013 |
| September | 5.48 | 3.30 | 0.018 | 0.011 |
| October | 3.63 | 1.60 | 0.012 | 0.005 |
| November | 3.43 | 1.92 | 0.018 | 0.007 |
| December | 3.70 | 2.03 | 0.072 | 0.011 |
| Yearly Average | 3.61 | 1.74 | 0.032 | 0.009 |

## SUMMARY

As with all clarification applications, the design of tertiary clarifiers should be approached from an understanding of the underlying basic principles and mechanisms that affect process costs and performance. Although the use of tertiary clarifiers is not common, substantial numbers are in service. Tertiary clarifiers are primarily used for phosphorus removal and, to a more limited extent, to remove or inactivate microbial contaminants and for membrane pretreatment. While no municipal facilities were specifically identified that use tertiary clarifiers for metals removal, this is a proven process in industrial applications. Coagulation is integral to the use of tertiary clarifiers and the design of rapid and flocculation facilities should be based on jar testing. Even though tertiary clarifiers have been used for many years, relatively few good data are published on the settling velocities of coagulated secondary effluent solids and tertiary chemical precipitates or the operating characteristics and performance of tertiary clarifiers. Hence, the designer is left with the choice of attempting to select an appropriate design velocity from literature that often does not adequately address wastewater applications or conducting settling tests. Settling tests are recommended. More engineering studies are needed that characterize secondary effluent solids and the implications of these characteristics for optimizing the performance of tertiary clarifiers. Potential exists to minimize the cost of tertiary treatment by careful evaluation of clarification alternatives. Such evaluations must include the judicious use of bench and pilot testing coupled with critical performance evaluations of existing systems.

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# Chapter 6

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# INTRODUCTION

**GENERAL.** The purpose of this chapter is to provide environmental engineers, waste treatment designers, and managers with an overview of the state of the art of clarifier modeling. A review of the applicable theory is given along with discussions of the use of models in design, troubleshooting, and plant operations.

Engineers use models to represent the physical, chemical, and biological processes of real systems. There are two types of models that are commonly used in water resources, namely, analogue and mathematical. The term *numerical* in this chapter will refer to models in which real phenomena are represented by mathematical relationships that are solved by digital computers. All models are idealizations of reality and as such have built-in limitations. Thus, the results of all models need to be interpreted and treated with a certain amount of caution.

This chapter includes presentations of several levels of model sophistication, namely physical models, numerical models in one dimension and two dimensions, and the classical state point analysis. The role of three-dimensional (3-D) models is also discussed but not treated in detail.

**TYPES OF MODELS.** Available models in wastewater engineering can be classified as either mathematical or analogue.

***Mathematical Models.*** If these models solve two- or three-dimensional momentum and continuity equations, they are also referred to as computational fluid dynamics (CFD) models. The mathematical relationships in these models may take several forms: empirical regression equations, algebraic equations based on simplified theory, ordinary differential equations, and partial differential equations. Differential equations are often discretized and solved as algebraic equations. Discretization involves converting continuous space and time into a finite number of points (coordinates) in the space–time domain. Information among these discrete points is obtained by interpolation.

The simplest models consist of mathematical functions that are statistically fitted to known inputs and outputs; such models are sometimes referred to as a ***black box*** (empirical) because they give very little insight to the controlling processes. The application of such a model is limited to the calibration conditions used in setting it up. In their most complex form, mathematical models of settling tanks attempt to represent all of important processes in the tank by solving the differential equations of continuity, momentum, energy, mass transport, and biological reactions subject to realistic boundary conditions; these models are called deterministic or ***glass box*** models because they reveal the role of natural laws in determining tank performance. These models can be applied outside of the range of calibration, albeit with caution. There is an intermediate class of models (***opaque or grey box***) that are based on gross simplifications of physical laws, for example, flux theory, plug-flow, and diffusion reactor models.

Mathematical models can also be classified by their spatial resolution as one, two, or three dimensional. There are simple two-cell models and complex multicell, 3-D models. In addition, models may simulate steady-state or unsteady conditions in the tank.

***Physical Models.*** *Physical models* refer to geometrically scaled models that have similar behavior to the full-scale system. In most cases, the physical model is an undistorted scale version of the full-scale system and typically uses the same fluid. Physical models that are used in wastewater treatment are typically reduced-scale versions of the full-scale structure. These models are designed and operated using selected laws of similitude (Kobus, 1980), depending on the dominant phenomenon that is being studied.

Physical scale models fill a similar role to glass box models in that they attempt to represent the physical processes in the real tank; however, true similarity is never achieved because all small-scale physical models are subject to some scale effects.

Typical applications for physical models in wastewater systems include

- Complex 3-D localized flows such as intakes, distribution chambers and manifolds, and energy dissipating inlets;
- Study of density currents in settling tanks;
- Residence time estimation for reservoirs, contact tanks, sedimentation basins, and mixing chambers;
- Demonstration of flow features in process units; and
- Pump wet wells.

A physical model is an analogue of the full-scale system. The full-scale system is referred to as the ***prototype***. Scaling laws are needed to convert the measured data in the model to the equivalent values in the prototype. These laws are based on the similitude or similarity requirements between the model and the prototype. Three types of similarity are required:

- Geometric—similar shape,

- Kinematic—similar velocity and acceleration patterns, and
- Dynamic—similar driving forces.

We define the length ratio, $L_r = L_p/L_m$, where subscript m = model, p = prototype, r = ratio, and $L$ is any corresponding length in meters or feet. Model scale is $S_r = 1/L_r$.

Geometric similarity means the model and prototype have a similar shape. This requires that the dimensionless length ratio,

$$L_r = L_p/L_m = \text{constant} \tag{6.1}$$

for all homologous points of the model and prototype. In addition, we also can write

$$A_r = L_r^2 = \text{area ratio} \tag{6.2}$$

$$\text{Vol}_r = L_r^3 = \text{volume ratio} \tag{6.3}$$

where $A_r$ and $\text{Vol}_r$ are dimensionless variables. Kinematic similarity means that the model and prototype have similar velocity and acceleration patterns. This will be satisfied if the velocity ratio between the prototype and the model is constant at all homologous points, that is,

$$V_r = V_p / V_m = \text{velocity ratio} = \text{constant} \tag{6.4}$$

where $V_r$ is dimensionless. The ratio of settling velocities in the model and prototype must also be constant.

Dynamic similarity means that the model and prototype have similar ratios of driving forces. In practice, complete dynamic or force similitude can never be satisfied. Consequently, engineers try to identify the dominant forces in the process and select the corresponding modeling laws. For example, in an inlet distribution channel, the important forces are gravity and friction, which are governed by the Froude and Reynolds laws, respectively,

$$N_{F\,m} = N_{F\,p} \tag{6.5}$$

and

$$N_{R\,m} = N_{R\,p} \tag{6.6}$$

Where

$N_F = V/(gD)^{1/2}$ (V = velocity [m/s or ft/sec]and $D$ = characteristic depth [m or ft]);
$N_R = VR_h/\upsilon$ ($R_h$ = hydraulic radius [m or ft] and $\upsilon$ = kinematic viscosity [m$^2$/s or sq ft/sec); and
subscripts m and p = model and prototype, respectively.

In practice, the Froude law is typically dominant and is completely satisfied; whereas the Reynolds law is modified as

$$N_{\mathrm{R\,m}} > N_{\mathrm{R\,minimum}} \sim 500 \text{ to } 3000 \tag{6.7}$$

where $N_{\mathrm{R\,m}}$ and $N_{\mathrm{R\,minimum}}$ are dimensionless. The flow in settling tanks is often subject to strong density currents (McCorquodale, 1976, 1987; McCorquodale and Zhou, 1993). To model these effects, we introduce a modification of the Froude law, known as the Densimetric Froude law given by

$$N_{\mathrm{F\,m}}{}' = N_{\mathrm{F\,p}}{}' \tag{6.8}$$

where $N_{\mathrm{F}}{}' = V/(g'D)^{1/2}$ is the nondimensional densimetric Froude number; $g' = g\Delta\rho/\rho_o$ in m/s$^2$ or ft/sec$^2$; the density difference, $\Delta\rho$, with respect to the reference density $\rho_o$ can be caused by differences in suspended solids (kg/m$^3$ or slug/cu ft), water temperature (°C), and/or dissolved solids. In prototype clarifiers, the flow in the inlet and settling zone is typically turbulent; therefore, it is essential that the model should also have turbulent flow. To accomplish this, it may be necessary to exaggerate the model Froude number while maintaining the densimetric Froude number and the minimum Reynolds number. This is typically necessary in modeling clarifiers.

**THE ROLE OF MODELS.** Models are tools that plant designers and operators can use to compliment other field and laboratory methods. Models should not be separated from other information. Even the most sophisticated 3-D models need to be validated with field data.

The purpose of a well-designed model is to answer specific questions. These questions may involve one of the following aspects of clarifiers (Ekama et al., 1997):

(a) Plant design,

(b) Operation,

(c) Training,

(d) Troubleshooting, and

(e) Research.

***Clarifier Design.*** Clarifier design involves several steps and various complexities of models. The preliminary design may be established based on precedence and empirical formulae involving design parameters such as dry weather and wet weather surface overflow rates (SORs), solids loading rate (SLR), detention times, and maintenance and operator preferences. For secondary settling tanks, the state point analysis is often used to estimate surface area of the tank. This will determine the approximate dimensions of the various tanks for the purposes of cost estimation.

During the design stage, models like those based on flux theory or coupled bioreactor–clarifier models as proposed by Ji et al. (1996) can be used to refine the tankage to account for shifts in sludge inventory. One-dimensional (1-D) models address many questions regarding the reactor and the solids inventory management of the coupled system. These coupled models consider interdependence of the clarifier (secondary settling tank [SST]) and the reactor. These models will help identify, at the design stage, any detrimental shift in the inventory of solids from the reactor to the clarifier.

Final detailed design, optimization of tank, and baffle arrangements require at least a two-dimensional (2-D) model and, in some cases, a 3-D model. For example, square settling tanks exhibit strong 3-D flow patterns, whereas circular tanks have been successfully modeled with 2-D approximations. Rectangular tanks are more prone to nonuniformly distributed influent and, therefore, may require 3-D models.

Physical models also have been used in some clarifier projects, for example, the Boston over-and-under clarifiers were studied in a physical model. McCorquodale (1976) used a physical model to explain the hydraulic effects of a thermal density current in the primary clarifiers at the West Windsor Pollution Control Plant (renamed Lou Romano Water Reclamation Plant, Windson, Ontario, Canada). This and subsequent studies have shown that primary clarifiers with temperature fluctuations as low as 0.5 ºC can experience strong density currents (Zhou et al., 1994). The combined use of physical models, numerical models, and full-scale plant studies will give the best results.

***Plant Operation and Control.*** In general, plant operation and control should be based on models such as GPX™ or Bio-Win™ that simulate the whole system. These models should be linked to monitoring data and subjected to continuous updating. Including 2-D settler capability within these models will produce better simulation of effluent quality. These models must be fast and robust.

***Training.*** One-dimensional settling tank models that are linked to appropriate models of all units of the plant are typically adequate for training purposes. These models can be empirical, neural network, or numerical models. Representative results can be obtained because these models can be calibrated or trained with actual plant data. Their response is fast and robust and they can realistically simulate mass inventory.

Graduate students in environmental engineering can benefit from CFD models, especially if they have a user friendly, interactive graphics interface with "animation" of output. These models show many of the liquid–solids interactions such as the effects of density currents, baffles, and sludge-withdrawal systems.

***Troubleshooting.*** Troubleshooting of clarifiers should start with standard field tests as outlined in the American Society of Civil Engineers (ASCE) Clarifier Research Technical Committee (CRTC) protocol (Wahlberg, 1995; Wahlberg et al., 1993) to determine whether a hydraulic problem is the likely cause of poor performance. Flux theory can then be applied to see if the tanks are overloaded. If necessary, this can be followed by 2- or 3-D CFD model simulations to determine hydrodynamic problems and assess the benefit of remedial measures.

***Research.*** Ekama et al. (1997) discusses the use of CFD modeling of secondary settling tanks as a research tool to better understand the effect of the internal geometry of these tanks on their performance. Computational fluid dynamics modeling offers the potential to optimize the tank internals and operating conditions subject to imposed loading and environmental factors such as atmospheric heat exchange and wind shear. Flocculation, sludge rheology, and other associated processes can be investigated in SSTs using calibrated 2- and 3-D models; however, such effects should be validated with field data, whenever possible.

**FIELD AND LABORATORY SUPPORT OF MODELS.** ***General.*** This stage of model development must define the process variables that are to be modeled and system context under which simulations are to be performed. The purpose or objective of the modelling exercise needs to be clearly stated. There is no point in developing a sophisticated model if it does not serve the project objective. For example, a 3-D model may be inappropriate for mass inventory simulation because a 1-D model coupled to the biological reactor is more appropriate and easier to apply.

Although there are similarities between numerical models for grit chambers, primary settling tanks, and SSTs, they are not interchangeable. The process or processes to be modeled must be understood and the final use of the model set out at the beginning. Models may fulfill several purposes as outlined in the Commercial Computational Fluid Dynamics Models section.

Depending on the end use of the settling tank model, it may be necessary to include the processes in other unit operations. For example, for solids inventory it will be necessary to include the biological reactor to be coupled with the SST. An operational training model should permit the inclusion of all relevant unit processes in the wastewater treatment plant; such a model must represent the dynamic nature of the physical and biological processes. The design of an SST for effective solids inventory management requires a dynamic model that couples the settling tank and biological reactor. A design model for optimizing settling tank size and internal geometry must have high spatial resolution and must accurately simulate physical processes in the tank.

Other items in the problem definition that should be specified are

(a) Type of tank, for example, circular, rectangular, square, and upflow;
(b) Type of feed, for example, center or peripheral;
(c) Type of effluent launders, for example, peripheral weir, inboard launder with weirs, and submerged manifold;
(d) Type of sludge withdrawal (suction or scraper with hopper);
(e) Type of flocculation zone;
(f) Hydraulic loading;
(g) Solids loading;
(h) Settling properties;
(i) Dominant spatial nature of flow, for example, 1-, 2-, or 3-D;
(j) Time dependence of tank processes, that is, is the flow steady, unsteady, or quasi-steady (slowly changing with time); and
(k) Significance of biological activity.

Figure 6.1 shows the general flow pattern suggested by Larsen (1977) for rectangular settling tanks. Larsen considered the importance of several mechanisms in controlling the flow pattern in a rectangular settling tank. These were

(a) Inlet kinetic energy (KE),
(b) Potential energy (PE) of influent suspended solids caused by density difference compared to the ambient fluid,
(c) Energy dissipation caused by friction,
(d) Gravitational work done on the fluid,
(e) Wind shear energy transfer, and
(f) Atmospheric heat exchange.

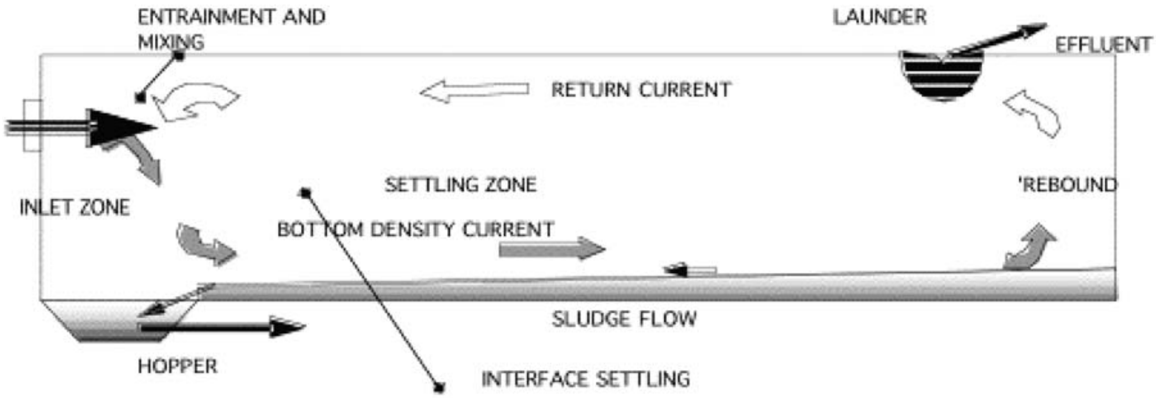


**FIGURE 6.1** Flow processes in a clarifier (adapted from Larsen, 1977).

Larsen also discussed time-dependent effects such as diurnal changes in influent temperature and total dissolved solids (TDS). His findings are summarized in Table 6.1 where the approximate relative energy rates are expressed in terms of percent of the inlet kinetic energy flux. The greatest energy fluxes are the inlet kinetic energy, the wind shear energy exchange, and the heat transfer; however, the inlet kinetic energy is more than 90% dissipated in the inlet zone. Wind shear is a stochastic effect. High winds can transfer a large amount of energy to the tank and possibly have an adverse effect on the hydraulics of the settling zone because of high internal mixing, setup, surface waves, and seiching in the tank. One effect of wind is to produce a 3-D flow pattern with nonuniform distribution of the hydraulic and solids flow over the effluent weirs. The friction energy loss related to "plug flow'" because of the mean flow is very small; however, in a real tank, the flow distribution is not uniform and friction forces are higher than in the ideal case. According to Larsen, the effect of the slope of the water surface is negligible, except near the weir. The PE as a result of the higher density of the influent is relatively small compared to the inlet KE but this PE is what causes the bottom density current. Settling tanks typically dissipate most of the inlet KE in the inlet zone; therefore, the PE flux caused by influent suspended solids has a significant influence on the flow pattern in the settling zone.

Figure 6.2 summarizes some of the important flow processes in a circular secondary settling tank (for example, Adams and Rodi, 1990; Anderson, 1945; Bretscher et al., 1992; Celik et al., 1985; Krebs, 1991; Krebs et al., 1995; Krebs et al., 1999; Lakehal et al. 1999; McCorquodale and Zhou, 1993; Robinson, 1974; Stamou and Rodi, 1984;

**TABLE 6.1** Energy balance in a secondary settling tank (Ekama et al., 1997; Larsen, 1977).

| | Energy rate compared to inlet KE (%) | Comments |
|---|---|---|
| Inlet KE flow | 100 | $\rho QV^2$ |
| Potential energy of inlet suspended solids | 20 | As a result of difference in density between MLSS and settling zone |
| Wind shear energy transfer | 0–>100 | Depends on tank exposure and shape |
| Water surface slope | < 5 | Expect near launder |
| Kinetic energy flux in settling zone | <5 | |
| Mechanical inputs (scrapers) | 5–10 | |
| Atmospheric heat transfer | >10 | Varies seasonally |
| Dissipation of KE in inlet zone | 90 | Could be enhanced by an inlet energy dissipation tub |
| Loss of inlet suspended solids PE | 10 | |
| Net friction loss rate | ~ 5–10 | |
| Loss of suspended solids PE in effluent | ~ 1 | |
| Loss of suspended solids PE in settling zone | 6 | |
| Net KE in outfall | <1 | |
| Other losses, e.g., internal and surface wave breaking, eddies | ~ 100 | |

Stamou et al., 1989; Szalai et al., 1994; Van Marle and Kranenburg, 1994; WPCF, 1985; Zhou and McCorquodale, 1992a, 1992b, 1992c). The important flow processes are (a) jet flow at inlet; (b) dissipation of KE in the inlet zone; (c) density water fall; (d) entrainment of clarified liquid into the density waterfall, increasing the total flow in the density waterfall; (e) formation of the bottom density current; (f) withdrawal of

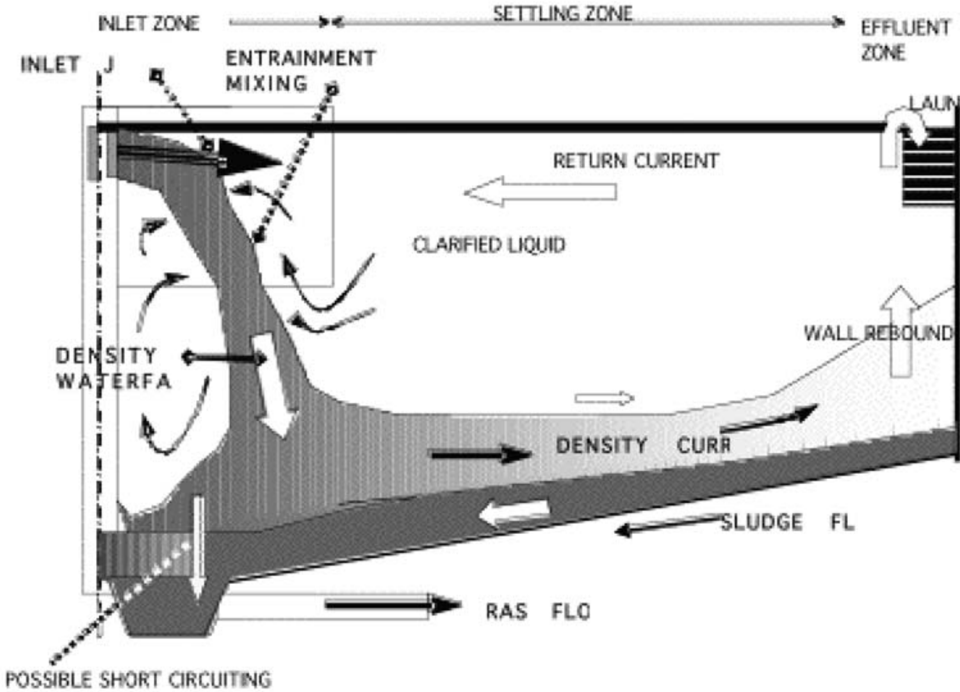


**FIGURE 6.2** Flow processes in a circular clarifier.

clarified liquid to the launder; (g) recirculation of excess flow (in the upper clarification zone); (h) sludge return flow (mostly caused by gravity assisted by scrappers); (i) return activated sludge (RAS) withdrawal via the sludge hopper; and (j) possible short-circuiting from the inlet zone to the RAS withdrawal. Rapid sludge withdrawal using an organ pipe or Tow-Bro system essentially distributes the RAS withdrawal over a greater area of the floor; however, flow processes are similar.

***Solids Loading.*** The solids loading rate is expressed as the rate of mass loading (mass/time) over the surface area of the settling tank ($L^2$). A common unit for SLR is kilograms per square meters per hour. This is useful for comparing solids loading for different plants. Typically, solids loading input to a model is expressed as a time series of suspended-solids concentration(s) in milligrams per liter and the incoming flow rate. In the case of SSTs, concentrations are mixed liquor suspended solids (MLSS) concentrations from the biological reactor, with typical values in the range 1000 to 3500 mg/L. The MLSS values have a diurnal variation.

***Hydraulic Loading Rate.*** The hydraulic loading rate refers to the volumetric effluent flow from the tank. This may be expressed as a total flow or as a flow per unit of surface area. The SOR is the effluent flow ($L^3/T$) divided by the surface area of the settling tank ($L^2$). Common units of SOR are meters per hour (or gallons per square foot per day). The total flow into a settling tank includes all recycled flows. In the case of an SST, this rate is referred to as RAS flow and can be quite high, for example, 25 to 100% of the effluent flow. The detention time of a settling tank is the tank volume ($L^3$) divided by the total flow through the tank ($L^3/T$). Hydrodynamic models require a time series of effluent flowrates and RAS flows. Under normal conditions, the flow series is diurnal and, for design purposes, can be represented by an average dry weather flow (ADWF) and a diurnal amplitude. Wet weather conditions are represented in a model by a hydrograph for the design event or as a peak wet weather flow (PWWF). It is noted that PWWF = $f_p$ x ADWF, where $f_p$ is a dimensionless peaking factor.

***Settling Characteristics.*** Figure 6.3 shows a general classification of sedimentation processes and factors affecting sedimentation in wastewater treatment plants (Imam, 1981). Ekama et al. (1997) have reviewed the various types of settling that can occur in an SST, that is, discrete settling, hindered settling, zone settling, and compression. In grit and primary settling tanks, the settling process is dominated by discrete settling, in which the settling velocity is given by the Stokes settling velocity,

$$V_o = \{4gD_p\,(S_s - 1)/(3\,C_d)\}^{1/2} \tag{6.9}$$

Where

$C_d$ = $24/N_R$ for low particle Reynolds numbers ($N_R = D_pV_o\,/\,\nu < 0.1$);
$D_p$ = particle diameter (m or ft); and
$S_s$ = particle specific gravity (nondimensional).

Vesilind (1968) developed a settling velocity equation for zone settling,

$$V_s = V_o e^{-kX} \tag{6.10}$$

Where

$k$ = constant ($m^3$/kg or L/mg),
$X$ = the local concentration of the suspended solids ($kg/m^3$ or mg/L),
$V_s$ = zone settling velocity (m/h), and
$V_o$ = Stokes settling velocity (m/h).

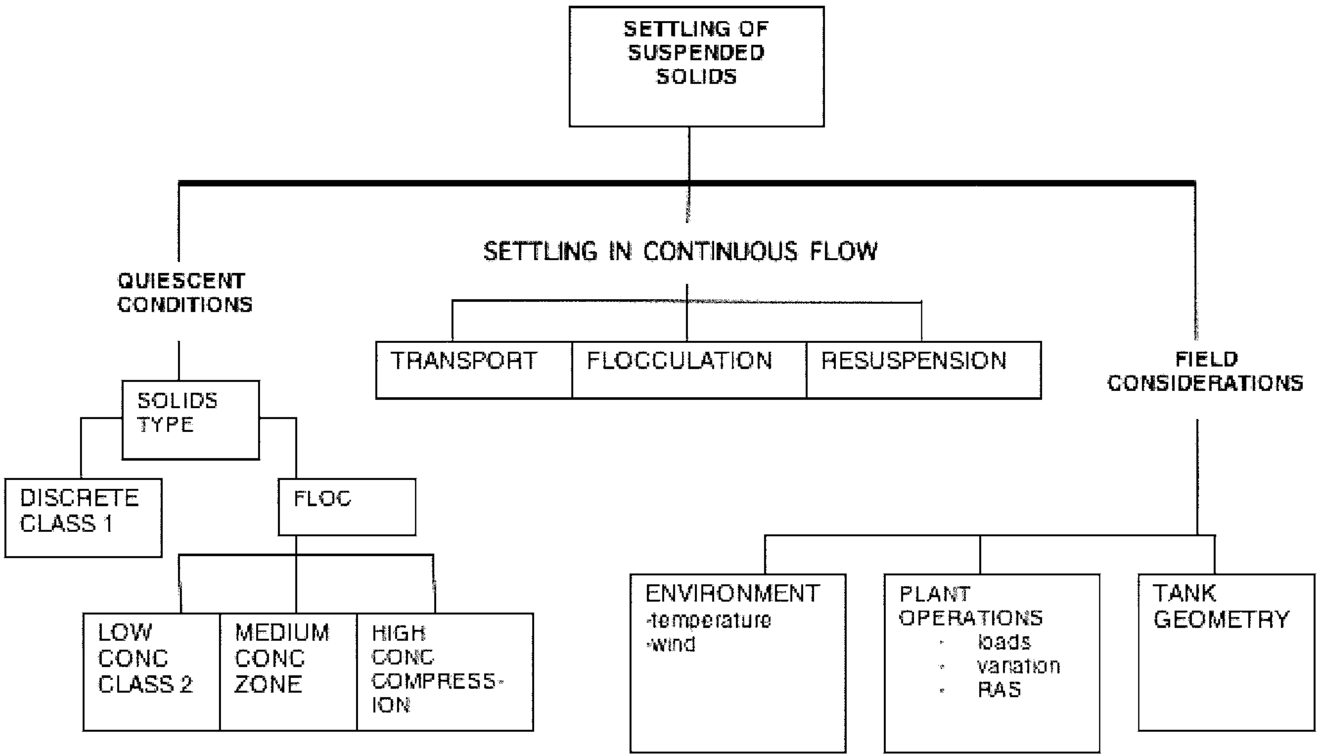


**FIGURE 6.3** Schematic of factors affecting settling of suspended solids (after Imam, 1981).

Ekama et al. (1997) presented several equations to relate $k$ and $V_o$ values to commonly measured sludge characteristics such as sludge volume index (SVI) or stirred SVI (sSVI) or diluted SVI. Takacs et al. (1991) proposed an improvement to the sludge settling equation, originally proposed by Vesilind (1968) to account for the behavior of discrete solids as well as zone settling. They proposed to model settling velocity (m/h) for suspended solids as follows:

$$V_s = V_o\,[e^{-K1(X\text{-}X\text{min})} - e^{-K2(X\text{-}X\text{min})}] \tag{6.11}$$

Where

$X_{min} = f_{min}X_o$ ($f_{min}$ = fraction of unsettleable solids in MLSS and $X_o$ = concentration of MLSS [mg/L]);
$K_1$ = hindered settling parameter for floc (L/mg);
$K_2$ = settling parameter for slowing settling solids (e.g., colloidal particles [L/mg]); and
$V_o$ = Stokes settling velocity for discrete flocs (m/h).

In addition, Takacs et al. (1991) gave the following guidelines:

(1) unsettleable solids: $X = X_{min}$ (a few milligrams per liter, e.g., $< 5$ mg/L);
(2) Slowly settleable solids, consisting of floc that have been separated from the large floc but can be reflocculated: $X_{min} < X < 100$ mg/L; and
(3) Highly settleable solids, consisting of large floc: $X > 100$ mg/L.

Wahlberg and Keinath (1988) provided a means to estimate the floc settling characteristics, that is,

$$V_o = 15.3 - 0.061\,5(\text{sSVI})\ (\text{m/h}) \tag{6.12a}$$

$$K_1 = -0.426 + 0.003\,84(\text{sSVI}) - 0.000\,054\,3(\text{sSVI})^2 \tag{6.12b}$$

where sSVI = stirred SVI (mL/g).

The coefficient $K_2$ depends on site-specific conditions. $K_2$ is a parameter that attempts to account for unflocculated particles in the clarified zone of the settling tank. Figure 6.4 shows the traditional single exponential equation and the double exponential formula. Curves representing good settling sludge and poor settling sludge are presented.

It is preferable to model all three classes of suspended solids: primary biotic particles, flocculated particles, and particles that cannot be flocculated by physical

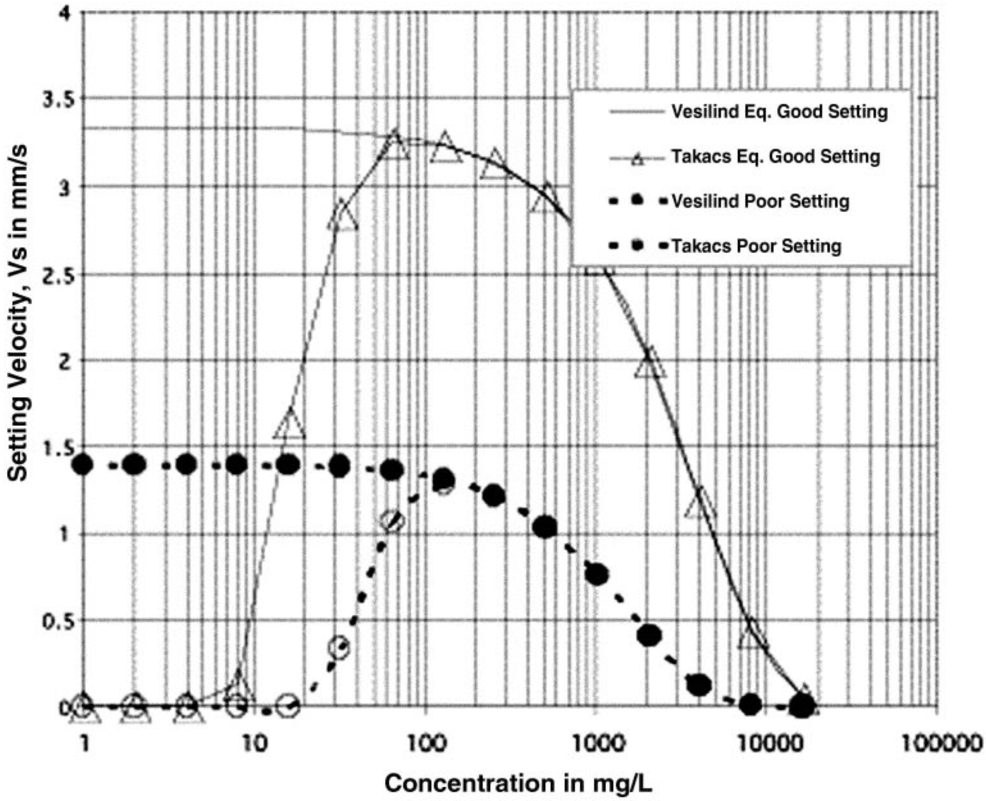


**FIGURE 6.4** Settling characteristics of activated sludge.

processes within the settling tank. The orthokinetic flocculation rate is considered to be first-order process, depending on the velocity gradient ($G$, $s^{-1}$), concentration of particles, and the floc volume fraction (Parker et al., 1970). Commonly available 2-D models compute the velocity field from which the local $G$ can be determined; alternately, the local value of $G$ can be estimated from $G = (\rho\varepsilon/\mu)^{1/2}$ where $\varepsilon$ = turbulent energy dissipation rate ($m^2/s^3$)and $\mu$ = the viscosity (N $s/m^2$)(Parker et al., 1970, 1972). Such models require that the solids be treated in multiple classes as was done for primary clarifiers by Abdel-Gawad (1983) and Abdel-Gawad and McCorquodale (1984) rather than in a single class as is currently the practice in SST models. The diagnostic tests for dispersed suspended solids (DSS, in milligrams per liter) and flocculated suspended solids (FSS, in milligrams per liter) suggested by Wahlberg et al. (1995) can be used for setting initial conditions on the nonsettleable and unflocculated classes. The DSS test indicates the unflocculated particles at the location in the system where the sample was taken. For example, comparing the DSS in the MLSS of the influent and the DSS of the liquid near the outer edge of the flocculation well indicates the flocculation that occurred in the center well. Similarly, the DSS at the effluent relative to the DSS at the center well shows the extent of flocculation in the settling zone of the clarifier. The FSS test is an estimate of the concentration of particles that can not be flocculated by normal physical processes. The difference between the DSS and FSS at the effluent is an indicator of a flocculation problem in the tank. The application of DSS and FSS as diagnostic tests is discussed in Chapter 2 and in Parker et al. (2000). Because FSS depends on $G$, it may be instructive to do the test for several $G$ values to determine the optimum $G$ for the minimum FSS.

Takacs et al. (1991) determined values for the parameters in eq. 6.11 from Pflanz's (1969) full-scale datasets. These data have been reviewed and extended as shown in Table 6.2.

McCorquodale et al. (2004) examined the settling properties of activated sludge in four categories: (1) nonsettleable particles, (2) discrete settling particles, (3) hindered or zone settling flocs, and (4) compression phase. Nonsettleable particles were assumed to be represented by FSS. Discrete settling was determined to correspond to particle settling at a diluted concentration of MLSS ($X_d$, the limit in milligrams per liter below which discrete settling can be assumed to occur), below which a "zone settling" interface could not be observed. A modified settling column test was performed to determine the fractions of the solids in four settling classes (e.g., $V_s = 0$, 1, 3, and 10.5 m/h). Figure 6.5 shows the solids distribution by velocity classes for a sample of MLSS from the Marrero wastewater treatment plant (WWTP) in Jefferson Parish, Louisiana; this sample was diluted from 3000 mg/L to 600 mg/L for this test.

**TABLE 6.2** Typical settling parameters (after Ekama et al., 1997; Takacs et al., 1991).

| Approximate SOR (m/h) | RAS (%) | MLSS (mg/L) | $V_o$ (m/h) | $K_1$ L/mg | $K_2$ L/mg | $f_{min}$ |
|---|---|---|---|---|---|---|
| 0.5–2 | 25–60 | 1500–2500<br>SVI ~ 80–100 | 5–15 | 0.000 2–0.001 | 0.005–0.1 | 0.001–0.003 |

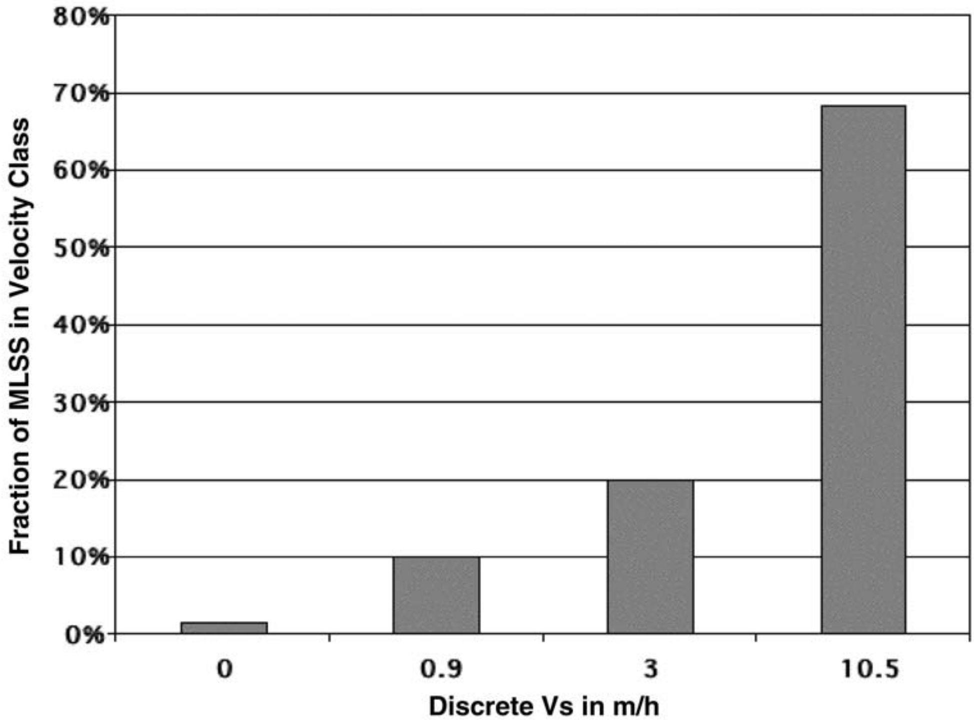


**FIGURE 6.5** Example of distribution of particle concentration in diluted MLSS from the Marrero WWTP.

Class "0" was estimated by the FSS test; classes "1" and "2" are assumed to be suitable for flocculation; and class "3" represents highly flocculated particles. The Parker et al. (1970) flocculation model is being used to simulate the transfer of mass among the classes "1", "2", and "3" within a 2-D clarifier model called *2-DC* (McCorquodale et al., 2004). Hindered settling (standard zone settling) tests were used to determine the $V_o$ and $K_1$ in the Vesilind equation (eq 6.10). In addition, to improve on the description of the compression zone, "zone" settling type tests were performed using RAS samples to estimate another Vesilind-like equation with its own $V_c$ and $K_c$ where $V_c$ is the compression rate (m/h) and $K_c$ is the compression parameter (L/mg). As indicated by Figure 6.6, the "zone settling" Vesilind curve intersects the "compression" Vesilind curve at $X_c$. It was noted that the compression curve tends to give higher compression rates than the extrapolation of the "zone settling" part of the

Vesilind equation. The *2-DC* model has three settling conditions: discrete class settling for $X < X_d$, zone settling for $X_c >$concentrations $> X_d$, and compression for $X > X_c$. Flocculation may occur under any of these conditions. An alternative to the Vesilind equation is a flux equation in the form given by Cho et al. (1993):

$$V_s = K_o\,[e^{-kX}]/X \tag{6.13}$$

where $K_o$ is a parameter determined from the zone settling test, mL/mg·h. This equation was found to fit the hindered and compression data in Figure 6.6 better than a single Vesilind equation; however, this equation failed to represent the discrete settling zone, for example, $X < 0.6$ g/L.

***Floc and Sludge Density.*** Density currents in secondary clarifiers occur because there is a difference in density between the influent MLSS and that of the contents of the tank. In SSTs, density currents are primarily caused by the fact that the MLSS in the influent make the inflow denser than the liquid in the settling zone. In municipal SSTs, temperature and TDS are less important than in primary settling tanks (PSTs)

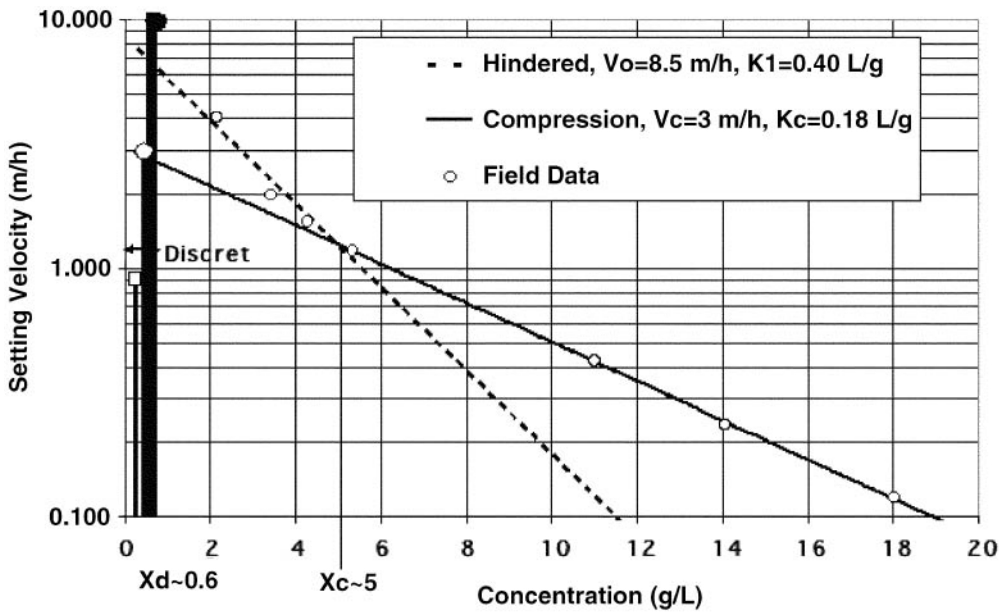


**FIGURE 6.6** Settling characteristics of activated sludge from the Marrero WWTP.

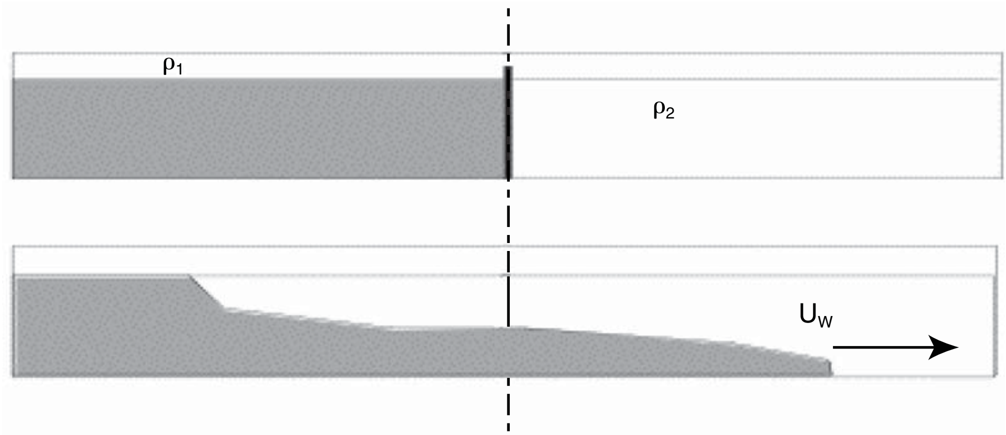


**Figure 6.7** Lock exchange method of estimating density of sludge and flocs (after Larsen, 1977).

or industrial SSTs. The density of the MLSS depends on its concentration and the density of the solids in the liquid–solids mixture. In PSTs, thermal effects are equally or more important than suspended-solids density effects.

Larsen (1977) quantified the density of the liquid–solids mixture by performing a "lock exchange" test. The test is made in a flume as illustrated in Figure 6.7. The liquid–solids mixture is placed on one side of the gate and clear water is placed on the other side. When the gate is lifted, heavier liquid flows under the lighter liquid and visa versa with velocities in meters per second that are related to the density difference by

$$U_w = 0.5(gh\Delta\rho/\rho)^{0.5} \tag{6.14}$$

Where

$\Delta\rho = (\rho_1 - \rho_2)$ = effective difference in the density of liquid–solids mixture and the clarified fluid ($\rho_2$ = reference density of ambient clarified liquid [kg/m$^3$]);
$\rho_1$ = density of liquid–solids mixture (kg/m$^3$);
$g$ = gravitational acceleration (m/s$^2$);
$h$ = tank depth (m); and
$U_w$ = advancing speed of interchange wave (m/s).

Larsen (1977) also used the following relationship for the mixture density:

$$\rho = \rho_r + X\,(1 - S_s^{-1}) \tag{6.15}$$

where $X$ = concentrations of suspended solids (the same units as the density, kg/m$^3$)and $S_s$ = specific gravity of the dry solids (no units). Larsen (1977) found that activated sludge had an $S_s$ in the range of 1.2 to 1.4 (dimensionless). Kinnear (2002) developed a centrifugal method of estimating $S_s$ and found higher values than Larsen. The specific gravity of the flocs, $S_{sf}$ (dimensionless), is much closer to 1 because of the larger amount of water contained in the floc structure (Li and Ganczarczyk, 1987). The discrete settling velocity depends on the effective specific gravity of the flocs ($S_{sf}$, which is nondimensional), whereas the density currents depend on $\Delta\rho$, which is a function of $X$ and $S_s$; the reference density of ambient clarified fluid depends on the fluid temperature ($T$) and the dissolved solids ($S_{TDS}$) and is based in part on an equation from Thomann and Mueller (1987):

$$\rho_r = a_o + a_1 T + a_2 T^2 + a_3 T^3 + a_4 T^4 + (0.802 - 0.002T)(S_{TDS} - 0.035) \tag{6.16}$$

Where

$\rho_r$ = liquid density excluding the suspended solids (kg/m$^3$);
$T$ = liquid temperature (°C);
$S_{TDS}$ = TDS (mg/L);
$a_o$ = 999.871 362;
$a_1$ = 0.066 418 255;
$a_2$ = − 0.008 872 389;
$a_3$ = 8.470 85 x 10$^{-05}$; and
$a_4$ = 8.450 21 x 10$^{-07}$.

***Compression Characteristics.*** Current CFD models treated the compression phase as an extension of the Vesilind equation. However, in the sludge blanket, the displacement of liquid by solids should be considered. This requires a two-phase approach. Wallis (1969) presents the basic principles of 1-D, two-phase flows of solids and liquids. He considers four possible stages in batch sedimentation: (1) clear liquid zone, (2) a nearly constant concentration or zone settling, (3) a variable concentration zone, and (4) a maximum density zone. He shows that batch consolidation in the variable concentration zone is a diffusion process similar to Terzaghi's soil consolidation equation (1925). Greimann and Holly (2001) applied two-phase principles to modeling steady-state solids concentration profiles in uniform open channel flows. Kinnear (2002) has developed a model for column settling of sludge based on two-phase flow theory. Kinnear has shown that the parameters needed to set up such a model can be derived from a modification of the settling column test and a floc density test. The compression rate of a thickened activated sludge is compared to the zone settling Vesilind equation in Figure 6.6; this suggests that the Vesilind curve should not be extrapolated into the thickened zone as shown by the results presented in Figure 6.6.

***Sludge Rheology.*** Wastewater sludges behave as non-Newtonian fluids at high solids concentrations such as those often found at the bottom of the sludge blanket (Dick and Ewing, 1967, Geinopolos and Katz, 1964, Sonanski et al., 1997). This has important implications in collection and removal of thickened sludge. Geinopolos and Katz (1964) showed that primary, secondary, and digested sludges behave as visco-plastic materials. Three common models are proposed to describe this behavior:

$$\tau = \mu D_v \tag{6.17}$$

$$\tau = \tau_o + K D_v \tag{6.18}$$

and

$$\tau = k D_v^{\,n} \tag{6.19}$$

The combination of eqs 6.18 and 6.19 gives

$$\tau = \tau_o + k D_v^{\,n} \tag{6.20}$$

Where

$\tau$ = the shear stress ($N/m^2$);
$\tau_o$ = the yield stress (function of solids concentration) ($N/m^2$);
$\mu$ = the dynamic viscosity (function of solids concentration) ($N\ s/m^2$);
$D_v$ = the shear rate, for example, $D_v \sim du/dy$ near the bed ($s^{-1}$);
$K$ = the consistency coefficient (function of solids concentration) ($N\ s/m^2$); and
$k$ and $n$ = empirical constants or functions of solids concentration.

DeClercq (2003) found that the mixed formulation of eq 6.20 was the best approximation for activated sludge. Lotito et al. (1997) presented the concentration function for the application of eqs 6.17 and 6.19 to four types of sludge: waste activated, raw, anerobically digested, and mechanically dewatered. Their activated sludge curves are shown in Figure 6.8. Similar work has been presented by Lakehal et al. (1999) and applied by Armbruster et al. (2001) to secondary clarifier modeling. Other approximations can be found in the research of Bokil and Bewtra (1972) and Casey (1992). Bird (1976) provides an introduction to non-Newtonian models.

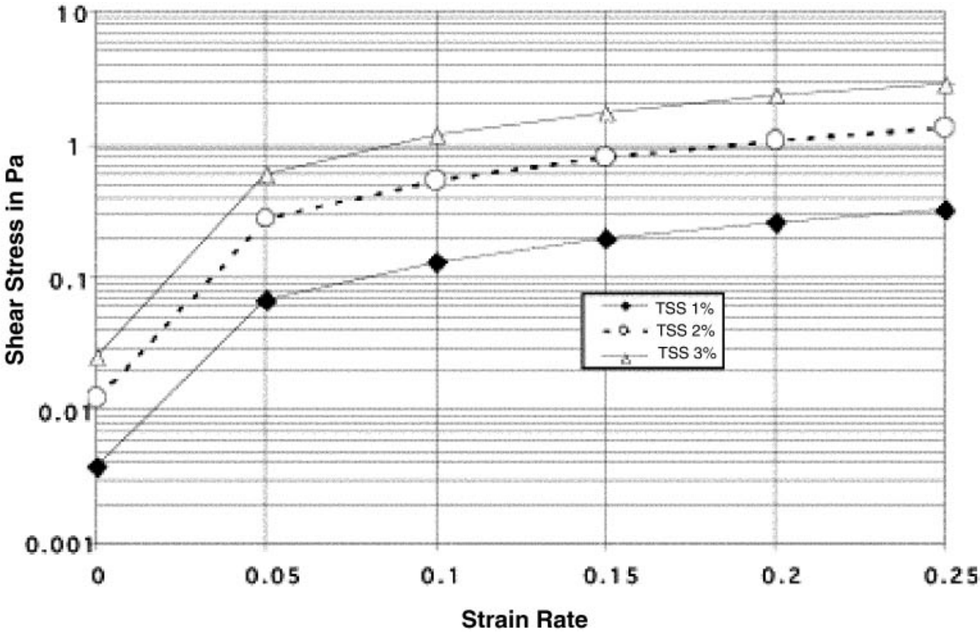


**FIGURE 6.8** Typical shear–strain rate curves for activated sludge (reprinted from *Water Science & Technology,* **36** (11), 79–85, with permission from the copyright holder, IWA).

***Flocculation Models.*** Flocculation is the accumulation of smaller particles into larger agglomerations. Flocculation is essential for effective settlement of biological and colloidal particles. The factors that affect floc formation and breakup include velocity gradient ($G$), concentration of suspended solids by class, differential settling velocities, and residence time.

Jiménez et al. (2003); La Motta et al. (accepted for publication); La Motta, Jiménez, Josse, and Manrique (2003); La Motta, Jiménez, Parker, and McManis (2003); Larsen (1977); Parker et al. (1970); and Wahlberg et al. (1994) reviewed and presented useful information on the flocculation process. Camp and Stein (1943) and Parker et al. (1972) used the following formula for the local velocity gradient:

$$G = (\rho\varepsilon/\mu)^{1/2} \tag{6.21}$$

where $\varepsilon$ = turbulent energy dissipation rate ($m^2/m^3$)and $\mu$ = the dynamic viscosity (N s/$m^2$). Because floc formation and breakup depend on $G$, the predicted distribution of $\varepsilon$ can be used to determine the local rate of floc formation or breakdown; by treating the discrete dispersed particles (class II floc) as a separate dependent variable, it is possible to more realistically model these slowly settling floc and the mass exchange with the large, well-formed floc. This approach will lead to a more general model that is less dependent on the tank for which it has been calibrated. Thus, it may be possible to avoid $K_2$ in eq. 6.11 as the major calibration variable in 2- and 3-D modeling. Wahlberg et al. (1994) showed that batch flocculation requires approximately 10 minutes and continuous flow requires approximately 20 minutes. The research of La Motta et al. (accepted for publication) support the findings of Wahlberg et al. (1994) that indicate that flocculation depends on the production of extracellular polymers, which bind biological and particulate carbonaceous oxygen demand (COD) into larger flocs that can be removed by sedimentation. They showed that the removal was a first-order process and could account for a significant proportion of the COD removal.

***Calibration Tests.*** Crosby (1984) provided insight to the hydraulics of settling tanks. He developed a "synoptic" dye test procedure that reveals aspects of the internal flow in a tank that could not be deduced from dye flowthrough curves (FTC) tests. The Crosby test gives us a series of "snapshots" of a dye front as it progresses through a tank. Because the Crosby test requires suspended-solids distribution to be collected at the same time, these data are of great value in calibrating and verifying numerical models. Figure 6.9 shows the results of a synoptic dye test of the Renton

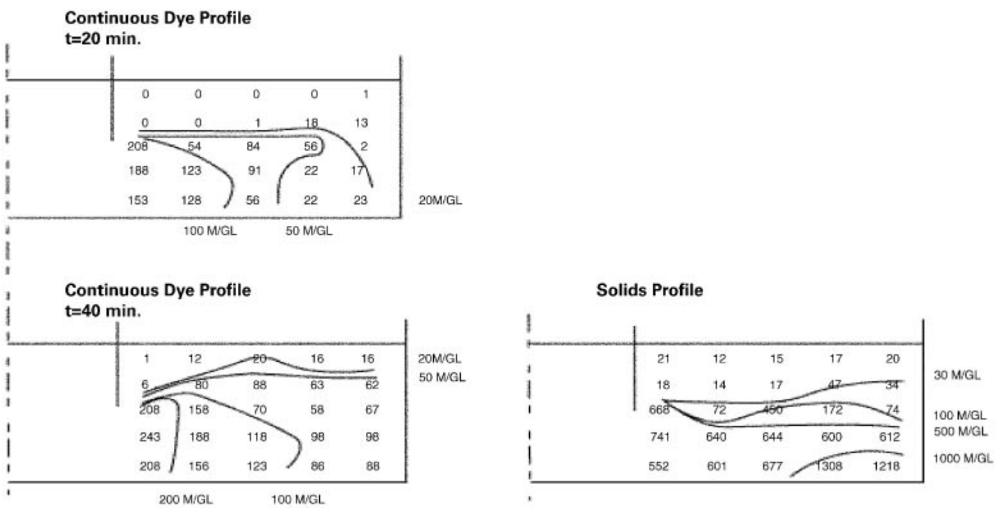


**FIGURE 6.9** Typical Crosby dye test with solids distribution at Renton WWTP (Samstag et al., 1992).

(Seattle, Washington) secondary clarifiers (Samstag et al., 1988, 1992). The advancing dye iso-concentration lines illustrate that there is a strong "density" current in the lower portion of a center-fed circular SST. The CRTC protocol (Wahlberg et al., 1993) has adopted the Crosby test as a standard. Recently, Kinnear (2002) applied an acoustic Doppler current profiler to measure the velocities in a secondary clarifier. His approach also gave the velocity profile and Reynolds stresses.

Examples of model calibration and verification are presented by Ekama and Marais (2002), Kleine and Reddy (2002), and STOWA (2002). Ekama and Marais (2002) compare the performance of 2-D models with flux models and full-scale tanks data. They showed that the 2-D model of Zhou and McCorquodale (1992a, 1992b) correctly predicted "failure/no failure" in 12 out of 15 full-scale stress tests; there was an incorrect prediction of "no failure"' in two of the cases that failed in the field. Vitasovic et al. (1997) presented an example of the calibration of a 2-D model using data from the Denver WWTP, Colorado.

***Example of a Two-Dimensional Model Calibration.*** The 2-DC model (McCorquodale et al., 2004) was calibrated using seven days of operational data from the Marrero WWTP, Jefferson Parish, Louisiana. The settling characteristics for suspended solids are similar to those given in Figures 6.5 and 6.6. Figure 6.10 presents a comparison of the predicted and measured effluent suspended solids

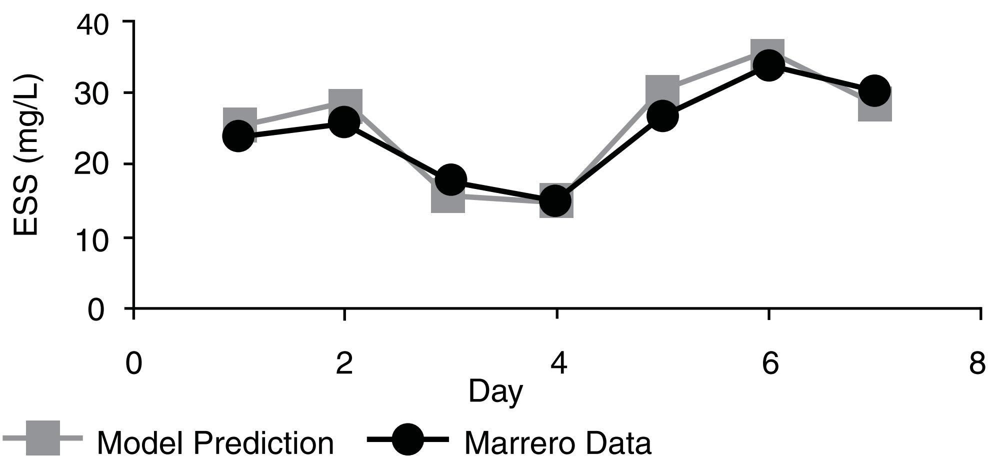


**FIGURE 6.10** Results of effluent suspended solids (ESS) calibration of a 2-D SST model for Marrero WWTP, Louisiana, SOR (0.7 to 1.6 m/h); average MLSS = 2800 mg/L.

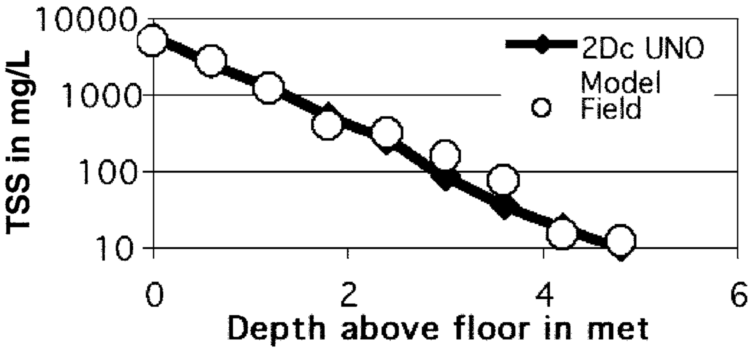


**FIGURE 6.11** Comparison of measured and 2-DC predicted solids distribution at midradius of Marrero WWTP, Louisiana.

(ESS). Figure 6.11 shows a comparison of the measured and modeled solids distribution averaged over the seven-day test. This example illustrates that a 2-D model can simulate the clarifier performance if appropriate settling properties are measured at the time of field data collection. The calibration is made by adjusting the "flocculation" parameter, $K_2$. A constant $K_2$ was assumed. However, because of variations in the settling properties, it is not possible to apply this calibrated model to predict future performance without some specific knowledge of the future settling characteristics, for example, sSVI, DSS, and FSS data. The sSVI during this calibration varied from 50 to 120 mL/g and SOR varied from 0.7 to 1.4 m/h during field calibration tests.

## GOVERNING EQUATIONS

**GENERAL EQUATIONS.** The application of the laws of conservation of mass, momentum, and energy successfully combines CFD with solids flux theory to provide a representative picture of the hydraulic and solids regimes within the clarifier. Typically, clarifier models contain a set of conservation equations for the liquid phase, solid phase, and momentum. The settling velocity function may be linked to the suspended solids and sSVI (e.g., Daigger and Roper, 1985; Takacs et al., 1991). The eddy viscosity is often determined by the k-ε turbulence model (Rodi, 1980). The model must accurately simulate the mass drawoff from the RAS flow and the sludge inventory in the tank.

The equations describing general turbulent flows used in clarifier models have been known for more than 100 years. These consist of 3-D unsteady Navier–Stokes equations (momentum equations), continuity equations, and mass-transfer equations. Any simplification in the basic equations could affect a model's ability to accurately simulate the significant physical process of sedimentation and flow dynamics under certain circumstances. A momentum or mass-transfer equation can be interpreted as follows:

1. Transient term that describes the variations of mass transfer with respect to time. This is essential in any unsteady clarifier model.
2. Three convection terms corresponding to three directions, which describe the mass-transfer process caused by flow movement in 3-D space.
3. Three turbulence diffusion terms in three directions, which describe the mass mixing process caused by turbulent diffusion. A turbulence model is needed to determine eddy viscosity in the diffusion terms.
4. A sink (or source) term, which often includes the flow driving force such as pressure gradients and density gravity terms, the solids settling flux, and the concentration decay rate.

Using the single-phase flow assumption, the equations described above can be considered as the theoretical model to present the major physical processes of solids movement. In cases with chemical or biological decay processes, the empirical or semiempirical relationship must be introduced to determine the decay rate used in the sink term in the mass-transfer equations for different water-quality components. The single-phase assumption implies that the volume occupied by the solids is negligible.

Useful references for the differential equations that are used in CFD modeling are Bird et al. (1960), Hossain and Rodi (1982), and Rodi (1980). The following conservation equations can be used to describe 2-D, unsteady, turbulent, and density stratified flow in a settling tank using either rectangular or cylindrical coordinates.

### *Continuity (Conservation of Fluid Mass) Equation*

$$\frac{\partial r^m u}{\partial r} + \frac{\partial r^m v}{\partial y} = 0 \tag{6.22}$$

### *Conservation of Momentum in the Radial Direction (r or x)*

$$\frac{\partial u}{\partial t} + u\frac{\partial u}{\partial r} + v\frac{\partial u}{\partial y} = \frac{1}{\rho}\frac{\partial p}{\partial r} + \frac{1}{r^m}\frac{\partial}{\partial r}\left(r^m \nu_t \frac{\partial u}{\partial r}\right) + \frac{1}{r^m}\frac{\partial}{\partial y}\left(r^m \nu_t \frac{\partial u}{\partial y}\right) + S_u \tag{6.23}$$

### *Conservation of Momentum in the Vertical Direction (y)*

$$\frac{\partial v}{\partial t}+u\frac{\partial v}{\partial r}+v\frac{\partial v}{\partial y}=-\frac{1}{\rho}\frac{\partial p}{\partial y}+\frac{1}{r^m}\frac{\partial}{\partial r}(r^m\nu_t\frac{\partial v}{\partial r})+\frac{1}{r^m}\frac{\partial}{\partial y}(r^m\nu_t\frac{\partial v}{\partial y})+g\frac{\rho-\rho_r}{\rho}+S_v \tag{6.24}$$

where

$$S_u=\frac{1}{r^m}\frac{\partial}{\partial r}(r^m\nu_t\frac{\partial u}{\partial r})+\frac{1}{r^m}\frac{\partial}{\partial y}(r^m\nu_t\frac{\partial v}{\partial r})2\frac{\nu_t}{r^2}um \tag{6.25}$$

and

$$S_v=\frac{1}{r^m}\frac{\partial}{\partial r}(r^m\nu_t\frac{\partial u}{\partial y})+\frac{1}{r^m}\frac{\partial}{\partial y}(r^m\nu_t\frac{\partial v}{\partial y}) \tag{6.26}$$

Where

$u$ and v = temporal mean velocity components in the $r$ and $y$ directions, respectively;
$p$ = the general pressure less the hydrostatic pressure at reference density $\rho_r$;
$\rho$ = the fluid density;
$g$ = the component of gravitational acceleration in the vertical direction (generally, $y$ is positive upward so $g$ will be negative in the governing equations);
$\nu_t$ = eddy viscosity ($m^2/s$); and
$S$ = source term with units of acceleration (e.g., $m/s^2$).
($m$ = 1 yields the cylindrical coordinates and m = 0 with $r = x$ gives the Cartesian coordinates).

### *Conservation of Particulate Mass (Solids Transport) or Concentration*

$$\frac{\partial X}{\partial t}+u\frac{\partial X}{\partial r}+v\frac{\partial X}{\partial y}=\frac{1}{r^m}\frac{\partial}{\partial r}(r^m\nu_{sr}\frac{\partial X}{\partial r})+\frac{1}{r^m}\frac{\partial}{\partial y}(r^m\nu_{sy}\frac{\partial X}{\partial y}+r^mV_sX) \tag{6.28}$$

Where

$X$ = concentration of suspended solids (mg/L or $kg/m^3$);
$\nu_{sr}$ = the eddy diffusivity of suspended solids ($m^2/s$) in the $r$-direction;
$\nu_{sy}$ = eddy diffusivity of suspended solids in the $y$-direction ($m^2/s$); and
$V_s$ = particle settling velocity (m/s).

By using the Reynolds analogy between mass transport and momentum transport, the sediment eddy diffusivity can be related to the eddy viscosity $\nu_t$ by the formulae

$$\nu_{sr} = \frac{\nu_t}{\sigma_{sr}}; \qquad \nu_{sy} = \frac{\nu_t}{\sigma_{sy}} \tag{6.29}$$

in which $\sigma_{sr}$ and $\sigma_{sy}$ are the nondimensional Schmidt numbers in the *r*-direction and the *y*-direction, respectively. Typical values of the Schmidt number are in the range 0.5 to 1.

***Conservation of Energy (Heat).*** The equation for the temperature field is obtained from the energy equation:

$$\rho\left(\frac{\partial T}{\partial t} + u_j \frac{\partial T}{\partial x_j}\right) = \frac{\partial}{\partial x_i}\left[\lambda \frac{\partial T}{\partial x_i} - \rho\overline{u_i T'}\right] \tag{6.30}$$

where $T$ and $T'$ (°C) are, respectively, the mean and fluctuating component of the temperature and λ is molecular diffusivity (N s/m$^2$) (Hossain and Rodi, 1982; Zhou et al., 1994), $x_j$ and $x_i$ are Cartesian coordinates in tensor notation (m), $u_i$ is the liquid velocity (m/s), $u_i'$ is the turbulent component of the velocity (m/s), and $t$ is time in seconds. Important boundary conditions include heat fluxes at the surface and wall and influent temperatures.

The local fluid density and the liquid–solid density are given by eqs 6.16 and 6.15, respectively. Equations 6.22 to 6.30 with eqs 6.11, 6.15, and 6.16 can be solved for mean velocity, pressure, and temperature if turbulence correlations $\overline{u'_i u'_j}$ and $\overline{u'_i T'}$ are determined by the turbulence model.

***Turbulence Closure.*** Several turbulence models have been used for settling tanks; these include models that relate the eddy viscosity $\nu_t$ to (a) a constant, (b) turbulence KE and dissipation rate, (c) the Prandtl mixing length concept, and (d) algebraic stress equations. The most commonly used model is the k-ε turbulence model (Rodi, 1980), which relates the eddy viscosity $\nu_t$ to turbulent KE $k$ (m$^2$/s$^2$) and the turbulent KE dissipation rate ε (m$^2$/s$^3$) by

$$\nu_t = C_\mu \frac{k^2}{\varepsilon} \tag{6.31}$$

where $C_\mu$ is a nondimensional constant used in the $k$- ε model. Values of $k$ and ε are required throughout the tank to determine the eddy viscosity. Distributions of $k$ and ε are calculated from the following semiempirical transport equations (Rodi,1980):

$$\frac{\partial k}{\partial t} + u\frac{\partial k}{\partial r} + v\frac{\partial k}{\partial y} = \frac{1}{r^m}\frac{\partial}{\partial r}\left(r^m \frac{\nu_t}{\sigma_k}\frac{\partial k}{\partial r}\right) + \frac{1}{r^m}\frac{\partial}{\partial y}\left(r^m \frac{\nu_t}{\sigma_k}\frac{\partial k}{\partial y}\right) + P\varepsilon + P_2 \tag{6.32}$$

and
$$\frac{\partial \varepsilon}{\partial t}+u\frac{\partial \varepsilon}{\partial r}+v\frac{\partial \varepsilon}{\partial y}=\frac{1}{r^m}\frac{\partial}{\partial r}\left(r^m\frac{\nu_t}{\sigma_\varepsilon}\frac{\partial \varepsilon}{\partial r}\right)+\frac{1}{r^m}\frac{\partial}{\partial y}\left(r^m\frac{\nu_t}{\sigma_\varepsilon}\frac{\partial \epsilon}{\partial y}\right)+C_1\frac{\varepsilon}{k}PC_2\frac{\varepsilon^2}{k} \tag{6.33}$$

where $P$ is the production of turbulent energy by the mean velocity gradients as

$$P=\nu_t\left[2\left(\frac{\partial u}{\partial r}\right)^2+2\left(\frac{\partial v}{\partial y}\right)^2+2\left(\frac{u}{r}\right)^2 m+\left(\frac{\partial u}{\partial y}+\frac{\partial v}{\partial r}\right)^2\right] \tag{6.34}$$

$$P_2=\frac{\nu_t}{\sigma}\frac{g}{\rho_r}\frac{\partial(\rho-\rho_r)}{\partial y} \tag{6.35}$$

The nondimensional $k$-$\varepsilon$ model constants, $C_1$, $C_2$, and $C_\mu$ as well as the turbulent Prandtl numbers for $k$ and $\varepsilon$, $\sigma_k$ and $\sigma_\varepsilon$ are given by Rodi (1980) as follows: $C_1 = 1.44$, $C_2 = 1.92$, $C_\mu = 0.09$, $\sigma_k = 1.0$, and $\sigma_\varepsilon = 1.3$. The buoyancy correction source $P_2$ involving the flux Richardson number in the $k$-$\varepsilon$ model is sometimes omitted as a first approximation (DeVantier and Larock, 1986, 1987).

The 3-D equations are obtained by including $z$ or $\theta$ directional contributions to advection, diffusion, and shear stresses. A common assumption is that the vertical momentum equation can be simplified by the hydrostatic pressure approximation.

**DRIFT-FLUX MODELING.** In fluids composed of multiple phases (e.g., fluid/particle, fluid/bubble, or fluid/droplet mixtures, having multiple components with differing densities), it is observed that the components can assume different flow velocities. Different velocities arise because of density differences between constituents, resulting in different responses to applied forces. Often, the differences in velocities can be very pronounced, for example, large rain drops falling through air or gravel sinking in water. Under some conditions, however, relative velocities are small enough to be described as a "drift" of one component through the other. Examples are dust in air and silt in water.

The "drift" distinction has to do with what controls the relative velocity between components. If the inertia of the relative motion can be ignored and the relative velocity reduced to a balance between a driving force (say a body force and/or pressure gradient) and an opposing drag force between the components, then we speak of a drift-flux approximation.

In a simplified way, a drift-flux approximation describes adequately the relative movement between components in a mixed liquor (*i.e.*, between suspended solids and the carrying fluid) because the relative velocity between components is small.

As input for clarifier modeling, a drift-flux model requires specification of influent suspended-solids concentration (expressed as a density), the definition of a maximum solids concentration, and the determination of a coefficient value that controls the rate of separation (i.e., solids settling). The value of this coupling coefficient can typically be determined from the results of settling tests performed in the laboratory. Kinnear (2002) used the two-phase drift-flux approach in his 1-D SST model.

**ONE-DIMENSIONAL MODELS.** There are two commonly used 1-D models for clarifiers: (1) the state point analysis model and (2) the multilayer 1-D model.

***State Point Analysis.*** The application of the state point analysis was introduced in Chapter 4. The following is a brief overview of the mathematical basis for state point analysis (also referred to as the limiting solids flux method). This procedure is commonly used to determine the surface area for the preliminary design of new SSTs and/or to determine the operating point for SSTs, that is, setting the RAS ratio and determining corresponding maximum solids flux, MLSS, and the equilibrium RAS suspended solids.

Solids-flux analysis is a well-established procedure for determining limiting solids loading or the required surface area for a clarifier. There are several simplifying assumptions for this method, for example, sludge withdrawal is assumed to be one dimensional with no shortcircuiting; flow and solids accumulation are assumed to be at steady state; the Vesilind equation is assumed to apply at the critical flux boundary; and compression or two-phase effects are assumed to be negligible. Furthermore, ESS is neglected.

The solids-flux procedure involves a field determination of the settling velocity of the MLSS as a function of concentration. Typically, the function that is used is the Vesilind equation (eq 6.10). Figure 6.12 shows an example of a settling velocity–concentration curve. Figure 6.13 (see Metcalf and Eddy, 1991) shows an idealized settling volume with a hypothetical solids-flux boundary boundary in which $S_{FL}$ = solids flux limit (kg/m$^2$·h); $\alpha$ = RAS ratio; $U_R$ = underflow rate (m/h); and $X_R$ = concentration of suspended solids in the underflow (kg/m$^3$).

The total solids-flux through this boundary consists of two parts:

(1) $S_{Fg}$ = solids-flux due to settling alone (gravity) (kg/m$^2$·h) = $V_sX$ (6.36)

(2) $S_{Fu}$ = solids-flux due to underflow = $U_RX$ where $U_R$ = under flow rate (m/h) (6.37)

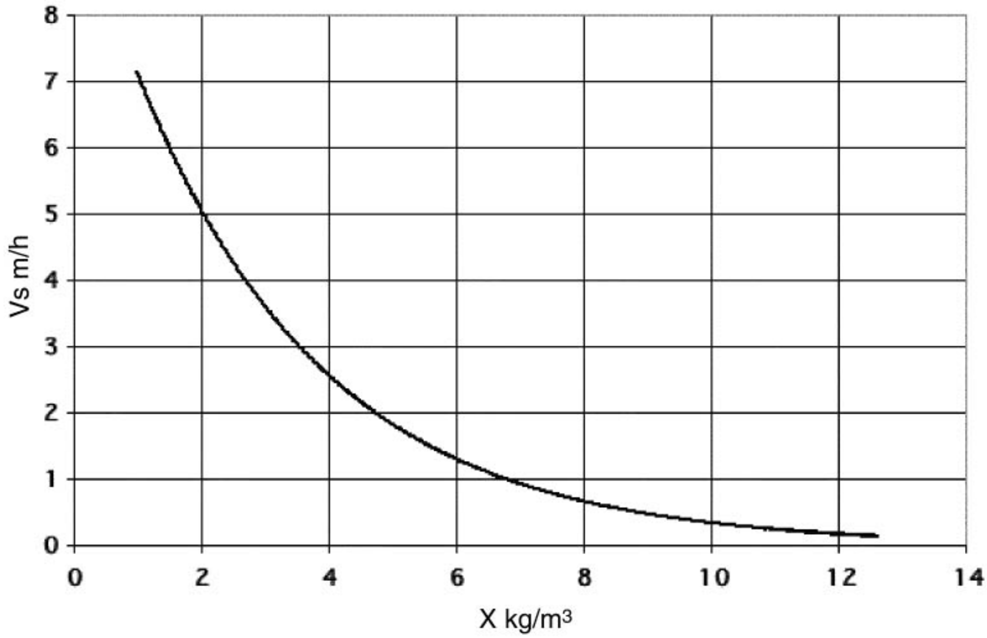


**FIGURE 6.12** Vesilind settling velocity curve for zone settling of MLSS ($V_o$ = 10 m/h and $K$I = 0.34).

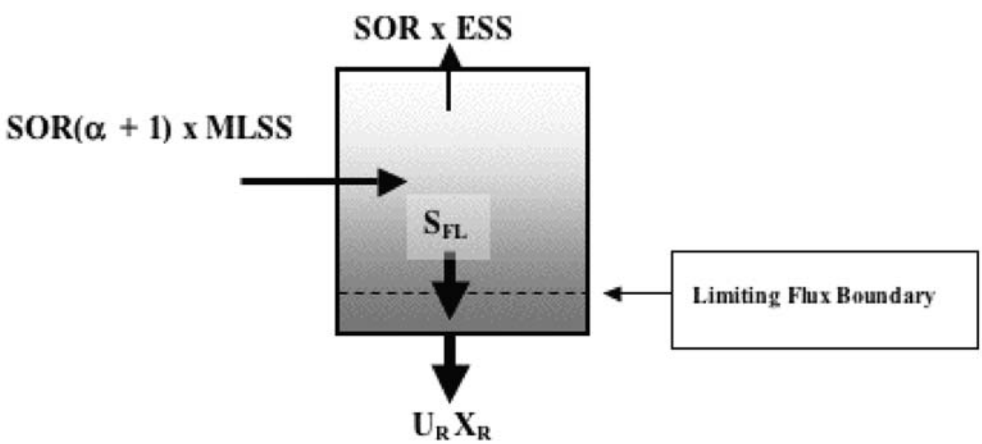


**FIGURE 6.13** Definition for solids-flux analysis (adapted from Metcalf and Eddy, 1991).

Where

$X$ = the concentration of solids at the boundary (kg/m$^3$);
$V_s$ = the settling velocity in m/h corresponding to the concentration $X$;
$U_R = Q_R/A_s$ = the underflow velocity in m/h $X$ = the concentration of solids at the boundary;
$V$s = the settling corresponding to the concentration $X$; and
$U_R = Q_R/A_s$ = the underflow velocity ($A_s$ = the surface area of the tank and $Q_R$ = the underflow or RAS flow).

The total solids-flux exiting this internal boundary is

$$\mathrm{S_{Ft} = S_{Fg} + S_{Fu}} \tag{6.38}$$

Where

$S_{Ft}$ = total solids flux = mass flux (kg/m$^2$·h);

If we introduce the Vesilind equation into eq 6.38, we obtain

$$\mathrm{S_{Ft} = V_o\, e^{-K1X}\, X + U_R\, X} \tag{6.39}$$

where $V_o$ and $K_1$ are the parameters in the Vesilind equations as obtained from the zone settling tests. For a specified $U_R$ , eq 6.39 gives a curve of $S_{Ft}$ versus $X$, similar to that shown in Figure 6.14 in which $S_{Ft}$ is the total solids flux, $X$ is the suspended solids concentration in (mg/L), $S_{Fg}$ is the solids flux due to gravity, and $S_{Fu}$ is the solids flux due to the under flow. This figure shows that eq 6.39 has a minimum value of $S_{Ft}$, which is called the limiting solids flux $S_{FL}$. Because $U_R = \alpha$ SOR, we can write eq 6.39 as

$$S_{Ft} = V_o\, e^{-K1X}\, X + \alpha\, \mathrm{SOR}\, X \tag{6.40}$$

Mathematically $S_{FL}$ is the solution of

$$dS_{Ft}/dX = 0 \tag{6.41a}$$

or

$$V_o\, e^{-K1XL}\, (X_L\, K_1 - 1) - \alpha\, \mathrm{SOR} = 0 \tag{6.41b}$$

which can be solved for the limiting concentrations $X_L$ (g/L or kg/m$^3$). The solution can be completed graphically or by using the solvers in most spreadsheet softwares, for example, "Goal Seek" in Microsoft Excel. The limiting flux then becomes

$$S_{FL} = V_o\, e^{-K1XL}\, X_L + \alpha\, \mathrm{SOR}\, X_L \tag{6.42}$$

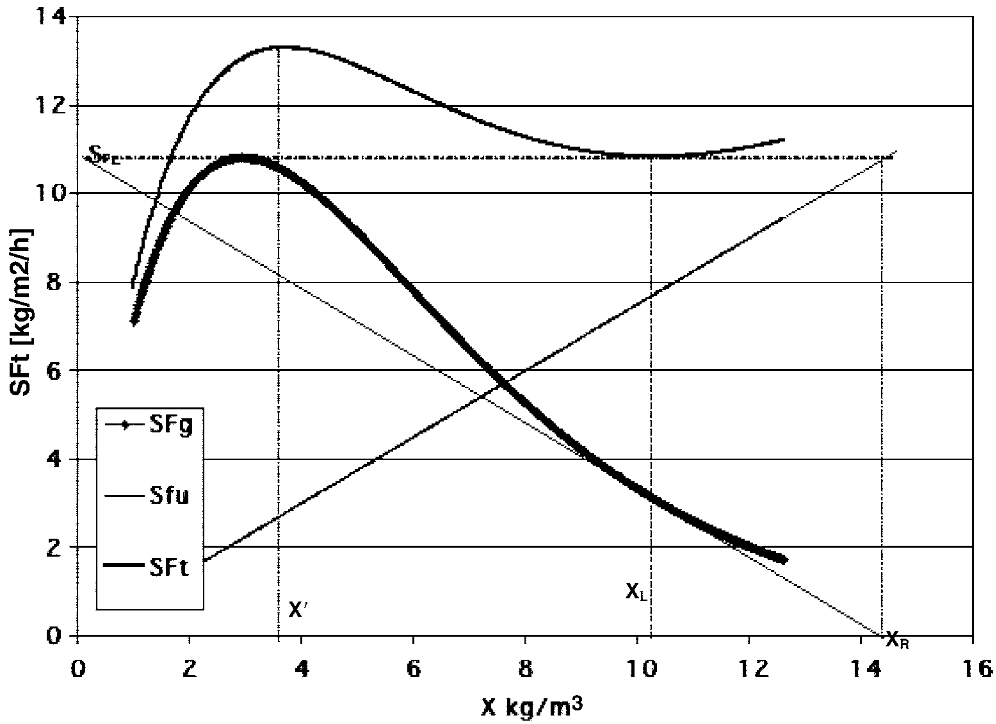


**FIGURE 6.14** Solids flux curve for SOR = 1.5 m/h, $\alpha = 0.5$, $Kl = 0.34$ m$^3$/kg, and $V_o = 10$ m/h.

and the underflow (RAS) concentration is

$$X_{RAS} = S_{FL}/(\alpha \text{ SOR}) \tag{6.43}$$

Then the MLSS at the limiting flux is given by

$$\text{MLSS} = \{(\alpha X_{RASc})/(1 + \alpha)\} \tag{6.44}$$

There are two possible roots for eq 6.42: (1) a local maximum value $X'$ and (2) the local minimum value $X_L$. When $X' \to X_L$, we are approaching the limiting $\alpha$ above which there is no solution for $S_{Ft}$ and $X_L$. The theoretical $X$ at which $X' = X_L$ is given by putting the second derivative of $S_{Ft}$ to zero, that is,

$$d^2S_{Ft}/dX^2 = 0 \tag{6.45a}$$

or

$$X' = X_L = 2/K_1 \tag{6.45b}$$

and a corresponding critical α of

$$\alpha_c = V_o\, e^{-2}\,(X'K_1 - 1)/\mathrm{SOR} \tag{6.46}$$

with the limiting flux of

$$S_{FLc} = V_o\, e^{-2} X' + \alpha_c\, \mathrm{SOR}\, X' \tag{6.47}$$

The underflow ($X_{RASc}$) concentration is

$$X_{RASc} = S_{FLc}/\,(\alpha_c\, \mathrm{SOR}) \tag{6.48}$$

The corresponding MLSS is given by

$$\mathrm{MLSS} = \{\alpha_c X_{RASc}/(1 + \alpha_c)\} \tag{6.49}$$

In the graphical solution, it is obvious which root is the correct one to get $X_L$; however, in computer solution it is necessary to "force" the solver to find the root corresponding to the minimum (positive second derivative). One practical way of doing this is to calculate $X' = 2/K_1$ first and use $2X'$ as a first guess at $X_L$.

Figure 6.15 shows the critical state point above which any increase in the α or the under flow rate ($U_R$) will not yield a solution for $S_{Ft}$. Figure 6.16 shows the variation of MLSS and SLR with α for SOR = 1.5 m/h = constant. Ekama and Marais (2002) separated SST behavior into two categories based on (1) sludge-handling criterion I, where the solids loading is limited by the solids flux at the state point, and (2) sludge-handling criterion II, where solids loading is limited by SOR, MLSS, and $V_s$ at $X$ = MLSS. In summary, the limiting solids flux for the two criteria are

(1) Criterion I: Limiting MLSS = $X_L\, \alpha/(1 + \alpha)$ for $\alpha < \alpha_c$ (6.50)

(2) Criterion II: Limiting MLSS = MLSS × $V_o\, e^{-K1\,\mathrm{MLSS}}$ /SOR or (6.51a)

$$\mathrm{MLSS}_L = \ln(V_o/\mathrm{SOR})/K_1 \text{ for } \alpha > \alpha_c \tag{6.51b}$$

where $\mathrm{MLSS}_L$ = limiting value of MLSS in kg/m$^3$. These criteria are illustrated in Figure 6.16 for SOR = 1.5 m/h.

The selection of the activated sludge operating parameters (recirculation ratio, sludge wasting ratio, solids retention time) can be made using recommendations presented in the literature. The operator typically obtains the best operating conditions by trial and error. The preceding state point analysis can aid in determining these relationships. A more complete model has been developed by La Motta (2004) to incorporate the activated sludge unit to the state point analysis. This part of a modeling package was submitted to the U.S. Environmental Protection Agency (U.S. EPA)

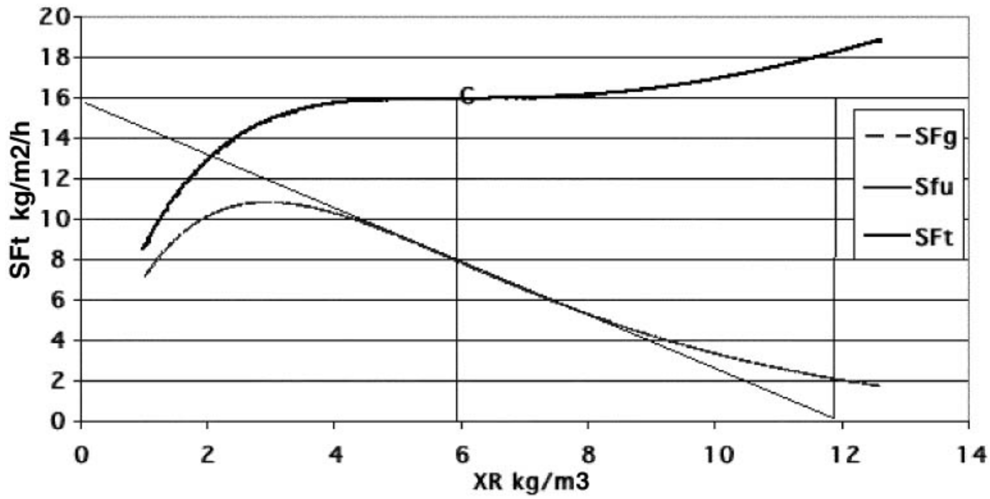


**FIGURE 6.15** Location of critical point on state point graph for SOR = 1.5 m/h, $\alpha$ = 0.90, $K$l = 0.34 $m^3$/kg, and $V_o$ = 10 m/h.

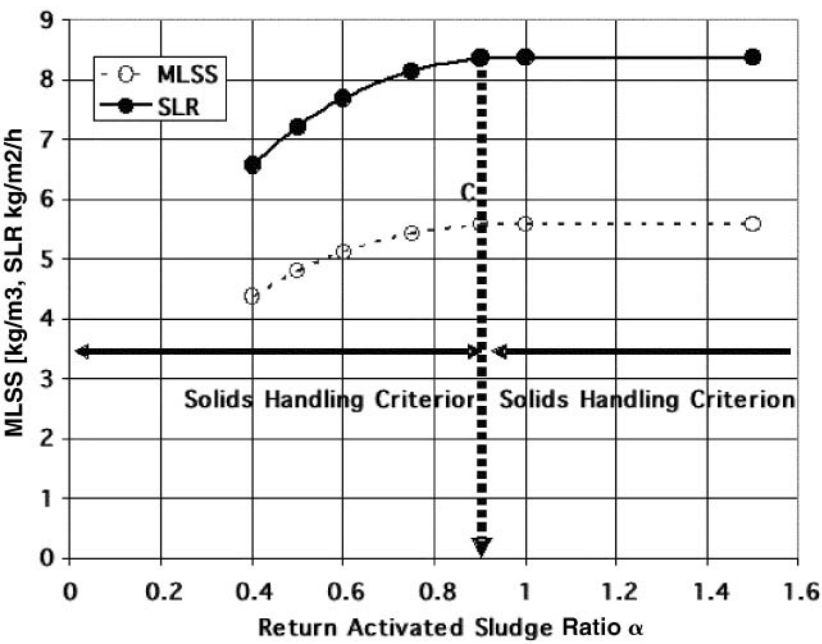


**FIGURE 6.16** Identification of solids loading criteria for SOR 1.5 m/h.

by McCorquodale et al. (2004). The defining diagram for this model is given in Figure 6.17. The model assumes 1-D, steady-state conditions and is programmed in a spreadsheet format. The model solves mass-balance equations for liquid, solids, and COD for the clarifier and the activated sludge reactor. Figure 6.18 is a sample of the spreadsheet input and solution page.

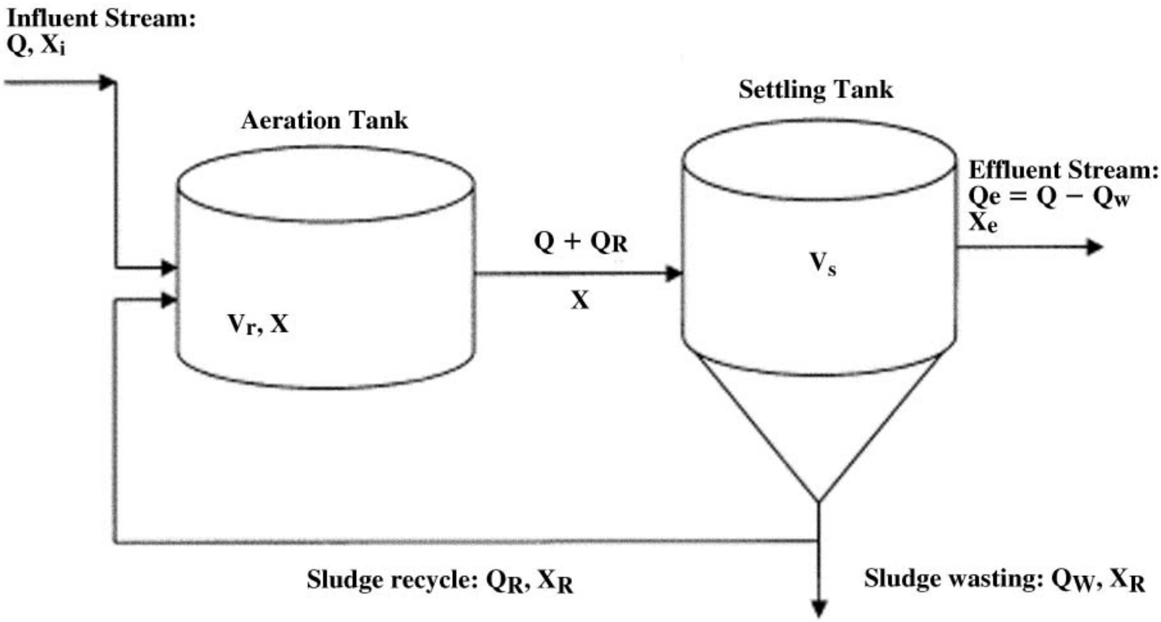


**FIGURE 6.17** Modeled activated sludge system ($A$ = settling tank area, m$^2$; $a_p$ = kinetic parameter of COD flocculation, kg/m$^3$; $a_x$ = kinetic parameter of TSS flocculation, kg/m$^3$; $C_i$ = sludge concentration, kg/m$^3$; $d$ = parameter of the correlation between PCOD and TCOD; $f_p$ = parameter of the correlation between PCOD and TCOD; $k_p$ = first-order constant of PCOD flocculation, m$^3$/d·kg; $k_x$ = first-order constant of TSS flocculation, m$^3$/d·kg; $k_g$ = first-order constant of TSS growth, m$^3$/d·kg; $k_{gp}$ = first-order constant of PCOD growth, m$^3$/d·kg; $k_d$ = endogenous respiration coefficient, d$^{-1}$; $F_B$ = batch sludge flux, kg/d·m$^2$; $F_L$ = limiting flux, kg/d·m$^2$; $n$ = (= $-k_1$) empirical parameter, m$^3$/kg ($n < 0$); $Q$ = influent flowrate, m$^3$/d; $Q_R$ = recycle flowrate, m$^3$/d; $Q_w$ = waste sludge flowrate, m$^3$/d; $r_g$ = rate of growth of suspended solids, kg SS/kgMLSS·d—which accounts for both microbial growth and growth of settleable particles due to flocculation of nonsettleable particles; $r_{gc}$ = rate of growth of colloidal particles, kg SS/kg MLSS·d; $r_{gp}$ = rate of growth of particulate COD, kg COD/kg MLSS·d; $r_{ng}$ = net rate of growth of microorganisms, kg SS/kg MLSS·d; $r_f$ = rate of flocculation of particles, kg SS/kg MLSS·d; $r_p$ = rate of flocculation of particulate COD, kg COD/kg MLSS·d; $SSS$ = supernatant suspended solids in the aerator, kg/m$^3$;

*(caption continued on next page)*

***Multilayered One-Dimensional Models.*** The treatment of clarifiers as 1-D tanks with multiple vertical layers is useful for solids inventory modeling and systems control models. Vitasovic (1985) coupled a 1-D, multilayered clarifier to an activated sludge system to demonstrate the possibility of optimizing and controlling plant operations. Similar 1-D clarifier models have been incorporated to a number of commercial codes for wastewater treatment trains, for example, the Bio-Win™ and GPX™ models. The 1-D model permits very short execution times, which make it feasible to test and optimize tank volumes and operational variables.

Figure 6.19 shows a typical arrangement for a 1-D clarifier model. The important features of this model are

(a) Discretization of the cylindrical tank into $N$ layers where $N$ is a user-defined number horizontal layer in the vertical column,

(b) Selection of an influent layer.

(c) All layers below the influent layer are assigned the UFR.

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*(Figure 6.17 caption continud)*

$(SSS)_R$ = supernatant suspended solids in the recycle line, kg/m$^3$; $S_P$ = particulate COD concentration in the aerator, kg/m$^3$; $S_{Pi}$ = particulate COD concentration in the influent stream, kg/m$^3$; $S_{po}$ = particulate COD concentration in the aerator corresponding to $\bar{t}$ = 0; $S_{Pr}$ = particulate COD concentration in the sludge return line, kg/m$^3$; $S_T$ = total COD concentration in the aerator, kg/m$^3$; $S_{Ti}$ = total COD concentration in the influent stream, kg/m$^3$; $S_{To}$ = total COD concentration in the aerator corresponding to $\bar{t}$ = 0; $S_{TR}$ = total COD concentration in the sludge return line, kg/m$^3$; $\bar{t}$ = hydraulic retention time, $V_r/Q$, d; $\bar{t}$ = solids retention time, d; $U$ = rate of uptake of dissolved COD, kg COD/kg MLSS·d; $V_r$ = reactor (aerator) volume, m$^3$; $V_s$ = settling tank volume, m$^3$; $w$ = sludge wasting ratio, $Q_w/Q$; $X$ = MLSS concentration in the aerator, kg/m$^3$; $X_i$ = suspended solids concentration in the influent to the aerator, kg/m$^3$; $X_e$ = suspended solids concentration in the final effluent, kg/m$^3$; $X_R$ = suspended solids concentration in the recycle line, kg/m$^3$; $X_w$ = suspended solids concentration in the waste stream, kg/m$^3$; $X$ = MLSS concentration in the aerator, kg/m$^3$; $X_i$ = suspended solids concentration in the influent to the aerator, kg/m$^3$; $X_e$ = suspended solids concentration in the final effluent, kg/m$^3$; $X_R$ = suspended solids concentration in the recycle line, kg/m$^3$; $X_w$ = suspended solids concentration in the waste stream, kg/m$^3$; $Y$ = true yield coefficient, kg biomass/kg DCOD consumed; $v_i$ = zone settling velocity of sludge at concentration $C_i$,, m/d; $v_0$ = settling velocity parameter, m/d; and $\alpha$= the recycle ratio = $Q/Q_R$) (after La Motta, 2004).

**CALCULATION OF RECIRCULATION RATIO AND SLUDGE WASTING IN ACTIVATED SLUDGE SYSTEMS**

***ENTRY: A blank if variable is unknown or value if it is known.***

**Variables that can be solved for:**

| | | |
|---|---|---|
| n = (-K1) | **-0.412** | **$m^3$/kg** |
| MLSS=X = | **3.524** | **kg/$m^3$** |
| $\alpha$ = | **0.23** | |

**Information Provided**

| | | |
|---|---|---|
| Settling tank area, A = | 0.44 | $m^2$ |
| Side water depth, D = | 0.27 | m |
| Vs = | 0.15 | $m^3$ |
| Q = | 6.171 | $m^3$/d |
| Vr = | 0.18 | $m^3$ |
| Vs/Vr = | 0.83 | |
| $\bar{t}$ = | 0.028 | d |
| $v_o$ = | 263 | m/d |
| Xo = | 0.063 | kg/$m^3$ |
| Xi = | 0.164 | kg/$m^3$ |
| $S_{Ti}$ = | 0.335 | kg/$m^3$ |
| $S_{T0}$ = | 0.175 | kg/$m^3$ |
| $f_p$ = | 0.71 | |
| Y = | 0.5 | |
| $k_d$ = | 0.06 | 1/d |
| $k_p$ = | 916 | $m^3$/kg·d |
| $k_x$ = | 290 | $m^3$/kg·d |
| $k_g$ = | 67 | $m^3$/kg·d |
| $k_{gp}$ = | 59 | $m^3$/kg·d |
| $a_p$ = | 0.016 | kg/$m^3$ |
| $a_x$ = | 0.010 | kg/$m^3$ |

**Predicted Values**

| | | |
|---|---|---|
| ST = | 0.026 | kg/$m^3$ |
| ESS=Xe ~ | 0.016 | kg/$m^3$ |
| $\bar{t}_c$ = | 0.50 | d |
| SFL = | 60.73 | kg/d·m 2 |
| XR = | 17.257 | kg/$m^3$ |
| w = | 0.02 | |
| Qw = | 0.13 | $m^3$/d |
| QR = | 1.42 | $m^3$/d |

FIGURE 6.18 Spreadsheet layout for coupled state point analysis (after La Motta, 2004).

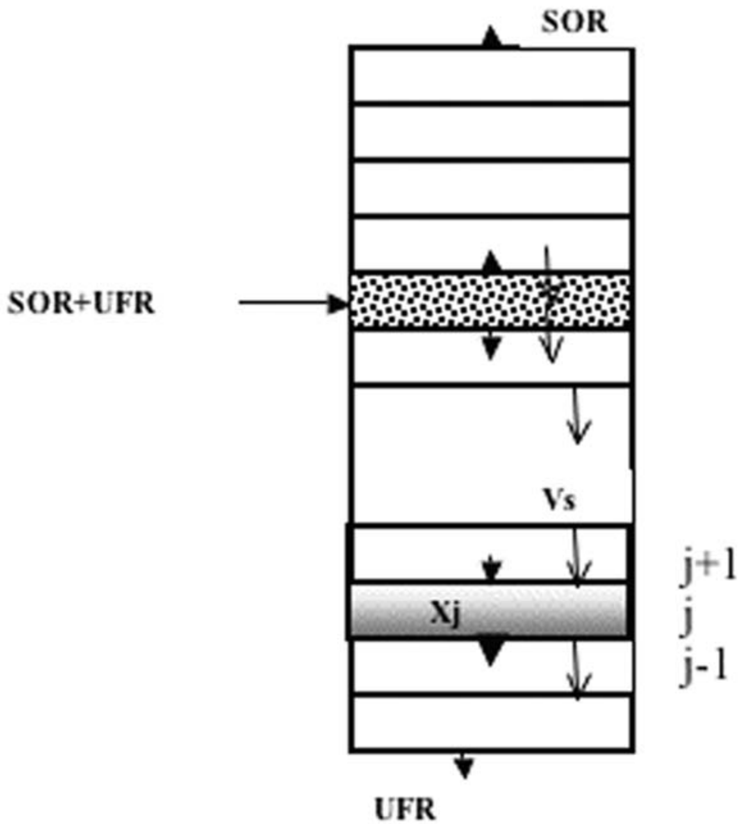


FIGURE 6.19 One-dimensional clarifier model.

(d) All layers above the influent layer are assigned a velocity equal to the SOR.

(e) The transfer of solids from layer to layer is determined by the imposed fluid velocities and the settling or slip velocity, which is often represented by the Vesilind equation (eq 6.10) or Takacs equation (eq 6.11).

(f) The concentration in each layer is computed by applying the mass-balance equation to that layer, that is,

$$dX_j/dt = \{X_{tj}(V - V_{stj}) - X_{bj}(V - V_{sbj})\} \tag{6.52}$$

Where

$X_j$ = solids concentration (kg/m$^3$);
$X_{tj}$ = concentration at the top of layer $j$ (kg/m$^3$);
$V_{stj}$ = settling velocity at the top of layer $j$ (m/h);
$X_{bj}$ = concentration at the bottom of layer $j$ (kg/m$^3$);
$V_{sbj}$ = settling velocity at the bottom of layer $j$ (m/h);
$V$ = SOR for $j > j_{\text{influent}}$ (m/h); and
$V = -$UFR for $j < j_{\text{influent}}$ (m/h).

**NUMERICAL METHODS.** There are three commonly used procedures for converting the governing partial differential equations to a discrete form of simultaneous algebraic or simultaneous ordinary differential equations. These methods are finite elements (FEM), finite volumes (FVM), and finite differences (FDM). The FEM uses an unstructured grid consisting of "elements" or subareas such as triangles or quadrilaterals of arbitrary size. Finite element method grids have the advantage that they can be constructed to fit very complex geometries. A potential problem with some FEM formulations is that they are only globally conservative. The FVM involves integrating the governing equations over the finite control volumes that make up the flow field. The FVM grids are typically structured grids; for example, the cells may be rectangular, orthogonal, or nonorthogonal quadrilaterals. This method is locally and globally conservative. The FDM is based on a Taylor Series expansion (Yakowitz and Szidarovszky, 1986) of the derivatives in the governing equations about discrete grid points. These grid points are arranged in a structured fashion, typically on a rectangular mesh. The FDM is known to have poor conservation properties and its use has declined. The mesh density is controlled to some extent by the geometric detail needed to adequately describe tank boundary conditions. The FVM may require more nodes than a corresponding FEM; however, the algebraic equations derived from the FVM are typically more readily solved than the FEM equations.

Finite difference, finite volume, and finite element methods have been applied to numerically formulate computer codes to solve eqs 6.22 to 6.34. For details of these numerical schemes, the reader is referred to DeClercq (2003), Gerges and McCorquodale (1997), Patankar (1980), Schamber and Larock (1981), Smith (1985), and Versteeg and Malalasekera (1995). Most codes are second-order accurate; that is, the truncation error is of the order of the grid size squared ($\Delta x^2$). A few codes are third-order accurate. Upwinding of the advected variable, which assumes that all information is coming from the upwind cell, is often used to obtain a more stable solution. First-order upwinding, such as in the Hybrid Method (Versteeg and Malalasekera, 1995), can lead to artificial (numerical) diffusion. Neglecting the transfer of information from all of the neighboring cells in the upwind direction can also lead to numerical diffusion. Second- and third-order skew upwind schemes address both of these sources of numerical diffusion. Gerges (1997), Gerges and McCorquodale (1997), and Zhou et al. (1993) presented discussions of these errors with an application to SSTs. Third-order methods are computationally more expensive than lower-order methods; however, for some applications (for example, dye simulation), the virtual elimination of artificial diffusion may justify the increase in computation time. In first- and second-order schemes, error as a result of numerical diffusion can be reduced by using a finer grid. Users of numerical codes should conduct sensitivity tests by varying the grid resolution and time step to ensure that solutions are independent of grid mesh size and time step.

## COMMERCIAL COMPUTATIONAL FLUID DYNAMICS PROGRAMS

**INTRODUCTION.** Throughout all engineering fields, numerical simulations of fluid behavior have historically been provided by universities and research institutions. Recently, however (owing, in part, to the development of the personal computer), commercial computing tools developed for the analysis and study of complex fluid flow problems have become more widely available and their use by practicing engineers is on the rise.

Computer algorithms for simulating fluid motions in three dimensions are referred to as CFD programs. Computational fluid dynamics programs are designed to produce simulations of fluid flows influenced by a wide variety of physical processes (for example, heat conduction, solidification, cavitation, and surface tension). Because these programs are based on the fundamental laws of mass, momentum, and energy conservation, they are applicable to almost any type of flow process. For this reason, CFD programs are referred to as "general purpose" solvers.

The roots of CFDs in the United States may be traced back to original developments at the Los Alamos National Laboratory (Los Alamos, New Mexico) beginning in the early 1960s. Many basic numerical techniques originated there for the solution of compressible and incompressible flow problems. Of particular interest are techniques for describing the behavior of free surface flows (Marker and Cell [MAC, SOLA], Volume of Fluid [VOF]), a technique for solving both compressible and incompressible flow problems with a single solution method (Implicit Continuous Fluid Eulerian [ICE]), and new types of grid and geometry models (Particle in Cell [PIC], Fluid in Cell [FLIC], staggered grids, Arbitrary Lagrangian Eulerian [ALE]). For more information, the reader is referred to http://www.lanl.gov/orgs/t/t3/history.shtml.

Widespread commercial use did not begin until the early 1980s, when CFD analysis was adopted by the aerospace industry for solving external flow problems (aerodynamics) and for designing fuel control systems (sloshing).

Today, CFD is used extensively by engineers in many fields. Hydraulic engineers engaged in the design of wastewater treatment facilities have used CFD to (a) predict the performance of hydraulic structures, (b) estimate flow distributions and mixing, (c) develop retrofits to existing facilities, and (c) troubleshoot and test design alternatives before construction.

**COMMERCIAL PROGRAMS.** Commercial CFD programs are offered by a number of companies worldwide. The oldest CFD companies in the United States are Flow Science, Inc. (Santa Fe, New Mexico), and Fluent, Inc. (Lebanon, New Hampshire). The largest CFD company in the United Kingdom is AEA Technologies (Harwell and Glasgow, United Kingdom). Training and technical support is included in the purchase and/or lease price of products offered by these organizations.

All commercial CFD programs are capable of solving flow problems involving complex geometries in two or three dimensions. Special preprocessing and postprocessing tools (included with most CFD packages) aid in problem setup and the analysis of results. Turbulence models and other physical models can be activated by the user when necessary. Clarifier modeling, for instance, requires the use of an additional physical model known as a drift-flux, or algebraic slip model (ASM). This algorithm is used to account for the settling of suspended material in the mixed liquor and to calculate spatially varying fluid densities within the clarifier being modeled.

**ADVANTAGES AND DISADVANTAGES OF COMMERCIAL COMPUTATIONAL FLUID DYNAMICS MODELS.** The results of a clarifier model using a commercial CFD program provide (a) estimates of fluid velocities within the clarifier, (b) estimates of total suspended-solids concentrations within the clarifier, and (c) an estimate of effluent suspended-solids concentration. Furthermore, the models may be operated in a steady-state or transient mode and used to gauge the performance of alternative designs (Richardson et al., 2000).

Deciding whether to acquire a commercial CFD program can be difficult. The lease and/or purchase price of this type of software is high. It is also difficult to find experienced personnel well versed in the operation and use of CFD programs. Training is, however, typically provided by most CFD vendors for free or for a nominal cost.

Lack of access to source code is another drawback of using a commercial software package. Most CFD programs are released with some source code available; thus, making it possible to make small changes to the program. The ability to access all source code routines or to make sweeping changes to the programs is not generally possible. This can be a problem if, for instance, a bug affects a part of the program that is not accessible by the user. Then, in this case, the user must rely on the vendor's ability to correct the problem in a timely fashion.

A simplified settling model is another shortcoming of many commercial CFD programs. The drift-flux or ASMs described earlier do not account for hindered settling and predict behavior that differs from that predicted by established settling models. However, the settling algorithms in commercial programs can be easily modified/upgraded and, for many problems, the results are relatively insensitive to the particular settling model used (that said, for some problems the ability to control the settling algorithm can be critically important).

Despite these drawbacks, today's commercial CFD programs are (a) flexible, (b) well tested, (c) well documented, (d) supported by knowledgeable staff, and (e) easy to use (for a well-trained engineer). Because commercial CFD programs are in a state of continuous development, it is clear that their use and reliability will increase with time. And, given recent increases in computing power, it will be possible to address more and more challenging problems with them in the future.

## APPLICATIONS OF COMPUTATIONAL FLUID DYNAMICS MODELS

**APPLICATION OF COMPUTATIONAL FLUID DYNAMICS MODELS TO PRIMARY SETTLING TANKS.** Hazen (1904) developed the earliest model for discrete settling that is typically assumed to occur in grit tanks and PSTs. Camp

(1946, 1952) and Dobbins (1944) introduced analytical solutions that allowed vertical mixing to be included in a Hazen-type model. Abdel-Gawad and McCorquodale (1984) used a strip integral model for the hydrodynamics and solids transport in PSTs. They classified influent solids in several distinct classes, including a nonsettleable class (McCorquodale and Bewtra, 1979). Their model was an improvement on the semianalytical models of Camp (1946) and Dobbins (1944) because realistic nonuniform velocity profiles could be simulated. Celik et al. (1985) applied the $k$-$\varepsilon$ model with a finite volume approach to simulate FTCs in Imam's (1981) neutral density clarifier. Recently, Parker et al. (2000) applied a CFD model to replace the strip integral model of Abdel-Gawad (1983). When PST models are supported by column settling tests, very good predictions of removal efficiency can be achieved. Density currents as a result of suspended solids can occur in PSTs as noted by Adams and Rodi (1990) but, typically, these are weaker than in SSTs. Thermal density currents can be relatively more important in PSTs than in SSTs (McCorquodale 1976, 1977, 1987; McCorquodale et al., 1995).

**APPLICATION OF COMPUTATIONAL FLUID DYNAMICS MODELS TO SECONDARY SETTLING TANKS.** ***Brief Historical Review of Two- and Three-Dimensional Clarifier Modeling of Secondary Settling Tanks.*** Secondary settling tanks have two functions: clarification of the wastewater and thickening of the sludge for return to the reactor or the waste stream. The performance of SSTs is determined by multiple factors, including sludge-settling properties, solids loading to the tank, hydraulic loading rate, sludge recycle rate, sludge density, tank geometry, variability of influent temperatures (TDS), influent total dissolved solids, and atmospheric conditions. The flow field in clarifiers has been found to be far from ideal. The velocity distribution is influenced by several factors such as density stratification caused by the presence of suspended solids. In general, because of the importance of density terms, solids transport cannot be decoupled from hydrodynamics in modellng SSTs.

The first numerical model for an SST was introduced by Larsen (1977). Schamber and Larock (1981) used the finite element technique with the $k$-$\varepsilon$ turbulence model to simulate neutral density flow in a settling tank. Imam et al. (1983a, 1983b) introduced a finite difference model with an eddy viscosity that was calibrated using experimental FTCs. Lyn and Zhang (1989) presented a 2-D numerical model for predicting turbulent flow in circular clarifiers without density currents. Adams and Rodi (1990) extended the work of Celik and Rodi (1986) and used a second-order finite volume technique known as Quadratic Upwind Interpolation for Convective Kinematics (QUICK) to simulate the dye transport in two different clarifier configurations. Szalai et al. (1994) advanced the work of Lyn and Zhang (1989) by taking into consideration the swirl effect. They used a low numerical diffusion technique referred to as a second-order Hybrid-Linear Parabolic Approximation (HLPA) and verified their results with the experiments of McCorquodale (1976). DeVantier and Larock (1987) presented a finite element model for stratified, turbulent, steady, 2-D flow. Sediment-driven density currents were simulated. Flow in the inlet zone was not modeled. McCorquodale et al. (1990, 1991) introduced a numerical model for unsteady flow in a circular clarifier for two cases: (1) diurnal variation in flow at a constant MLSS concentration and (2) a sudden increase in MLSS. The model included a description of density currents in the settling zone only. Lyn et al. (1992) present a model that included density currents. Zhou and McCorquodale (1992a, 1992b, 1992c) presented models based on the HYBRID (combine central difference and first-order upwind treatment of the advective transport terms) approach, which considered both density currents and the inlet zone. Gerges and McCorquodale (1997) introduced a finite volume numerical model based on a skew third-order upwinding scheme (STOUS). They simulated the dye experiments of Imam et al. (1983a, 1983b) and compared the FTCs predicted by STOUS and HYBRID, which showed that HYBRID suffers from severe numerical diffusion. Krebs (1991) and Krebs et al. (1999) applied CFD modeling to investigate the role of inlet geometry. A recent paper by Lakehal et al. (1999) presents an investigation of shear flow at the sludge blanket in SSTs. Zhou et al. (1997) applied a 3-D model to successfully aid in solving a performance problem for a large rectangular clarifier.

The CRTC field protocol (Wahlberg et al., 1993) was applied in a performance evaluation of two circular secondary clarifiers located in the South Secondary Complex at Metro Wastewater Reclamation District's (MWRD's) Central Treatment Plant, Denver, Colorado (Wahlberg et al., 1995). The protocol from this study provides a standard for field testing of clarifiers. Vitasovic et al. (1997) used the Denver dataset to calibrate and validate a 2-D model.

Equations 6.22 to 6.35 represent the 2-D form of the single-phase CFD equations of a clarifier. These equations can readily be extended to three dimension flow by adding another momentum equation and including additional advective and diffusion type terms in all of the equations. It is noted that these equations are based on the assumptions that the liquid–solid mixture can be treated as a homogeneous liquid. This assumption is valid when the volume of solids is very small compared with the total volume of liquid. This may not be the case in the sludge blanket, where the volume fraction of solids can exceed 1%.

Clarifier hydraulic regimes differ significantly in tanks with different sludge inventories. An existing clarifier that has good performance with a shallow sludge blanket does not necessarily give the best performance in operations with a large sludge inventory. The effect of any design detail on clarifier behavior depends on the operating conditions that include (1) SOR; (2) RAS; (3) a combination of high MLSS and poor SVI; (4) relatively low flow; and (5) combination of MLSS, settling, and compression properties ($V_o$, $K1$, sSVI). A modification that does not work well under one operational condition may have a beneficial effect on tank performance under second loading combination. A CFD model is one tool to achieve an optimized clarifier design with respect to cost and/or efficiency under the site conditions. A 2- or 3-D clarifier model can be used to investigate the effect of design or operation changes on tank performance. Computational fluid dynamics models have been used to study the effects of any combination of the following clarifier modifications: influent structures (for example, energy dissipating inlet [EDI] or influent momentum dissipating inlet [IMD], flocculation baffle [center well], and canopy baffle); clarifier effluent structures (for example, in-board launder, finger launders, and Stamford baffle); internal baffles (for example, midradius or Crosby baffle and perforated baffle); and sludge-withdrawal systems (sloping floor, scrapers, and rapid sludge withdrawal such the Tow-Bro system).

**GUIDELINES FOR SELECTION OF DESIGN FEATURES FOR CLARIFIERS.** ***Inlet Structures.*** The optimization of the flocculating zone design involves selecting the best depth and the position for the flocculating baffle (skirt), often in combination with an EDI or IMD. In large rectangular clarifiers, perforated baffles have been successfully used to dissipate the momentum and KE resulting from high inflows to enhance the effluent quality and sludge compression. The optimum skirt depth depends on the solids loading, SVI, and hydraulic loading. For ADWF and shallow blankets, deep baffles tend to produce better effluent. In secondary clarifiers, which are often operated under heavy flow conditions (for example, PWWF), a deep reaction baffle may result in very poor performance for high blanket levels. The selection of the best baffle depth will be a compromise between the arrangement that works well for both ADWF and for PWWF. The volume of the flocculating well should be determined considering two factors: (1) the hydraulic detention time required for flocculation (15 to 20 minutes) and (2) the effect of the baffle location on tank hydrodynamics and effluent TSS. Fortunately, these two considerations often lead to a similar solution. A poorly designed flocculating zone may cause a clarifier to lose 10 to 25% of its capacity or lead to increases in effluent suspended solids of 15 to 35%. A clarifier model could be used to test the performance of tanks with different flocculating zones (different depths and baffle locations). This test requires the incorporation to the clarifier model of a flocculation submodel; otherwise, only the effect of the baffle on the tank hydrodynamics can be studied. Krebs et al. (1992) used a model to show that porous walls could increase clarifier efficiency.

***Clarifier Effluent Structures.*** Effluent launder modifications can have a significant effect on clarifier performance. A good launder system can give more evenly distributed effluent flow. The effect of launder modifications is greatest for cases with a very deep sludge blanket. Launder modification is a promising way to increase clarifier capacity.

For clarifiers with a peripheral effluent weir, another promising option is the addition of a Stamford baffle. There is some empirical and modeling evidence that Stamford baffles can reduce the effect of flow rebound at the effluent weir and, therefore, have a positive effect on effluent suspended solids concentration. A clarifier model can be used to assess and compare the behavior of a clarifier with and without the Stamford baffle.

***Sludge Drawoff Facilities.*** Under heavy solids loading conditions, there may be benefits to optimizing the sludge-removal mechanism. According to field measurements (Albertson and Okey, 1992), a poorly designed sludge-removal facility could reduce the tank capacity by 15 to 30%. An optimized sludge-removal mechanism could significantly reduce the possibility of short-circuiting or watery sludge caused by insufficient sludge-transport ability. An optimized sludge-transfer facility could achieve rapid sludge removal and a more concentrated sludge blanket, thus resulting in higher tank capacity and better effluent quality. The parameters that can be investigated in a model include the size of the scraper blade, the interval between the scraper blades, the moving speed of the rake arm, and the number and the location of the sludge hoppers. DeClercq (2003) showed that the non-Newtonian nature of the flow has a significant effect on sludge withdrawal. Chain and scraper systems, sludge suction tubes, or Tow-Bro systems can be tested by using a 3-D clarifier model under different sludge withdrawal flowrates.

Ekama et al. (1997) identified short-circuiting of influent MLSS to the sludge hopper as a potential cause of increased blanket levels and degraded performance. Canopy baffles have been used to reduce this effect (Krebs et al., 1995, 1999).

In some process retrofit projects, one of the objectives is to improve flow distribution and RAS. It may be required to evaluate flow pacing of the RAS based on the clarifier influent flow. In a secondary treatment process, increasing RAS flow can lower the RAS draw-off concentration required to achieve an equilibrium state. On the other hand, the higher recycle flow rate may reduce sludge compression in the tank because of the stronger turbulence and shorter solids detention time. Any modification of the RAS pumping system could be expensive. Therefore, it is important for design engineers to know the effect of RAS flow variations on process performance for a given range of process operating conditions. Modeling results can be used to determine optimum RAS flows for different clarifier loading conditions and solids settling properties.

***Clarifier Water Depth and Bottom Slope.*** Field data show that an improperly designed clarifier bottom slope or side water depth can result in a large reduction of effective tank capacity compared with a good design. For many years, design engineers have determined clarifier water depth based on their experience because conventional design theory based on the assumption of piston flow is not able to consider any effect of water depth on clarifier capacity or performance. This empirical approach is valid only for a certain range of process and operational conditions. Once process or operation is outside of the range of experience, the risk of a poor design increases.

The review of many clarifier designs in Europe and Australia indicates that some design factors that need more evaluation are steep tank bottom slopes (20 to 40%) and very shallow side water depths (approximately 2 m). Steep bottom slopes may offer stronger sludge flow towards the center sludge hopper and reduce the burden of the sludge scraper system. However, the reverse sludge flow may be too strong to maintain favorable sludge compression. Although the shallow side water depth could somewhat reduce the turbulence in the tank, the sludge-storage capacity is primarily dependent on the depth.

***Modification Packages and Cost-Effectiveness Analysis.*** The beneficial effects of clarifier modifications are not always additive. Furthermore, the benefits of modifications that work well under one operating condition may have the reverse effect under a different operating condition.

In addition to the screening study of individual modifications, the clarifier model can be used to test the effect of combinations of modifications to obtain the highest efficiency. Clarifier modeling helps design engineers to implement cost-effective clarifier modifications. Modeling can reduce the construction and/or maintenance costs while enhancing clarifier performance.

***Storing Biosolids Temporarily in Aeration Basins During High Flow.*** Consultants often have to look at the modifications needed for storing biosolids temporarily in aeration basins under wet weather conditions. The effect of possible modifications on system performance could be evaluated by using a coupled activated sludge model. Ji et al. (1996) presented a paper describing a coupled activated sludge secondary clarifier model. This model was applied during the design phase of the Utoy Creek WWTP (Atlanta, Georgia) upgrade project to investigate different options for managing solids during design wet weather loading.

Adding an aeration tank component, the 2- or 3-D fully mass conservative clarifier model can be used to simulate the whole secondary treatment system. The basic principle is that, for a given process design (mass inventory in aeration tank), the system model can accurately simulate mass distribution and effluent quality under diurnal flow or wet weather conditions. In a selected process, modeling results can help the designer to get the best combination of aeration tanks and settling tanks (thus, the highest cost effectiveness for whole system) rather than the best results for each component.

***Optimization of Construction, Operation, and Overall Cost.*** In a secondary treatment process, system effluent quality is affected by clarifier geometry and hydraulics, solids settling properties, influent concentrations, and flows. To improve the performance of an existing clarifier, many alternatives (with different costs and efficiencies) can be considered, such as those described in this section. Using modeling results, the overall cost, including both construction and operational cost, could be optimized. The final goal of a modeling study is to provide the basic information to the project owner to aid in decisions related to secondary treatment, for example, clarifier expansion or modifications to sustain target loading conditions. The results could also help to establish a reliable basis for cost-effectiveness analysis in the selection of tank modifications.

***Assessment Aspects of the Clarifier Performance.*** Evaluations of clarifier performance, based on clarifier modeling, should include two significant aspects: clarifier capacity and clarifier effluent quality. In the modeling study, tank capacity is measured and expressed in terms of hydraulic and solids loading that clarifiers are able to accommodate. If the tank operation can achieve an equilibrium state and tank clarification is functioning well, imposed solids loading is considered to be the sustained load of the clarifier. When the tank operation approaches an equilibrium state, most of the influent mass is withdrawn from the RAS flow and the mass accumulation in the tank is negligible, that is,

$$Q_o X_o = Q_R X_{RAS} + Q_{eff} X_{eff} \tag{6.53}$$

Where

$Q_o$ = influent flow (e.g., $m^3/h$),
$X_o$ = MLSS concentration (mg/L),
$Q_R$ = RAS flow (e.g., $m^3/h$),
$X_{RAS}$ = RAS concentration (mg/L),
$Q_{eff}$ = effluent flow (e.g., $m^3/h$), and
$X_{eff}$ = effluent concentration in mg/L.

In some cases with peak flow (or solids) loading conditions, clarifier operation may not reach equilibrium even at the end of the event. Peak flow is still considered to be sustained as long as the clarifier effluent concentration does not exceed permit levels during the given peak flow period. In a clarifier stress test with a sludge-thickening problem, the clarification failure occurs when the effective clarifier storage capacity is exceeded. However, the rate of the mass accumulation depends on such factors as sludge compression, loading conditions, tank storage, clarifier hydraulic efficiency, efficiency of sludge-withdrawal facilities, and short-circuiting of influent to the RAS.

Some indicators of clarifier performance are effluent quality, sludge compression, RAS concentration, sludge inventory in the tank, and the hydraulic behavior of the clarifier (for example, from hydraulic residence times). Tank effluent quality is simply defined as the averaged ESS concentration (clarification). Sludge compression is evaluated by examining the sludge blanket level and the RAS draw-off concentration (thickening). Also, in some cases, the sludge compression ratio (SCR) is used. The definition of the SCR is the RAS withdrawal concentration divided by the sludge compression standard. The sludge compression standard, which is the concentration required to achieve the *equilibrium state* (eq 6.53), is only dependent on influent flow and MLSS and RAS flow. The sludge-withdrawal concentration, which is the actual RAS concentration removed in the underflow, is dependent on various tank design features and operation factors such as flow pattern, strength, turbulence intensity, sludge settling property, influent MLSS, and RAS draw-off mechanism.

Using detailed clarifier modeling results, the hydraulic behavior of the clarifier can be assessed by considering the following:

1. Relationship between the sludge blanket level and effluent solids concentration;
2. Effect of clarifier hydraulic loading (and solids loading) on effluent concentration, RAS concentration, sludge blanket level, and sludge inventory;
3. Variations of flow patterns with respect to the sludge blanket; and
4. Solids distribution within the tank.

***Field Validation of Computational Fluid Dynamics Models for Secondary Settling Tanks.*** An essential part of the acceptance of a numerical model is validation using field observations. Once the model has been calibrated for a given clarifier, it should be validated by running the model with an independent input dataset; the results should be compared with the observations for the event corresponding to the dataset. At this stage, there should be no additional adjustment of any of the calibration parameters. For example, a clarifier may be calibrated for existing conditions and validated with field tests for a new internal geometry. Two examples of model validation are given here.

The CFD high accuracy clarifier model (HACM) (Gerges, 1997) was used to study the behavior of two plants that were exhibiting unreliable performance. The model was run and calibrated for the existing conditions. It was then used to simulate several proposed changes. Selected modifications to the tanks were made and a new set of field data were collected. The model results were then compared with these field results. Borkman et al. (2004) using HACM found that installation of a 1.5-m deep baffle at approximately 4 m from the inlet of a rectangular clarifier at the Irvine Ranch Water District WWTP (Irvine, California) would result in an improved effluent and more consistent performance. The baffle was installed and the plant ESS decreased from an average of 21 mg/L to less than 5 mg/L. The improvement was attributed to improved flocculation and an improved flow pattern.

Bodeaux and Gerges (2004) reported their experience with a center-feed "squircal" clarifier at Fairfield-Suisun Sewer District (Suisun, California). The CFD model (HACM) was again used to investigate several alternatives. The model indicated that a deeper inlet skirt (flocculating center well) and a 1.5-m peripheral baffle could improve tank hydraulics and reduce ESS. Subsequent installation and side-by-side field testing confirmed what the model had predicted, showing a more than 40% reduction in ESS.

# PRACTICAL EXAMPLE OF THE APPLICATION OF COMPUTATIONAL FLUID DYNAMICS MODELS

**CIRCULAR CLARIFIERS.** Figure 6.20 shows an idealized secondary clarifier used to illustrate some benefits of CFD modeling.

The 3-D clarifier model (Zhou et al., 1992, 1997) was applied to evaluate a design of a circular clarifier that has a center-feed and peripheral effluent weir configuration. The tank dimensions are shown in Figure 6.20 (diameter of clarifier = 36.6 m, side wall depth = 4.47 m, and almost flat bottom). The tank influent flow is introduced in the center feedwell through vertical slots on the wall of a vertical feed pipe with a diameter of 1.1 m. The top of the slots is submerged by 0.76 m. The clarifier influent jet directly enters the flocculation well in the existing design. The radius of influent flocculation well in the existing design is 3.74 m and the submerged depth of the skirt is 2.4 m. Double center-mounted rotational Tow-Bro sludge-withdrawal tubes are used.

***Modifications.*** An influent momentum dissipating column (MDC) with a radius of 1.3 m and depth of 2.15 m is proposed to surround the existing influent column. The total slot space is 40%.

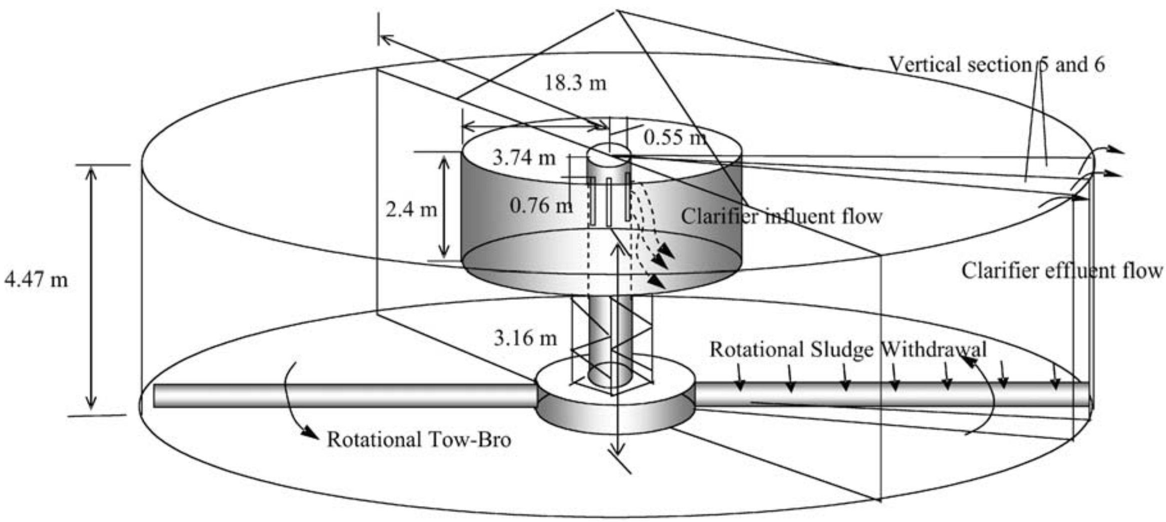


**FIGURE 6.20** Idealized center-feed clarifier.

***Performance.*** Figures 6.21(a) to 6.21(d) present the velocity vectors at sections 5 and 6 (see Figure 6.20) under the ultimate flow condition (SOR of 2.2 m/h) combined with MLSS of 2800 mg/L in tanks with and without an MDC. In the clarifier simulations with no MDC, the forward momentum is more concentrated near the bottom under the flocculation well compared to the case with an MDC. This is caused by the greater effect of the influent on the skirt when there is no MDC. Because of the deep sludge blanket, this results in a stronger upward (buoyant) deflection of the influent flow downstream of the flocculation skirt. To further show the efficiency of MDC on momentum dissipation, Figures 6.22 and 6.23 present the effect of MDC on the horizontal velocity components. The results were taken along the water depth right after the perforated column in sections 5 and 6 for an ultimate SOR of 2.2 m/h. The maximum horizontal velocity reduces from 21 to approximately 15 cm/s because of the impingement of influent flow with MDC, indicating that the MDC proposed in this study can effectively reduce influent momentum along the radial direction and more evenly distribute flow in the flocculation well under high flow conditions. For SOR = 2.2 m/h, the clarifier effluent concentration drops from 51.7 to 35.6 mg/L because of the MDC. Under a relatively low SOR of 1.0 m/h, the clarifier effluent concentration actually slightly increases from 13.2 mg/L in the existing design to 13.8 mg/L. In the clarifier with an MDC, the sludge blanket level is found to be lower than that in the clarifiers with no MDC. The influent structure, in the unmodified design, provides sufficient influent momentum to obtain good influent mixing in the flocculation well under relatively low clarifier flow conditions (SOR of 1.0). However, the structure is not able to effectively dissipate the very strong influent momentum with a very high flow condition (SOR of 2.2).

The modeling results show that the sludge inventory in clarifiers with an MDC is approximately 8% less than that in clarifiers with no MDC although the same Tow-Bro sludge suction tubes are used in both tanks for an ultimate SOR of 2.2. Reasons for improved performance are reduced resuspension and a transient reduction in the RAS concentration as a result of the greater penetration of the influent in the region of the skirt. In the simulation with an SOR of 1.0, the model predicts no significant differences of sludge inventory between clarifiers with and without an MDC. Figure 6.24 presents the clarifier ESS versus SOR in clarifiers with and without an MDC for MLSS of 2800 mg/L, sSVI of 100, and RAS of 50%. The MDC increases the longest peak flow period (SOR = 2.2) from 3 hours to approximately 5 hours for ESS $<$ 30 mg/L.

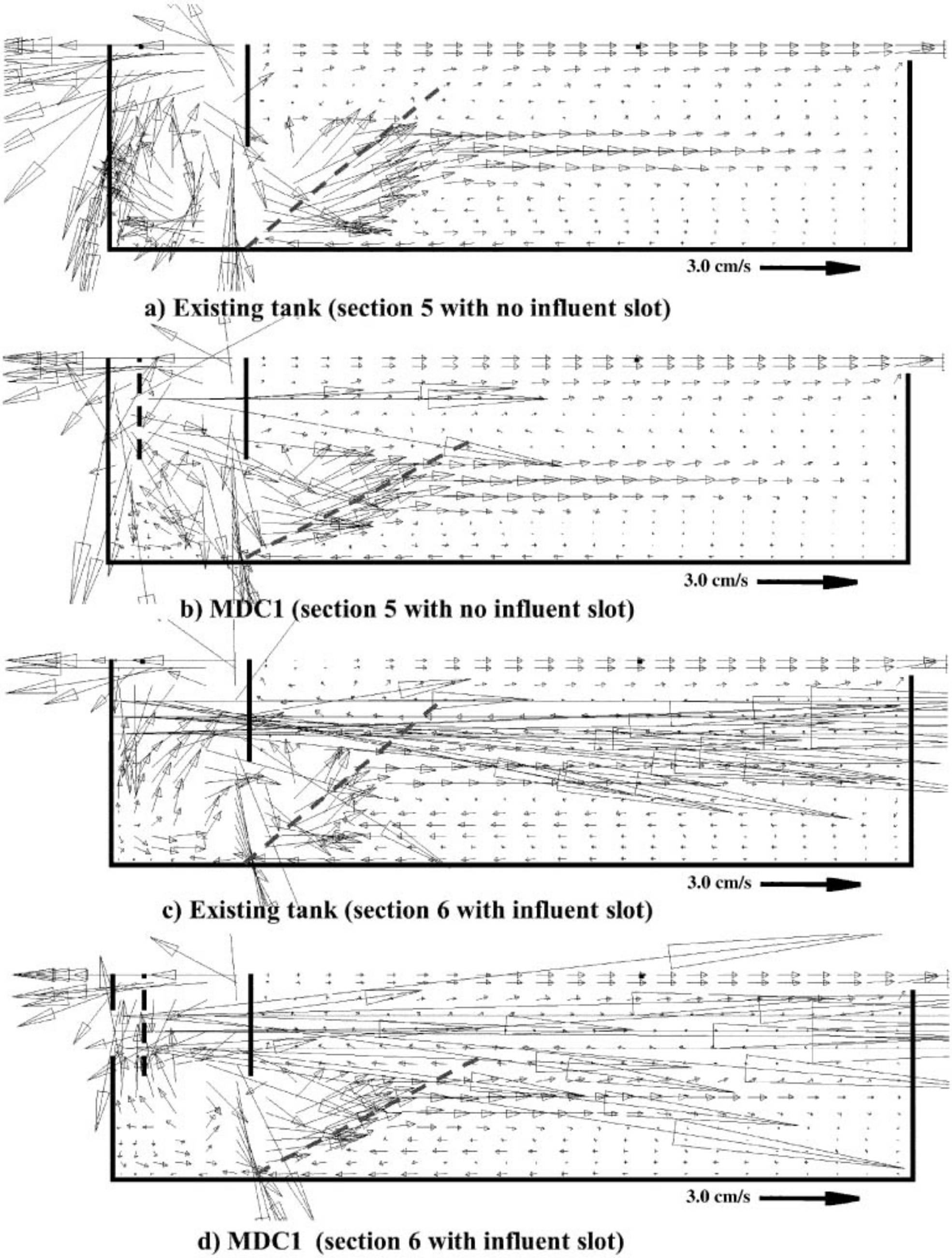


**FIGURE 6.21** Effect of MDCl on flow pattern in clarifier with ultimate SOR of 2.2 = $m^3/m^2{\cdot}h$.

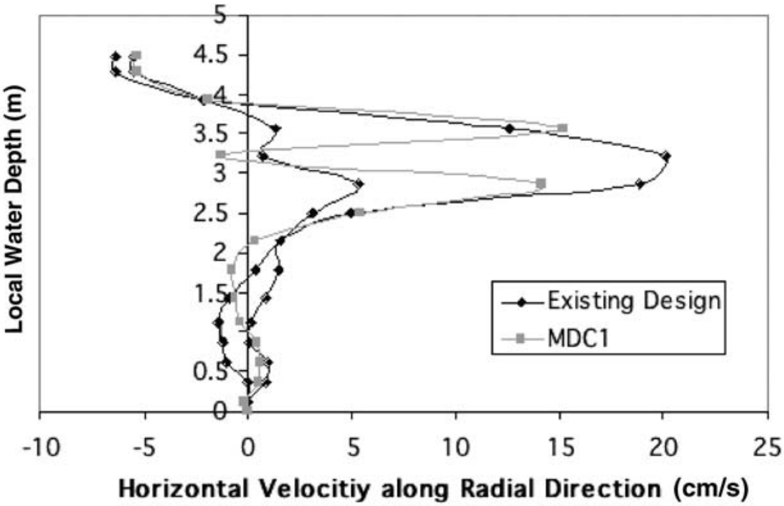


**FIGURE 6.22** Effect of MDC on horizontal velocity right after MDC in section 6 (with influent slot) for SOR = 2.2 m/h.

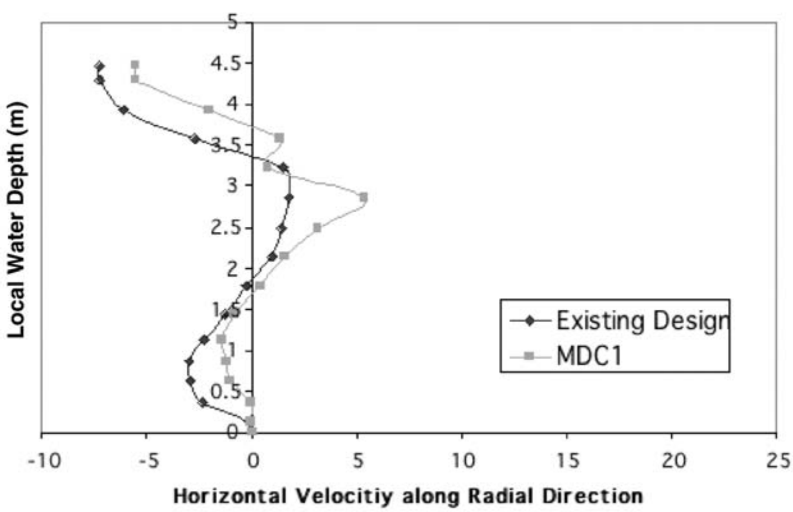


**FIGURE 6.23** Effect of MDC on horizontal velocity right after MDC in section 5 (with no influent slot) for SOR = 2.2 m/h.

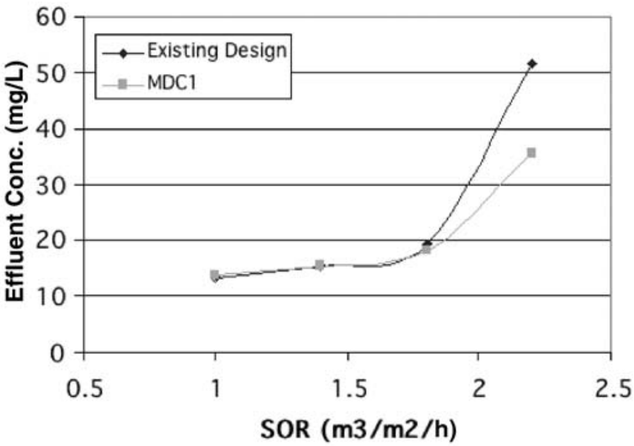


**FIGURE 6.24** Effect of MDC on clarifier capacity (MLSS = 2800 mg/L and sSVI = 100).

To investigate the design alternatives that could be used to further improve clarifier performance, the clarifier model has been used to evaluate the performance of a clarifier with a modification package, which includes an MDC plus a Stamford baffle. Figure 6.25 presents the predicted clarifier ESS as a function of SOR with and without the modification package after 5 hours of simulation. The modification package

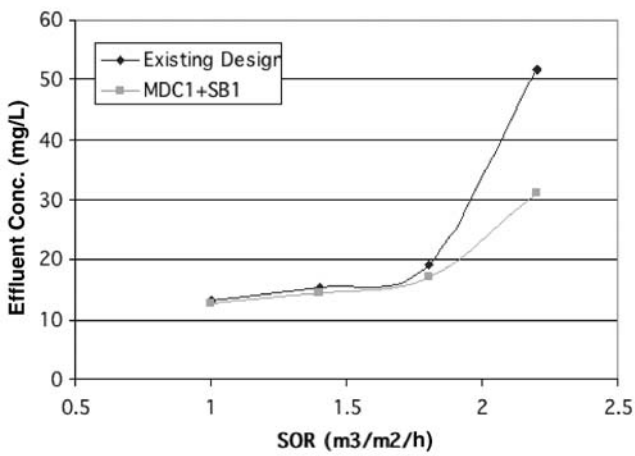


**FIGURE 6.25** Effect of modification package on clarifier capacity (MLSS = 2800 mg/L).

reduces the ESS from 52 mg/L in the existing design to 31 mg/L at the end of the simulation of 5 hours.

## LIMITATIONS AND NEW DIRECTIONS IN MODELING CLARIFIERS

All models have simplifying assumptions that limit the range of applicability of the models. The following is a summary of some common limitations and/or modeling challenges to currently available clarifier models:

- Most models assume that the liquid–solids mixture can be treated as an equivalent homogeneous liquid for the solution of momentum and continuity equations. Increasing the number of phases that are modelled will increase the computation time; however, new processor speeds and parallel architectures that are becoming available for desktop machines, make it feasible to incorporate this advancement.
- The density state equation is a very important link between solids concentration and hydrodynamic effects. At the moment, floc density is only approximated in most models. Because floc density is highly variable ($1.0 <$ dry floc specific gravity $< 1.6$), it is advisable to use actual field measurements for each site.
- Present models ignore the mixing effect of surface heat transfer and density effects related to diurnal variations in influent temperature. Most models do not include temperature as a dependent variable. This may be a serious problem in primary settling tanks where the thermal density currents are more important relative to solids-induced density effects; however, temperature effects on SSTs have been documented (Wells and LaLiberte, 1998). A new public domain model has been developed by McCorquodale et al. (2004) that includes the energy equation with heat exchange. The model shows significant night and winter upsets in ESS as a result of high radiative heat loss.
- Total dissolved solids in the influent can vary with tides and result in TDS-induced density currents. Although this can readily be included in most existing models, it is typically omitted.
- Wind shear can transfer a large amount of energy to the tanks; however, it is typically ignored.

- Advanced flocculation models are not incorporated to most CFD models. Most models are using the $K_2$ in Takacs equation (eq 6.11) as a calibration parameter. This limits the range of application of the model to the geometry and flows that were used in the calibration. Li and Ganczarczyk (1987) have measured floc settling velocities and attempted to relate these to floc size and porosity. The beta version of the model recently submitted to U.S. EPA (McCorquodale et al., 2004) includes a simplified flocculation model.
- The consolidation phase of sludge thickening is often treated as an extension of the Vesilind equation (eq 6.10); however, the zone settling tests that are used to develop the Vesilind equation may not be applicable to the sludge blanket, where two phase theory should be considered.
- More research is required to determine the effectiveness of scrapers in circular tanks. Billmeier (1988) presented a study of the effectiveness of scraper blades. DeClercq (2003) showed that scraper blade movements might result in local hydrodynamic waves that could be detrimental to clarifier performance. Scraper performance depends on the local speed, angle, sludge rheology, sludge concentration, depth, and tank slope.
- The turbulence model in most clarifier models is a form of the $k$-$\varepsilon$ model, with added damping near the sludge blanket–clarified liquid interface. Some of the assumptions in the $k$-$\varepsilon$ model are not valid in clarifiers, especially in and above the sludge blanket.
- The presence of a gas phase is ignored in present clarifier models. If denitrification occurs in a clarifier, resulting gas bubbles can cause a rising blanket and significant alteration of tank hydrodynamics.
- Most models are not coupled to other units in the treatment plant, but 1-D settling models have been successfully applied to simulated solids inventory in activated sludge treatment processes (Vitasovic, 1985). Ji et al. (1996) developed a dynamic solids-inventory model that had a CFD-based, 2-D clarifier submodel. The computational capacity of modern desktop computers makes it possible to integrate high-resolution 2- and 3-D models with biological reactors.
- Plant permits include several parameters besides TSS, for example, biochemical oxygen demand/COD, total phosphorus, ammonia, and nitrate. Future models should include all of the variables typically included in plant permits.

- Future settling tank models should be designed to model floating solids as well as settling solids.
- Clarifier models commonly neglect the biochemical processes in the clarifier. The hydraulic detention time in the clarifier is not negligible. Dissolved oxygen uptake continues in the clarifier, in particular in the sludge blanket. Modeling the sludge residence time in the clarifier would give designers and operators new insight to the best methods of sludge removal.
- There have been several recent advances in sludge rheology. The sludge can be considered as a Bingham fluid (see Bird, 1976; DeClercq, 2003; as well as http://www.tu-dresden.de/mw/ilr/lampe/bingham/bingeng.htm) with two or three parameters that are functions of the solids concentration, the mechanical history of the sludge, and the biological history of the sludge. More research is needed to relate rheological properties to biological unit processes. Gravity flow and forced transport caused by scrapers is not very well modeled by present clarifier models. Models may help to resolve questions related to what is the best sludge-withdrawal arrangement (Wahlberg et al., 1993).
- Clarifier models can only crudely predict the potential for “rat-holing”, that is, short-circuiting of diluted MLSS directly to the return sludge line. As the grid resolution in 3-D models improves, these models will be able to accurately predict “rat-holing”.

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# Chapter 7

# Field Testing


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# INTRODUCTION

The evaluation of secondary clarifiers became more important with the growing popularity of the activated sludge process in the 1940s. The observation of currents that appeared to be carrying large amounts of suspended solids to the effluent prompted Norval Anderson to begin his extensive research. Using some unique research tools, such as drogues, he was able to document the presence of a density current. He further demonstrated that this density current led to a "wall effect"—the impact of the density current with the perimeter wall, causing an upwelling of solids near the weir.

The growth of secondary treatment processes also led to many individual research efforts by manufacturers of process equipment. These efforts, in turn, brought many different variations of secondary clarifiers—rectangular, square, and circular—into popular use.

In the late 1970s, Robert Crosby began a more detailed evaluation of circular clarifiers as a part of a U.S. Environmental Protection Agency (U.S. EPA) (Washington, D.C.) research project. His study brought to light the effect of such factors as flow distribution and return sludge flow control, along with the effects of the density current. Crosby's studies also demonstrated the value of different density current baffles to improve clarifier performance. This research effort resulted in the presentation of more definitive evaluation tools, such as dye testing and sludge-blanket profiling.

In 1986, U.S. EPA and Environment Canada (Ontario) joined forces to form a Clarifier Research Technical Committee (CRTC). Composed of members who were clarifier experts from around the world, the committee's charge was to (a) determine the research needs for secondary clarifiers and (b) develop a protocol for evaluating clarifier performance. This committee later became known as the American Society of Civil Engineers–CRTC. Among the most important conclusions of the Committee was the consensus that, to determine a clarifier's capabilities, testing of an actual clarifier in the field was required.

As a part of its effort to demonstrate the usefulness of standard evaluation techniques, the committee conducted several extensive field tests of different shaped secondary clarifiers. These field tests resulted in published project reports and, eventually, a protocol for field testing (Wahlberg, 2001).

In separate field evaluation efforts, researchers have sought other tools and techniques to identify performance characteristics and document clarifier performance. Some of these tools, such as acoustical Doppler meters, are able to measure clarifier currents more precisely than other techniques, while other tools, such as hand-held fluorometers, simply provide the same data in a different manner.

## PURPOSE OF FIELD TESTING

Field testing of clarifiers has many uses. Probably the most popular reason for testing is to determine the performance characteristics of existing units, particularly their weaknesses, to identify the most appropriate upgrade features. The ultimate goal is generally to optimize the clarifier's performance while using the existing basic components.

In many cases, the objective is to improve suspended solids capture, either to meet certain total suspended solids (TSS) or turbidity limits or to achieve a certain level of phosphorus removal. Another growing need is to increase the capacity of the existing units to meet the demands of a growing population base or to treat more of the wet weather flows through the existing plant. In either case, the primary objective is to increase the clarifier capacity. This would generally necessitate some testing at the higher anticipated hydraulic loadings.

A third popular reason for conducting field tests is to determine the actual capacity of a clarifier. In this case, the clarifier is ultimately brought to the point of failure, whether through forced hydraulic or solids loadings, or both.

Occasionally, a new or modified clarifier must be tested to demonstrate overall conformance to the design performance capabilities. Although this can be accomplished, in part, by monitoring effluent quality over a period of time, it often may be accompanied by more detailed evaluation of the clarifier's hydraulic characteristics.

## INITIAL STEPS IN ANALYZING A CLARIFIER'S PERFORMANCE

One of the first steps in conducting a field evaluation is the identification of all the pertinent design details, not only for the individual clarifier, but also for the entire system. As the testing progresses to the field, it becomes more focused on actual performance characteristics, such as described in the following sections.

**DETERMINING CLARIFIER FLOWS.** Determining clarifier flows involves the following:

- Verifying full-plant flow,
- Determining the flow to each battery of clarifiers, and
- Determining the flow to and from an individual clarifier.

Many assumptions of flow and solids loadings are based on plant flow-meter readings, so the accuracy of that meter should be verified. With reliance on the full-plant flow, the flows to an individual clarifier or battery of clarifiers can often be determined by the use of simple flow measurement techniques, such as the insertion of rectangular weirs in the effluent launders. The only requirement is, in this case, to determine the relative flows to each unit.

Because the flow to a clarifier is often the sum of the through-flow and recycled flow, it is also necessary to determine the recycled flow component. This can be as simple as measuring the return activated sludge (RAS) rate from a test clarifier. At the other end of the spectrum, the entire recycled flow to a fixed-film reactor may have to be measured. The most convenient technique for measuring these flows is the "bucket and stopwatch" method; using known clarifier dimensions to constitute the "bucket", the flow to the clarifier is stopped while maintaining the return sludge or recycled flow. The rate of withdrawal is determined by measuring the rate that the water surface recedes.

**ASSESSING THE BIOLOGICAL OR CHEMICAL PROCESS PERFORMANCE.** The eventual performance capability of a clarifier is generally a function of the floc characteristics of the feed from the biological or chemical reactors preceding it. If an activated sludge has slow settling qualities or a chemical floc is poorly formed, the clarifier performance will be limited by that condition. If the existing conditions are anomalous to the normal conditions, they should be rectified before conducting capacity tests.

The ability of a biological or chemical floc to agglomerate and settle is typically measured with a settleometer, noting the settled sludge volume at various time intervals as the mixture settles. At the end of a 30-minute settling test, the suspended solids in the supernatant should also be measured as an indication of how well the mixture can clarify the wastewater. A further test of the ultimate capability of the mixture to clarify includes flocculating the mixture at, for example, 50 rpm for 20 minutes before settling. The supernatant of the flocculated sample then represents the best product of an ideal clarifier: one which flocculates and provides for quiescent settling.

In addition to the routine settling tests, a biological floc should be examined microscopically to determine the condition of the mixture and the types of biota present. If filaments are present, additional settling tests with various dilutions of the mixed liquor can determine whether or not the filaments present are limiting settling and compaction.

***Conditions for Testing for the Formation of the Floc.*** Testing for the formation of the floc involves the following:

- Determining the condition of the floc before entering the clarifier, for example, at the end of the aeration or flocculation compartment;
- Determining the condition of the floc within a flocculation zone in a clarifier; and
- Determining the condition of the floc approaching the effluent launders or in the final effluent.

The basic techniques for testing the floc formation include a visual examination of a biological floc with a microscope and jar testing the product for settleability at various concentrations of sludge and various additions of coagulant aids.

The most intense level of jar testing involves the use of a "gang stirrer", where multiple samples can be analyzed at the same time. A simpler test involves the use of a single flocculation container for determining the flocculation potential of single samples. In each case, the supernatant, after 30 minutes of settling a flocculated sample, is analyzed for suspended solids to determine its potential to clarify the wastewater.

***Testing for Activated Sludge Settling Properties.*** The settling rate and compactibility of an activated sludge are often the major determinants of a clarifier's treatment capacity. The basic test for settling rate and compaction is the 60-minute settleometer test. By relating the 30-minute settled sludge volume to the mixed liquor concentration, the "sludge volume index" (SVI) can be determined. This SVI value can then be compared to published data to determine the capacity of the clarifier to thicken the activated sludge.

A more rigorous, and often more accurate, technique uses tall, stirred settling columns to determine the settling characteristics. Called a "state point" analysis, this is a graphical analysis that uses the settling rates of various dilutions and concentrations of the mixed liquor to predict the solids loading at which the clarifier will fail. The term "fail", as used in this analysis, refers to the solids loading rate at which the sludge blanket will continue to increase in depth.

At this point in the testing, it may be obvious that the activated sludge settling characteristics have departed far enough from the expected performance value that no further testing should be performed until this problem has been overcome.

**DETERMINING INDIVIDUAL CLARIFIER EFFLUENT QUALITY.** The ability of a clarifier to produce an acceptable effluent at a given loading is the ultimate information sought. This is why testing the effluent at different loading rates, either for effluent total suspended solids (ETSS) or turbidity, is so important to an evaluation. Effluent sampling can be performed by use of manual grab samples or through the use of automatic composite samplers. The automatic sampling should preferably be sequential samplers that can be programmed to furnish discrete samples on a 30- or 60-minute basis.

This same time frame can be used to observe effluent turbidity as a real-time indication of effluent suspended solids quality. The turbidity can be measured using a portable field turbidimeter or an online turbidimeter.

**MONITORING BLANKET PROFILES AT SELECTED LOCATIONS.** One of the most important performance characteristics observed during a field evaluation is the change in the formation of the sludge blanket during the test period. At the base level of effort, a simple core sampler can be used to determine the level of the surface of the sludge blanket. Unless they are specially designed, however, core samplers are limited to providing a vertical composite of the clarifier contents. Each sample taken is analyzed gravimetrically for TSS.

Experience has shown, however, that much more intense monitoring of the sludge blanket is required to adequately characterize the performance of the test clarifier. By frequent measurement of the vertical solids profiles (VSPs), the actual concentration of the solids at various elevations and locations in the clarifier can be determined for the test period. Another way that this can be accomplished, with great effort, however, is by using a core sampler that is tapped for discrete samples at every foot. These individual samples are then taken to a laboratory for a gravimetric analysis. This can be very time-consuming and expensive, which imposes a cost limit on the amount of VSP information that can be obtained.

With the development of portable electronic suspended solids analyzers, the amount of VSP information obtained during a test can be expanded to include multiple locations along the length or radius of a clarifier, at frequent time intervals during the stress testing period. These devices have enabled the evaluator to identify failure conditions more accurately and to quantify the amount of solids present in each test clarifier as the test progresses.

**OBSERVING ETSS VARIATIONS.** The effluent TSS is the primary indicator of clarifier performance. Most plants have this information available on a daily basis from a composite sample of the plant's effluent. This existing data can be evaluated in terms of various plant operating parameters, such as

- Overflow rates, based on the average daily flow rate,
- Sludge volume index,
- Solids loading, and
- Sludge blanket levels.

A more intensive evaluation would include the use of sequential (hourly) composite samplers so that the effect of the diurnal variations of flow and blanket levels would be available.

To differentiate between the performance of multiple clarifiers, it is necessary to gather ETSS data for each clarifier. At a basic level, the comparison of a sufficient number of daily grab samples can provide a reasonable basis for evaluating individual clarifier performance. For this data, the use of turbidity data would be sufficient, if correlated for that plant to the actual TSS value.

## DETERMINING HYDRAULIC CHARACTERISTICS

**FLOW CURVE TEST.** ***General Description.*** A flow curve is graphical representation of the time that it takes for a mass of a tracer to pass through a reactor, such as a clarifier. This is the basic test of the hydraulic characteristics of a clarifier. Depending on the hydraulic regime, the flow curve produced will follow one of three basic patterns (See Figure 7.1). A plug flow regime, such as the flow in a pipe, or similar to the flow in a well-designed chlorine contact tank, will generate a flow curve similar to that in Figure 7.1a. A completely mixed reactor, such as a chemical feed tank or some aeration reactors, will generate a flow curve following the shape of Figure 7.1b. A clarifier, on the other hand, is more of an arbitrary flow reactor, with some of the tracer reaching the effluent in a rather short time frame, followed by the bulk of the tracer, with some of the dye lingering much longer than the theoretical hydraulic retention time (Figure 7.1c).

***Flow Curves in a Single Clarifier.*** For this test, a slug of a fluorescent dye is introduced at a point just upstream of the inlet structure where the dye will be well-mixed with the influent. Samples of the dye are then obtained from the effluent at time intervals

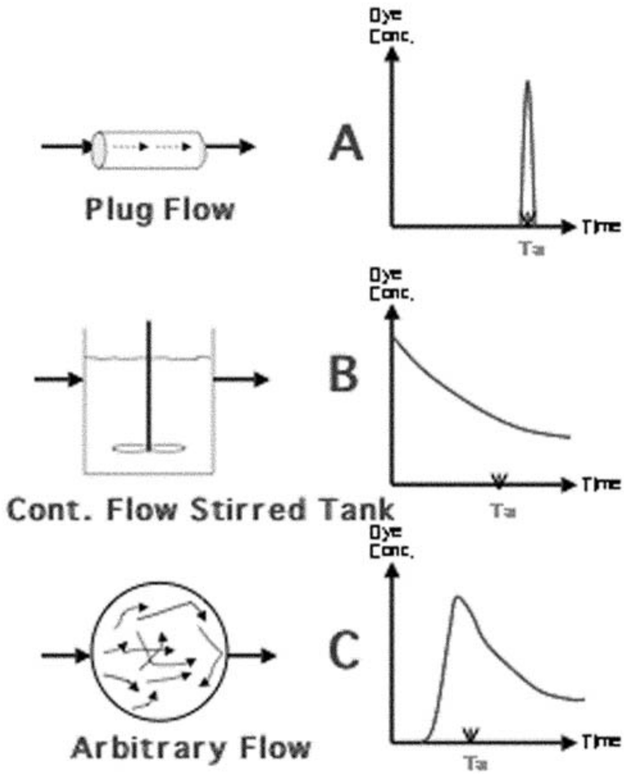


**FIGURE 7.1** Reactor configurations and flow curves: (a) plug flow, (b) continuous flow stirred tank, and (c) arbitrary flow.

that are frequent enough so that a "flow curve" can be developed comparing the effluent concentration versus time. This data can then be evaluated to determine the following:

- Time of the initial appearance of the dye,
- Time at which the peak dye concentration occurs,
- Actual (or operating) detention time as represented by the time to the centroid of the area under the curve.

Figure 7.2 depicts an example of a graphed flow curve/detention time test. With this flow curve data, comparisons can also be made with flow curves from other similar clarifiers, or even other types of clarifiers. Figure 7.3 depicts an example of a flow curve/clarifier detention time test with severe short-circuiting. Figure 7.4 depicts an

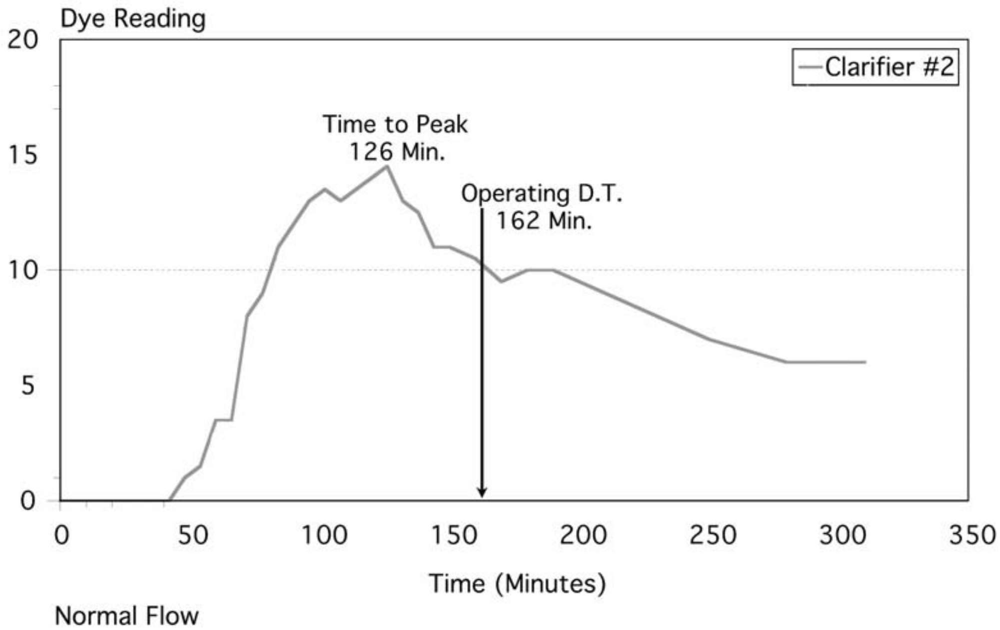


**FIGURE 7.2** Example of a flow curve/detention time test.

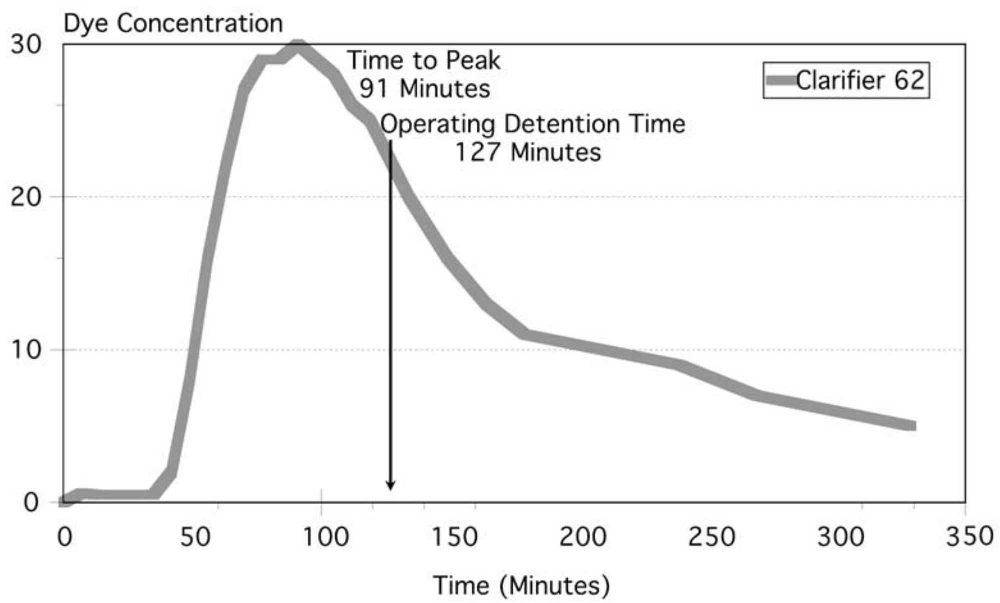


**FIGURE 7.3** Example of a flow curve/clarifier detention time test with severe short-circuiting.

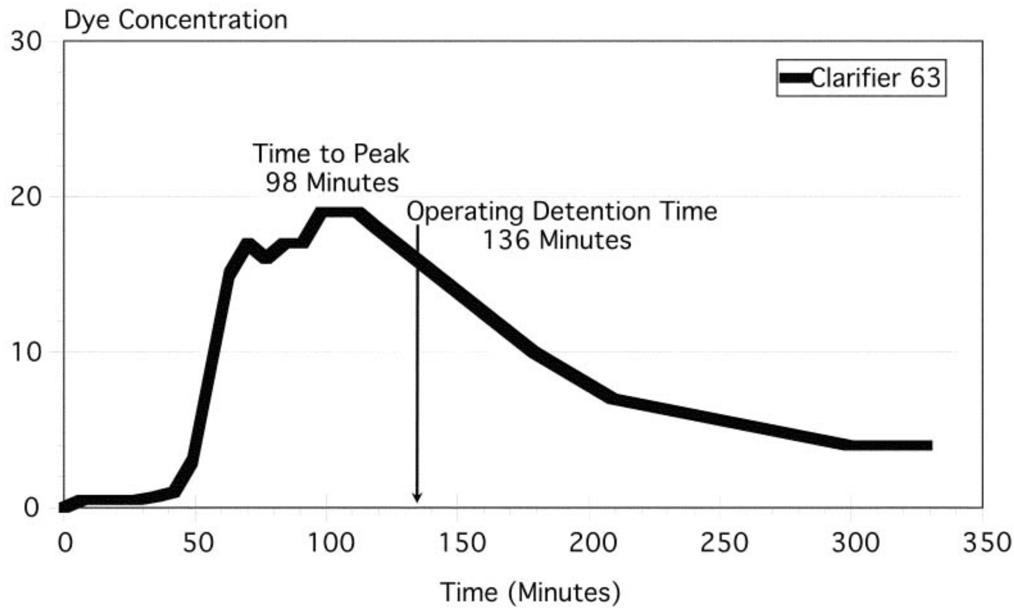


**FIGURE 7.4** Example of a flow curve/clarifier detention time test with moderate short-circuiting.

example of a flow curve/clarifier detention time test with moderate short-circuiting. The flow curve for a clarifier with severe short-circuiting is faster rising to the peak concentration and faster receding from the peak than it would be in a clarifier with lesser degrees of short-circuiting.

***Flow Curves in a Clarifier System.*** If there is more than one clarifier in the system, it is important to conduct this flow curve test in all the clarifiers at the same time. The results of this battery of tests will help identify flow differences and other general hydraulic differences in what appear to be the same clarifiers. Subsequent clarifier testing should then be focused on both the best-performing clarifiers and the poorest-performing clarifiers. Figure 7.5 depicts a flow curve/detention time comparison in a battery of clarifiers.

***Flow Curves at Different Locations.*** During a flow curve test, additional samples can be collected from individual weirs or launders, from the inside and outside of a single launder (as shown in Figure 7.6), or from both sides of a clarifier. Each of these

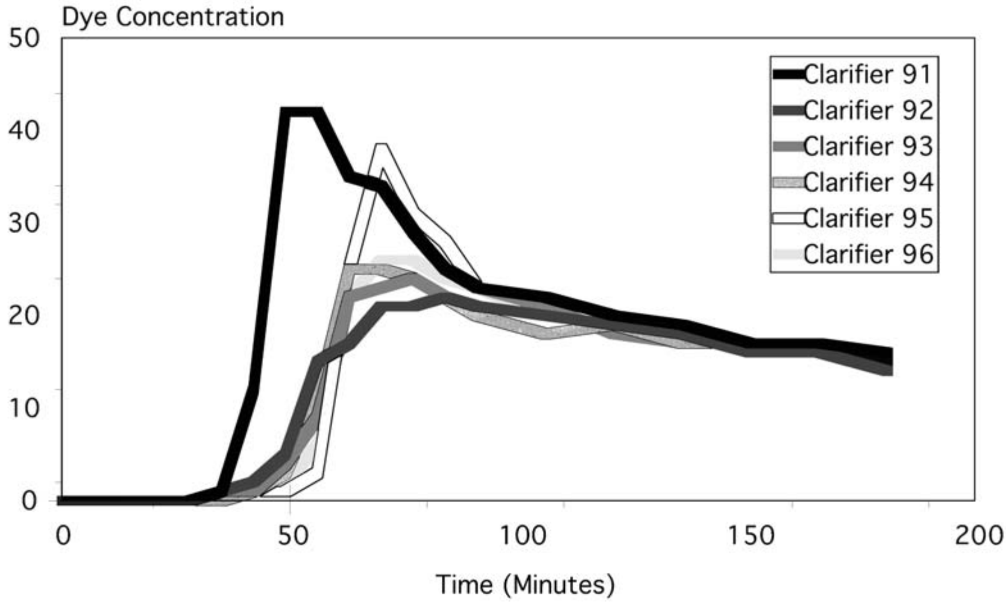


**FIGURE 7.5** Example of flow curve/detention time comparison in a battery of clarifiers.

tests will help to develop a better understanding of the clarifier's hydraulic characteristics. Figure 7.6 (Clarifier #80 launder comparison) is an example of the flow curves generated by the flow over the front and back sides of one section of the effluent launder in a rectangular clarifier.

**DYE TRACER TEST.** Along with flow curve data, it is very useful to be able to delineate the actual progress of the current through the clarifier. If the clarifier behaved as an "ideal" clarifier, the flow would cover a broad band over the full depth of the clarifier from the inlet to the effluent weirs. Instead, the actual flow follows certain defined pathways or currents. By locating and delineating these currents, better decisions can be made regarding their role in determining a clarifier's efficiency.

To define the location of the major short-circuiting current in a clarifier, a special technique, called the *dye tracer test*, can be used. For this test, a solution of dye is injected at a constant rate to a location upstream of the inlet, such as in the mixed liquor channel. For the duration of the test, the dye will naturally follow the flow

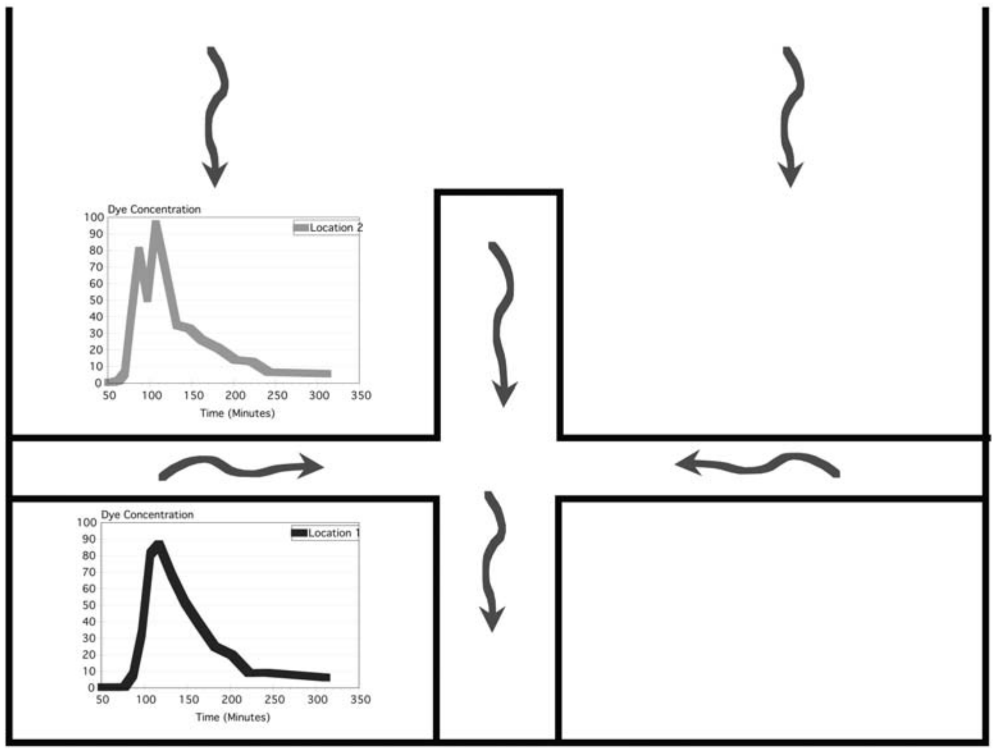


**FIGURE 7.6** Clarifier #80 launder comparison.

paths that exist in the clarifier. By sampling different vertical cross-sections of the clarifier at various time intervals, the actual flow path can be outlined by the isogonal dye concentration lines, i.e., the lines of equal dye concentrations. Figure 7.7 depicts an example of a tracer test results in a circular clarifier. Figure 7.8 depicts an example of a tracer test in a large rectangular clarifier.

**DROGUE CURRENT TEST.** The currents in a clarifier often vary considerably by depth and distance from the influent. To locate and quantify these currents, an X-vaned type of "flow catcher", called a *drogue*, can be used (Figure 7.9).

The drogue is set in a circular clarifier from the walkway or from along the side-wall of a rectangular clarifier. A lanyard is attached to the drogue to suspend it from a float at whatever depth the currents are sought. The float allows the movement of the drogue to be tracked on the surface. As the float moves through the clarifier, its

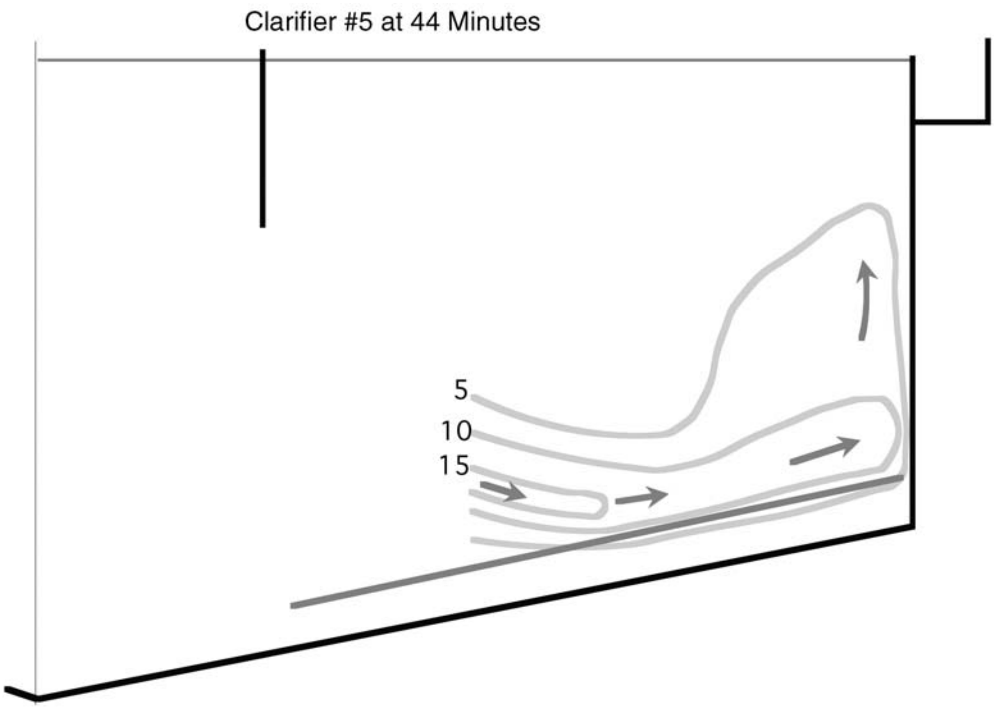


**Figure 7.7** Example of a tracer test result in a circular clarifier.

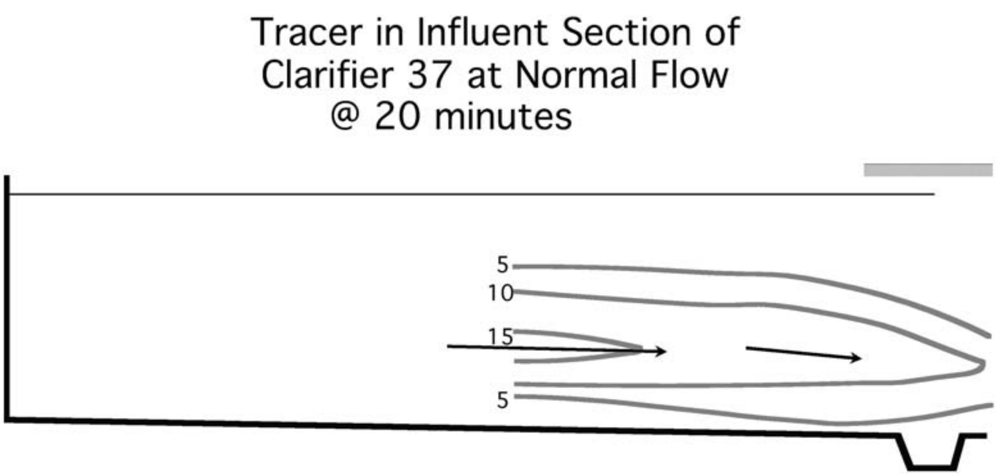


**Figure 7.8** Example of a tracer test in a large rectangular clarifier.

**FIGURE 7.9** A drogue ready for use in a rectangular clarifier.

position is noted at set time intervals so that it can later be reproduced on paper. This process is repeated at multiple depths and at different locations along the walkway or sidewall. The movement of the drogue is a relatively exact indication of the current at each depth and location. Figure 7.10 depicts a rectangular clarifier example with drogues at normal flow.

Note that the drogue movement in the circular clarifier at the −3.4 m (−11 ft), −4.0 m (−13 ft), and −4.6 m (−15 ft) elevations indicate the strong movement of the density current toward the perimeter weirs. The drogue data for the rectangular

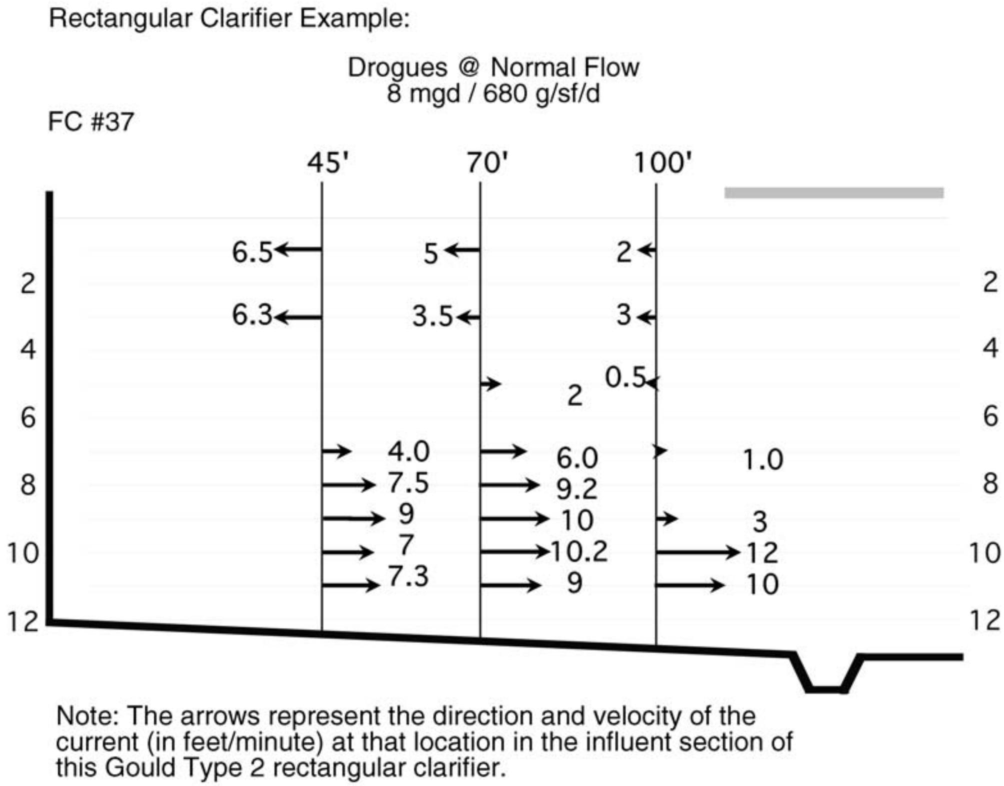


**FIGURE 7.10** Example of drogue data in a rectangular clarifier (note: the arrows represent the direction and velocity of the current [in feet per minute] at that location in the influent section of this Gould type 2 rectangular clarifier) (gpd/sq ft × 0.047 4 = m/d; ft/min × 5.080 = mm/s; mgd × 3785 = $m^3$/d).

clarifier also indicates the movement of the density current at the −2.7 m (−9 ft), –3.0 m (−10 ft) and −3.4 m (−11 ft) elevations, reaching velocities as great as 61 mm/s (12 ft/min). Note also that the drogues near the surface of the rectangular clarifier show that the current as far down as −0.9 m (−3 ft) below the surface is moving towards the influent.

**DETERMINING HYDRAULIC CHARACTERISTICS FOR DIFFERENT CONDITIONS.** Each of these tests will yield the hydraulic information for the clarifier under the operating conditions present during the test. It is generally of interest to characterize a clarifier's performance under different, and generally more stressful, conditions, such as higher influent or recycle flows or solids loadings.

Depending on the capabilities of the system, these different conditions can be obtained either by increasing the flow to the entire system or by taking one or more clarifiers offline to increase the loadings to the test clarifiers. When changing flowrates, it is helpful to slowly increase the flow rather than to increase it all at once. The changed flowrate should be allowed to stabilize in the clarifier for approximately one hydraulic retention time before starting further testing. Although opinions vary on how long the flow should be held constant before starting a test, approximately one hydraulic retention time should be sufficient for most testing purposes.

Occasionally, the temperature of the wastewater affects a clarifier's performance. This is particularly true in primary clarifiers and in clarifiers following fixed film reactors. In each of these situations, if the wastewater to be settled enters a clarifier whose contents are colder than the entering wastewater, the warmer wastewater will have a tendency to flow over the cooler clarifier contents. This is one of the few cases where temperature can affect settling.

Another case where temperature can play a profound role in settling is with above-ground steel clarifiers in warm climates. In these situations, the sun can warm the walls of the clarifier and induce rising convection currents along the southerly wall.

Return sludge flow rates can also affect a clarifier's performance. In addition to the requirement to return sludge at a rate consistent with process control needs, the effect of the RAS flowrate on the clarifier's hydraulic regime can be a factor.

## ADDITIONAL FIELD TESTS

**VERTICAL SOLIDS PROFILES.** The main purpose of a clarifier is to separate the solids in the suspension and to provide a clear effluent. While some clarifier operators choose to run their clarifiers without any sludge blanket, others either choose to operate with sludge blankets or, because of operating conditions, are unable to avoid carrying a blanket. The location and condition of this blanket can have a great effect on the clarifier's hydraulic characteristics and overall performance.

In the past, most investigators relied on the use of clear plastic core samplers to determine the presence of a sludge blanket and its interface. The typical observations had to do with changes in the blanket levels during the tests.

A more rigorous form of sludge blanket testing uses specially constructed core samplers that permit discrete sampling of the blanket concentrations at various depths. This kind of sampling provides much more information than simply reporting the blanket interface. This technique, however, is relatively expensive and time-consuming.

With the advent of portable, hand-held suspended solids analyzers, this kind of data collection has become much more efficient, and maybe even more reliable. Vertical solids profiling techniques, using these in-situ TSS analyzers, are now providing an effective insight to the formation and movement of sludge blankets during the various phases of a test. With the cost of data collection reduced to the labor factor, these devices enable an order of magnitude more VSP data than was previously possible. An example of a Royce Model 711 portable TSS and interface is shown in Figure 7.11 (Sanitaire/Royce Technologies, New Orleans, Louisiana).

During a field test, multiple VSPs are collected in the same time frame, either along the side of a rectangular clarifier or along the access bridge of a circular clarifier. The number of sampling sites is determined by the level of effort demanded by the evaluation. In circular clarifiers, it is important to have at least one sampling station within the center well. These VSP results can then be tabulated and compared to record the progress of the sludge blankets.


**FIGURE 7.11** Example of a model 711 portable TSS and interface detector (Sanitaire/Royce Technologies, New Orleans, Louisiana).

In rectangular clarifiers, the tabular VSP results permit a direct clarifier-to-clarifier comparison on a "pounds-in-the-clarifier" basis. In circular clarifiers, the comparisons can still be made, but not on a direct quantitative basis.

The instrument typically has an upper limit for recording, such as 10 000 mg/L. To permit tabulation, results over the instrument's upper limit are recorded as "10". Table 7.1 depicts an example of the results of a comparison of two rectangular clari-

**TABLE 7.1** Results of a comparison of two rectangular clarifiers. The tabular entries are the suspended solids concentrations at each depth in milligrams per liter.

| Clarifier #1 at 13 247 $m^3$/d (3.5 mgd) | | | | | | Clarifier #2 at 13 247 $m^3$/d (3.5 mgd) | | | | | |
|---|---|---|---|---|---|---|---|---|---|---|---|
| **Location and depth** | **3.0 m (10 ft)** | **9.1 m (30 ft)** | **15 m (50 ft)** | **21 m (70 ft)** | **32 m (105 ft)** | **Location and depth** | **3.0 m (10 ft)** | **9.1 m (30 ft)** | **15 m (50 ft)** | **21 m (70 ft)** | **32 m (105 ft)** |
| −0.3 m (−1 ft) | 0.09 | 0.02 | 0.02 | 0.02 | 0.02 | −0.3 m (−1 ft) | 0.17 | 0.11 | 0.09 | 0.08 | 0.07 |
| −6 m (−2 ft) | 0.13 | 0.09 | 0.07 | 0.03 | 0.03 | −6 m (−2 ft) | 1.10 | 0.09 | 0.12 | 0.08 | 0.07 |
| −0.9 m (−3 ft) | 1.82 | 0.56 | 0.08 | 0.03 | 0.03 | −0.9 m (−3 ft) | 1.89 | 0.09 | 0.13 | 0.08 | 0.08 |
| −1.2 m (−4 ft) | 2.25 | 0.55 | 0.06 | 0.04 | 0.03 | −1.2 m (−4 ft) | 3.15 | 0.13 | 0.14 | 0.08 | 0.08 |
| −1.5 m (−5 ft) | 2.48 | 1.40 | 0.11 | 0.04 | 0.03 | −1.5 m (−5 ft) | 3.33 | 2.80 | 2.50 | 0.09 | 0.08 |
| −1.8 m (−6 ft) | 3.45 | 4.00 | 6.10 | 8.60 | 6.03 | −1.8 m (−6 ft) | 3.50 | 4.70 | 3.70 | 4.40 | 7.28 |
| −2.1 m (−7 ft) | 5.45 | 8.96 | 10 | 10 | 10 | −2.1 m (−7 ft) | 7.58 | 10 | 10 | 10 | 10 |
| −2.4 m (−8 ft) | 10 | 10 | 10 | 10 | 10 | −2.4 m (−8 ft) | 10 | 10 | 10 | 10 | 10 |
| Totals | 25.7 | 25.6 | 26.4 | 28.7 | 26.2 | Totals | 30.7 | 27.9 | 26.9 | 24.8 | 27.6 |
| | Sum | >>> | 133 units | | | | Sum | >>> | 138 units | | |

fiers. The meter readings are indicated the table in milligrams per liter. Note that the amount of solids in each clarifier is relatively the same.

Table 7.2 shows the results of a comparison of another two circular clarifiers. Again, the tabular entries are the suspended solids concentrations at each depth in milligrams per liter. Although the columns and tables are summed, the totals do not directly represent the amount of solids in each clarifier. The best comparisons that can be made from this table are the actual VSP results at each of the sampling stations.

**TEMPERATURE PROFILES.** The temperature of the wastewater has only a general effect on the performance of a clarifier because of the increase in viscosity as the water temperature decreases. The change in the density of the wastewater as a result of temperature may have a more noticeable effect.

In an activated sludge clarifier, the mixing action resulting from the formation of a density current near the clarifier bottom and a reverse current near the surface is so great that there is no stratification resulting from temperature in these clarifiers. In clarifiers following fixed film reactors, however, there can be a marked layering of the clarifier contents. Because of the minimal density of a fixed film reactor's effluent, there is no discernible density current formed. There can be a stratification of the clarifier contents because of a warmer influent wastewater overlaying the cooler contents of the clarifier.

To determine if there is any temperature stratification, a conventional electronic thermometer can be used to measure the vertical profiles at various locations in the clarifier.

**SALINITY.** Salinity may be a concern in industrial wastewater clarifiers or in clarifiers receiving significant amounts of saltwater intrusion. To determine if there is any stratification resulting from salinity, a total dissolved solids probe could be used to measure salinity variations in the vertical profiles at various locations in the clarifier.

**STATE POINT ANALYSIS.** To estimate the theoretical capacity of a clarifier to separate and thicken the activated sludge, a "state point" analysis can be performed. This is a type of graphical analysis that uses the settling rates of various dilutions and concentrations of the mixed liquor to predict the solids loading at which the clarifier will fail. The term "fail", as used in this analysis, refers to the solids loading rate at which the sludge blanket will continue to increase in depth. For more detail on this analysis, refer to the Testing for Activated Sludge Settling Properties section.

**Table 7.2** Results of a comparison of two additional rectangular clarifiers. The tabular entries are the suspended solids concentrations at each depth in milligrams per liter.

| Clarifier #3 at 5 667 $m^3/d$ (1.5 mgd) | | | | |
|---|---|---|---|---|
| **Location and depth** | **1.8 m (6 ft)** | **7.0 m (23 ft)** | **12 m (40 ft)** | **Sum (units)** |
| −0.3 m (−1 ft) | 1.45 | 0 | 0 | |
| −0.6 m (−2 ft) | 1.99 | 0 | 0 | |
| −0.9 m (−3 ft) | 0.4 | 0 | 0 | |
| −1.2 m (−4 ft) | 0 | 0 | 0 | |
| −1.5 m (−5 ft) | 0 | 0 | 0 | |
| −1.8 m (−6 ft) | 0 | 0 | 0 | |
| −2.1 m (−7 ft) | 0 | 0 | 0 | |
| −2.4 m (−8 ft) | 0.11 | 0 | 0 | |
| −2.7 m (−9 ft) | 0 | 0 | 0 | |
| −3.0 m (−10 ft) | 0.62 | 0.33 | 0.26 | |
| −3.4 m (−11 ft) | 0.96 | 0.39 | 4.21 | |
| −3.7 m (−12 ft) | 1.35 | 10 | 10 | |
| Totals | 6.9 | 10.7 | 14.5 | 32.1 |

| Clarifier #4 at 5 667 $m^3/d$ (1.5 mgd) | | | | |
|---|---|---|---|---|
| **Location and depth** | **1.8 m (6 ft)** | **7.0 m (23 ft)** | **12 m (40 ft)** | **Sum (units)** |
| −0.3 m (−1 ft) | 0 | 0 | 0 | |
| −0.6 m (−2 ft) | 0 | 0 | 0 | |
| −0.9 m (−3 ft) | 0 | 0 | 0 | |
| −1.2 m (−4 ft) | 0 | 0 | 0 | |
| −1.5 m (−5 ft) | 0 | 0 | 0 | |
| −1.8 m (−6 ft) | 0 | 0 | 0 | |
| −2.1 m (−7 ft) | 0 | 0 | 0.27 | |
| −2.4 m (−8 ft) | 0 | 1.18 | 0.67 | |
| −2.7 m (−9 ft) | 1.15 | 1.28 | 2.04 | |
| −3.0 m (−10 ft) | 2.02 | 2.90 | 3.28 | |
| −3.4 m (−11 ft) | 5.69 | 7.15 | 9.10 | |
| −3.7 m (−12 ft) | 10 | 10 | 10 | |
| Totals | 18.9 | 11.5 | 25.4 | 65.7 |

## RELEVANCE TO DESIGN

If the project involves the modification of an existing clarifier, a field evaluation is invaluable in identifying the actual performance characteristics. Considering the work of Anderson, his field evaluations were conducted as a means to discover how a particular existing design performed. Supported by this field data, the owner was then able to proceed with the construction of more of the same type clarifier with more confidence. In other cases, where existing clarifiers have been evaluated, designers have been able to either incorporate unique design features that responded to their particular circumstances or use more common modifications with more confidence.

In many cases, the information gained from field testing has led an owner or designer to modify certain clarifier details, such as an inlet baffle or an effluent weir. In other cases, the evaluation may identify the need for other types of revisions, such as better flow distribution or return sludge flow control.

One of the most beneficial aspects of identifying the hydraulic characteristics of a clarifier is gaining the knowledge of how the currents are formed and travel in a clarifier. With this information, a designer is able to address the particular problems identified with specific design features.

## CASE STUDIES

**CIRCULAR CLARIFIERS.** An example of a circular clarifier follows:

- Design flow: 75 700 $m^3$/d (20 mgd);
- Activated sludge process;
- Four 30.5-m (100-ft) diameter × 3.7-m (12-ft) side water depth (SWD) secondary clarifiers; and
- Rapid sludge withdrawal system with four suction tubes per arm.

The primary objective was to compare a clarifier with a modified center well with an unmodified clarifier.

The major field tests used were the following:

- Flow curves at normal flow and high flow,
- Dye tracer tests at both flow rates,
- Drogue current velocity tests, and
- Vertical solids profiles of the sludge blankets

Before each test, the flowrates were checked by job-built flow measurement weirs inserted in the effluent launders. These rates were maintained the same to each clarifier throughout the test period.

An example of the test results is this flow curve comparison shown in Figure 7.12. It can be seen that clarifier #5, which has the modified center well, has much better hydraulic characteristics than does clarifier #6.

Other elements of the field data, such as the VSPs and the drogue data, showed that the sludge blankets in clarifier #5 were more compact and the currents were also less intense than those in clarifier #6.

**RECTANGULAR CLARIFIERS.** An example of a rectangular clarifier follows:

- Design flow: 22 710 $m^3/d$ (6 mgd);
- Activated sludge process;
- Five 26-m- (85-ft-) long × 4.3-m- (14-ft-) wide × 0.9-m (9-ft) SWD secondary clarifiers; and

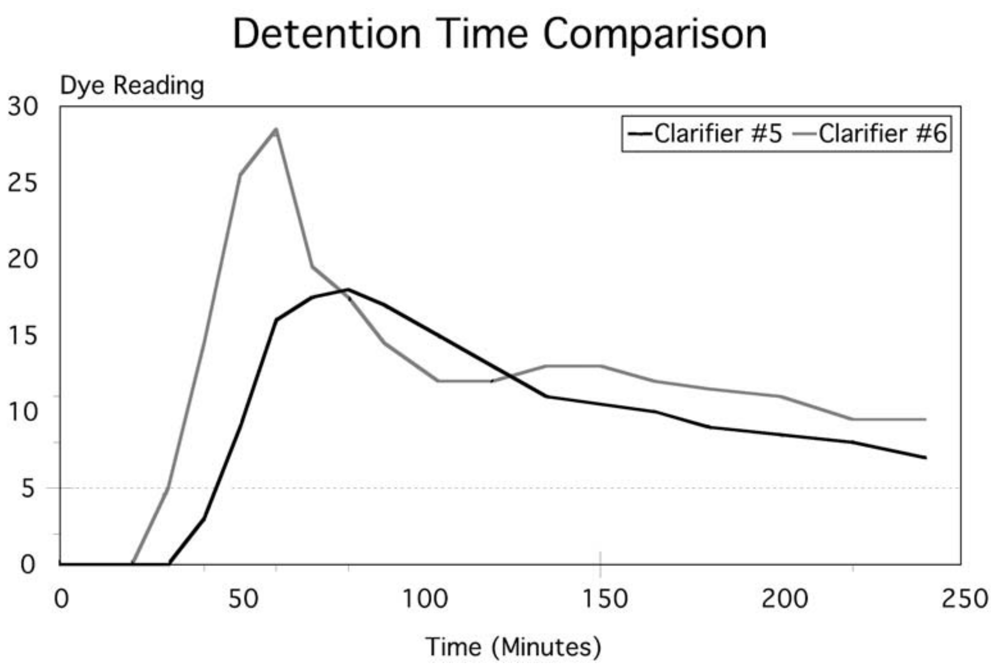


FIGURE 7.12 Example of detention time comparison.

- Conventional chain and flight sludge collectors with the sludge hopper at the influent end.

Three of the clarifiers had been modified with various configurations of baffles. The primary objective was to compare the performance of the new baffle configurations with the performance of the unbaffled clarifiers.

The major field tests used were the following:

- Flow curves at normal and high flow,
- Dye tracer tests at both flowrates,
- Drogue current velocity tests, and
- Vertical solids profiles of the sludge blankets.

The flow curve comparison tests gave a good indication of the effectiveness of the baffle configurations in reducing the currents in the clarifiers. Figure 7.13 depicts the flow curves at normal flow.

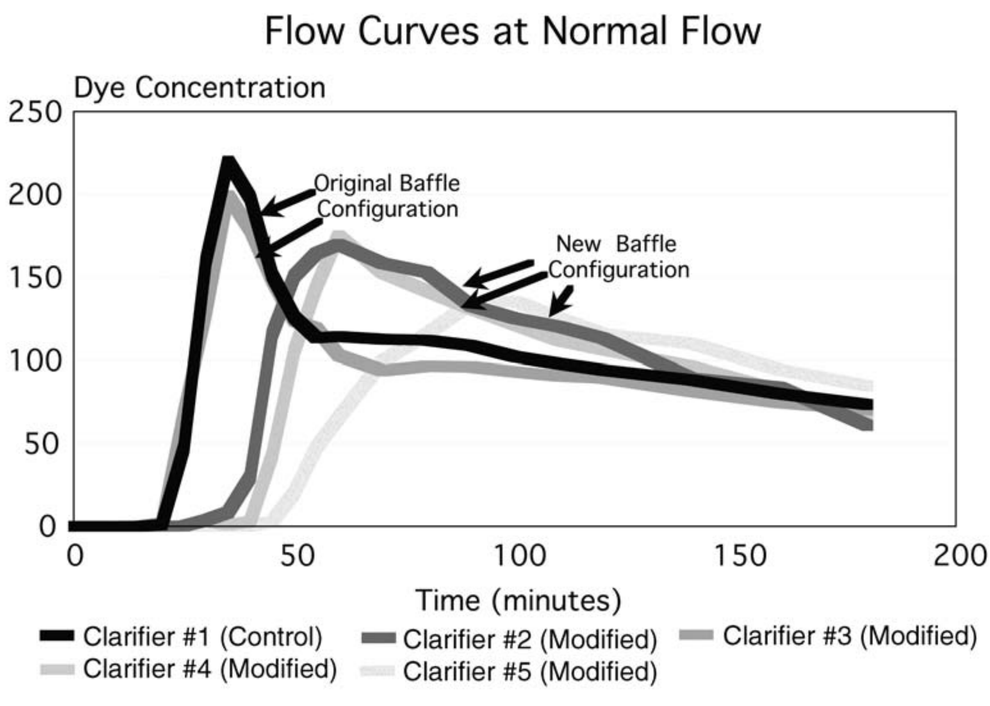


**FIGURE 7.13** Example of flow curves at normal flow.

Other elements of the field evaluation, such as the dye tracer test and the drogue current test, were conducted to identify the reasons for the differences in the flow patterns.

## REFERENCE

Wahlberg, E. J. (2001) *WERF/CRTC Protocols for Evaluating Secondary Clarifier Performance*; Project Number: 00-CTS-1; Water Environment Research Foundation: Alexandria, Virginia.

# Chapter 8

# Circular Clarifiers


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## INTRODUCTION

Circular clarifiers have earned a reputation for being the most trouble-free with respect to the sludge-collection mechanisms. Square, hexagonal, and octagonal tanks are somewhat like circular in the form of the hydraulic flow regimes that are typically established, but have certain differences that limit their popularity. If filets are used in the corners and simple collection mechanisms are used, these alternate shapes have nearly all the advantages of circular tanks. For purposes of this chapter, tanks of these shapes are considered essentially equivalent to circular tanks. Where differences are significant, they are pointed out.

Circular tanks have the disadvantage of taking more footprint for equivalent capacity than rectangular units built with common wall construction. Furthermore, circular tanks require more feed and sludge piping and separate pumping stations to remove sludge. Square, hexagonal, and octagonal units have some common wall construction, but this advantage over circular is offset by a requirement for thicker walls.

The circular shape inherently leads to separate structures for flow splitting ahead of the tanks and for sludge pump stations. For flow splitting, the most common and effective structure involves feeding the structure at a low elevation, causing flow to rise vertically and then dividing by flowing over two or more weirs, each of the latter feeding a circular tank. Another concept is to provide overflow weirs along an aerated channel that has very low horizontal velocities. On some large plants, modulating butterfly gates with computer-controlled operators have been used successfully.

For sludge removal, it is important to have independently measured and controlled withdrawal for each clarifier. In many instances, measuring hydraulic flow is satisfactory; some design engineers include solids concentration measurement that enables mass flux monitoring without sampling. Such instrumentation is generally located in a separate pump station serving circular tanks; whereas, they are commonly located in galleries for rectangular units.

This chapter emphasizes the aspects of design important to clarifier performance. Early chapters should be referred to for theory and for sizing the number and diameter of circular tanks. Practical limits for size are discussed in this chapter.

Circular clarifiers are used extensively in all three levels of wastewater treatment. These include primary, secondary, and tertiary treatment. Primary treatment generally consists of settling raw wastewater, but the addition of inorganic coagulants and polymers to enhance performance is practiced at an increasing number of plants.

Secondary clarifiers serve both fixed- and suspended-growth biological systems. Because of the heavier solids loading of suspended-growth systems, special design features are provided.

For tertiary treatment, some of the technologies initially developed for water treatment are used. Chemicals are commonly added to improve suspended solids removal. These concepts include solids contact clarification, solids recycle systems, and ballasted clarification, such as discussed in Chapter 5. Rather than have a separate, exclusive section on design features of tertiary clarifiers alone, the text of this chapter includes features of tertiary tanks along with features of primary and secondary tanks.

It is not the intent to cover all operations aspects of clarifiers in this chapter. Some comments are made about the differences of operational requirements among alternative designs, but other manuals and parts of this manual are recommended for that purpose.

Proper design, specification, and review of submittals are extremely important for the success of circular clarifiers. There is not a "one-size-fits-all" circular tank design that is recommended. This text points out many of the features that must be carefully combined to obtain best value for each unique site.

The science of circular clarifier design is continuing to evolve. New features for removing sludge, controlling algae, and other goals of treatment are found on almost an annual basis.

This chapter does not present much detail relating to materials selection in clarifier design. For large municipal plants, clarifier walls and floors are made of concrete. For industrial applications, coated steel is used more extensively. Internal mechanisms have most commonly been constructed using coated steel; however, stainless steel and fiberglass have been gaining market share because of their abilities to avoid corrosion.

## DESIGN

Circular clarifiers used in primary, secondary, and tertiary wastewater treatment have tank diameters that range from 3m (10 ft) to greater than 100 m (300 ft). However, for most plants, diameters are kept to less than 50 m (150 ft) to avoid the adverse effects of wind on the surface.

Once the number, shape, and sizing of clarifiers are complete, detailed design becomes the most important step in obtaining sedimentation facilities that are successful in performance, operations, and maintenance. This section covers the design of circular clarifier inlets, outlets, depth, interior baffles, sludge removal systems, skimming systems, algae control facilities, and other ancillary facilities, such as walkways, railings, and lighting.

**INLET PIPE AND PORTS.** Circular clarifiers can be fed by several different inlet configurations. Most plants are fed from the center; however, as shown in Figure 8.1, there are two basic peripheral feed alternatives. Mid-radius feeding devices have been developed and even patented; however, their use is so rare that the subject is not discussed further in this text.

The location of the feed point determines the internal hydraulic regime of the tank. Center feeding causes the flow to move radially outward toward the weir, and, in many tanks, there is a doughnut-shaped roll pattern formed, which results in some surface flow back towards the center. The opposite pattern is formed by

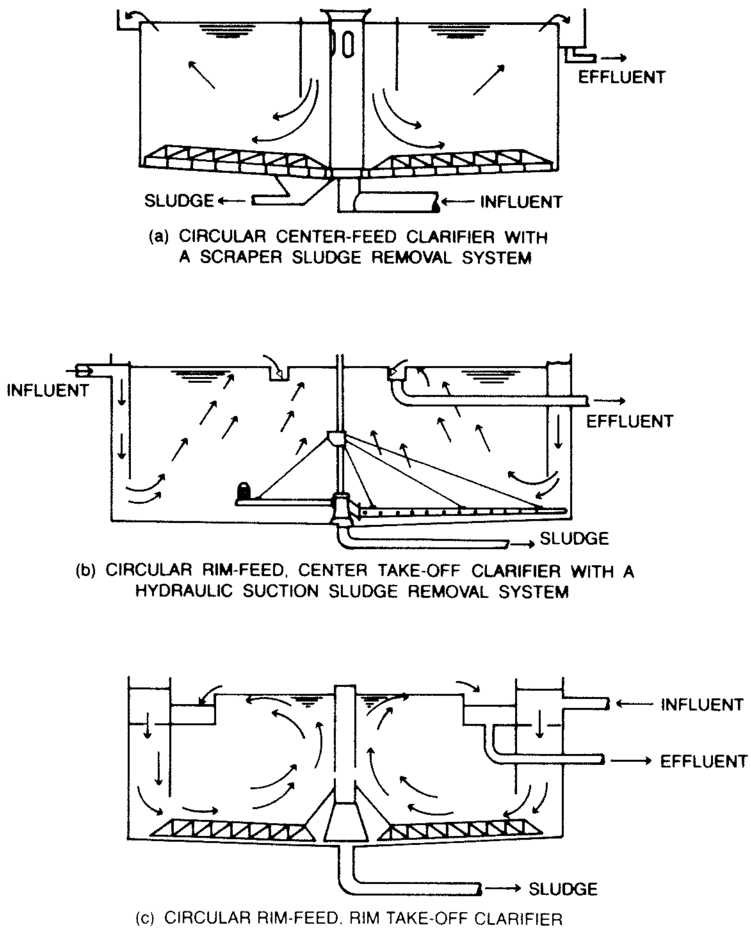


**FIGURE 8.1** Typical circular clarifier configurations and flow patterns.

peripheral feed devices. Additional details are presented in the Inlet Geometry section of this chapter.

**PIPE SIZE AND VELOCITIES.** Most United States clarifiers are equipped with the mechanism drive located at the top of the center column. The center-feed pipe must then serve a dual role of bringing influent into the tank and transmitting rotational torque from the drive into the bottom foundation. Special structural calculations beyond the scope of this text are required to determine the physical requirements of the center column for load transmission.

From a treatment and hydraulic standpoint, the influent pipe should be sized to keep material in suspension, but keep velocities low enough to avoid floc breakup and excessive head loss. Many manufacturers design the influent velocity at peak hour flow and maximum return activated sludge (RAS) flow, with one unit out of service, not to exceed approximately 1.4 m/s (4 ft/sec). Some other designers lower this to approximately one-half of this value to minimize floc breakup. For peripheral feed tanks, the velocity of inflow to the distribution feed trough or skirt should also be kept to less than this value.

For center feed tanks with ports that transmit flow from the feed pipe into the feedwell, port velocities should not exceed feed pipe velocities discussed above.

Most center feed columns have four rectangular opening ports. They are often submerged, although some designs may show the top several centimeters of the ports exposed. Instead of ports, another popular feed pipe opening concept is to connect two segments of pipe with four vertical structural steel channels welded to each pipe exterior.

For peripheral feed tanks, some designs have a raceway with multiple ports at its bottom. Others have an open raceway, and the tangential dispersion of influent is achieved by introducing a directional spiral feed pattern. For those inlets with multiple ports, the port spacing and size is generally performed by equipment manufacturers who have computerized hydraulic models for this purpose. Most design engineers specify, for a given range of flows, that the relative flows leaving the different ports do not vary by more than 5% (or such value) from the total flow divided by the number of ports.

In many center feed inlet designs, the inlet ports discharge freely into the inlet feedwell. In some, however, deflectors are constructed just downstream of each port to break up the jetting velocities into the inlet baffled area. Likewise, for peripheral feed tanks with multiple port bottom openings, a deflector plate is typically located immediately downstream of each port opening. This diffuses inflow at that point and prevents jetting of flow down below the influent skirt and into the settling zone.

As shown in Figure 8.2, center feed tanks can also be fed with horizontal pipes or vertical pipes that discharge freely at their end. Some of these pipes can also be equipped with a bell-mouth outlet that reduces the release velocity into the tank center.

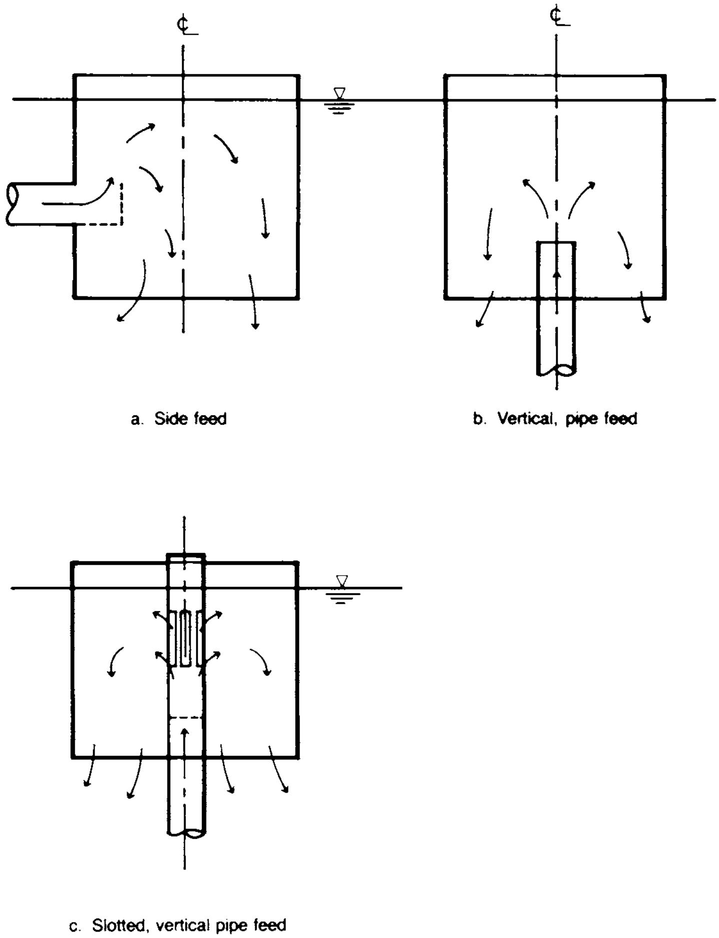


**FIGURE 8.2** Various conventional center feed inlet designs: (a) side feed; (b) vertical, pipe feed; and (c) slotted, vertical pipe feed.

It is important to have a termination baffle or an upturned elbow on a horizontal feed pipe so that it does not release flow with any residual horizontal velocities. Such unbalanced velocity vectors can seriously disturb the internal flow patterns of the clarifier and affect effluent quality.

**INLET GEOMETRY.** ***Center Feed.*** A standard center feed inlet design for a circular tank is shown in Figure 8.3. It remains one of the most common ways of

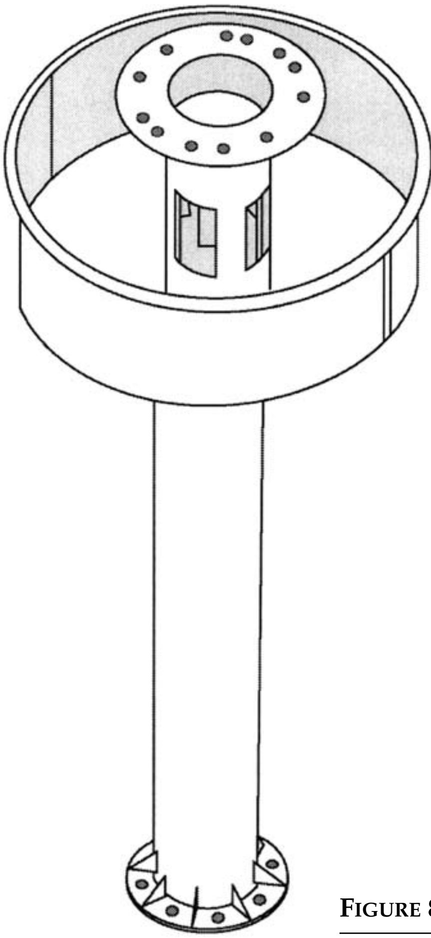

**FIGURE 8.3** Standard center inlet design.

feeding center-feed primary clarifiers. The diameter of the feedwell equals 15 to 20% of the tank diameter for most primary clarifiers and secondary clarifiers following fixed-film reactors. For activated sludge facilities, the feedwell size is often 20 to 25% of the tank diameter. This enlargement accommodates the recycling of return sludge, which can be as high as 100 to 150% of the average plant flowrate. The diameter of the simple feedwell is typically determined by the criteria for downward velocity of flow in it. Some designers and manufacturers advise that the feedwell diameters do not exceed 10 to 15 m (35 to 45 ft), regardless of tank size. Likewise, downward flow velocities leaving the feedwell are often limited to approximately 0.7 m/min (2 or 2.5 ft/min).

The top elevation of the feedwell is generally designed to extend above the water surface at peak hour flow with one unit of service. A few ports are typically cut into the top portion of the baffle to allow scum to move from the feedwell into the tank proper. It is common to place four such openings equidistant around the baffle.

A typical center feedwell extends downward from as little as 30% to as much as 75% of the tank depth. Several manufacturers recommend that submergence be 25 to 50% of the side water depth.

It is also common that the center feedwell bottom edge be located approximately 0.3 m (1 ft) below the bottom of the center feed pipe ports. It must be low enough so that the flow jetting out of the ports does not get below the baffle and out into the settling zone.

One design concept recommends that the cylindrical area below the feedwell be approximately equal to the feedwell cross-sectional area. This prevents a velocity increase as the liquid enters the lower portion of the clarifier. In this case, the opening under the feedwell would be measured as the side water depth minus the feedwell depth. This requirement may conflict with the clarifier feedwell velocity criteria and side water depth criteria discussed above. Therefore, it is often necessary to find a compromise that meets most of the criteria simultaneously.

In some conventional tanks, the feedwell rotates with the sludge scraper mechanism, whereas, in others, it remains stationary. The feedwell can be supported from the bridge or from the sludge collector mechanism. If it is supported by the bridge and does not rotate, care should be taken to avoid aligning the feed pipe ports with the scum port openings.

A typical inlet velocity pattern resulting from the use of the simple center feed inlet is illustrated in Figure 8.4. In activated sludge treatment, incoming mixed liquor has a higher density than the supernatant content in the tank. This leads to the formation of a “waterfall” effect, as illustrated in Figure 8.5. Murphy (1984) observed this effect and found the influent to flow radially outward in a relatively thin sheet across the lower elevations of the tank, just above the interface with the settled sludge. Crosby (1980) also reported that the influent velocity vectors can be distorted by the sludge collector riser pipes if they pass in front of the inlet ports of the feed pipe.

To reduce the cascading effect of the influent flow in activated sludge treatment, a flat circular baffle, similar to that shown in Figure 8.6, has been recommended by McKinney (1977) and implemented in several designs. In the United States, however, this baffle is not often constructed. The baffle is most valuable in tanks with plows

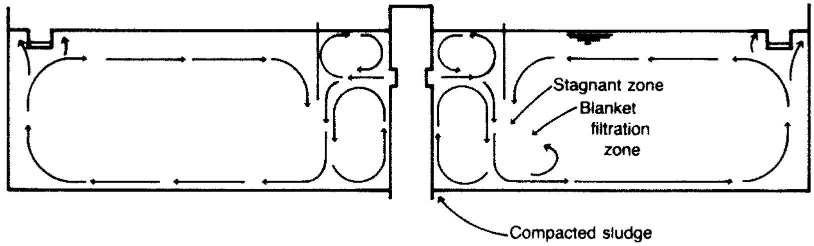


**FIGURE 8.4** Typical velocity pattern of center feed tank.

and central hoppers for sludge removal. It prevents scouring of the sludge hopper and facilitates the plowing of sludge radially inward as the influent flow moves in the opposite direction.

In clarifiers settling raw wastewater or fixed-film secondary process effluent, the sludge layers are much thinner and solids are generally removed by scraper mechanisms. The simple center feedwells discussed above have been giving satisfactory performance over a range of reasonable submergence depths.

For activated sludge clarifiers, however, the bottom elevation of the center feedwell has a significant effect on performance. The relative level of the sludge blanket surface must then be considered in both the design and operation of the clarifier. Sorenson (1979) examined the strategy of maintaining a deep sludge blanket within the clarifier. His data indicated that maintenance of a high blanket

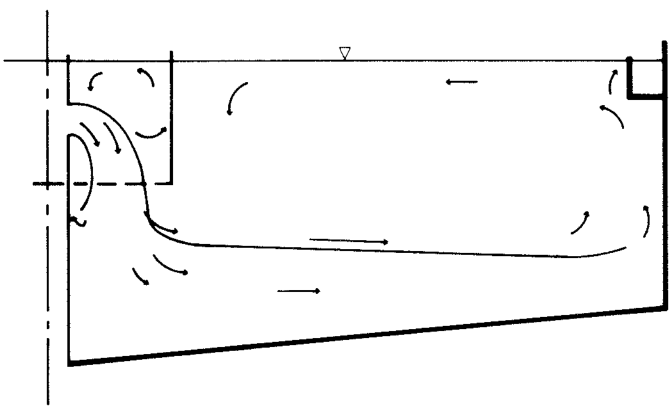

**FIGURE 8.5** Possible solids cascading phenomenon in clarification of activated sludge.

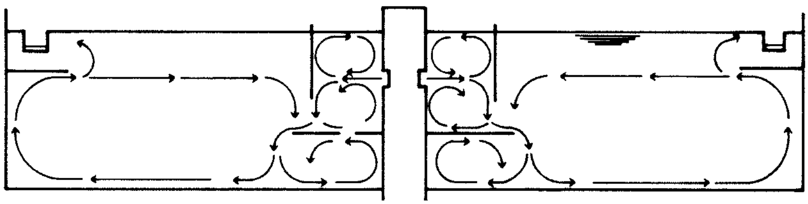

FIGURE 8.6 Circular baffle provided to reduce cascade effect in influent mixed liquor flow.

using automatic control produced a better effluent quality in comparison to a tank with manual control.

Evidence from thickening of some suspensions suggests that the introduction of feed below the blanket level is detrimental to the solids concentration of the underflow. Also, hydraulic surges that enter the final settling tank tend to fluidize a deep sludge blanket and scour solids from it.

Full-scale tank studies by Crosby (1980) showed that better performance would generally be obtained with a center feedwell bottom that is either well above the sludge blanket or somewhat below the top of the sludge blanket. A shallow blanket separated from the well bottom is considered optimal for sludges that settle well, but not always so for sludges that settle poorly. In the latter case, it may be possible for an operator to improve performance by carrying a relatively thick blanket that provides some degree of solids filtration and settling.

Operating with the bottom of the feedwell at nearly the same elevation as the top of the sludge blanket is a condition to be avoided. The sludge blanket provides, in effect, an artificial bottom to the tank. Having it near the bottom of the inlet creates a relatively high radial velocity and flow turbulence across the top of the sludge blanket. This can keep influent solids from settling and even sweep along the solids from the top of the blanket towards the effluent structure.

***Flocculating Center Feed.*** In many plant designs, suspensions arriving at the settling tank are not fully flocculated. Provisions to add polymers are often made. Tank performance may be improved by provision of a separate flocculation zone. Simply increasing the size of the center feedwell is one approach. Some have provided mechanical flocculators within this zone, whereas others have provided an energy-dissipating inlet (EDI) to distribute the flow into the flocculation zone. An example of this concept is shown in Figure 8.7.

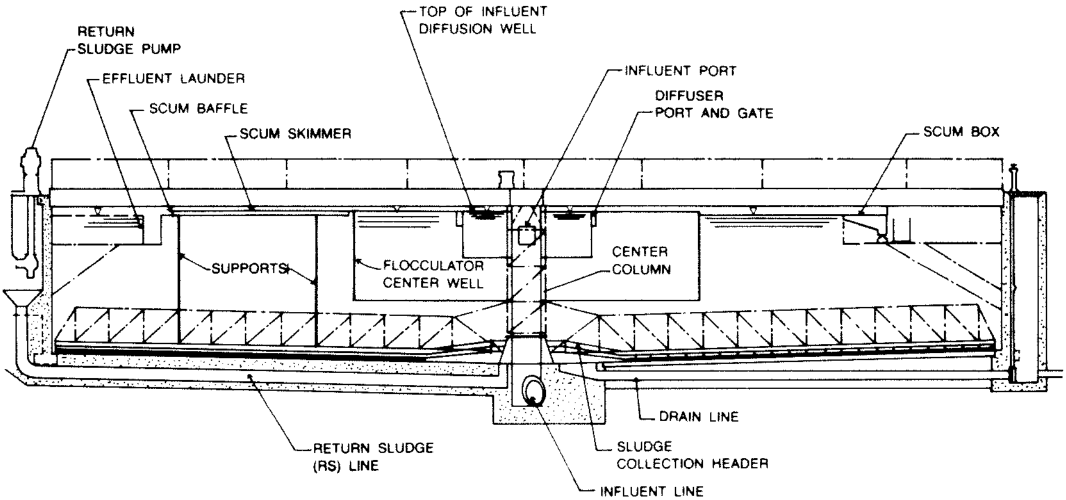


**FIGURE 8.7** Cross-section of secondary clarifier incorporating flocculator center well features.

An enlarged flocculation zone can be beneficial to treating raw wastewater, especially if inorganic and/or polymer coagulants are added. In activated sludge treatment, the mixed liquor suspended solids (MLSS) are often broken up in transfer from the aeration basin to the clarifier. The flocculation zone allows for these broken particles to be recombined into larger, more settleable floc. The amount of tank surface area lost for flocculation is considered minor compared to the improved settleability of the suspension, in most cases.

The sizing of the flocculation centerwell has been the subject of research by Wahlberg et al. (1994a), Parker et al. (1971), and others. It is been shown that a detention time of approximately 20 minutes achieves well over 90% of the obtainable degree of floc formation. Therefore, a rule of thumb has been to size the flocculation well to obtain 20 minutes of residence time at average dry weather flow with an additional allowance of 50% for RAS flow.

This criterion has also been compared to a more simplistic approach of setting it equal to 30 to 35% of the clarifier diameter. Wahlberg et al. (1994b), presented dye study information that showed the size of flocculation chamber can be made too large, resulting in short circuiting of influent within it.

The depth of projection into the clarifier by the flocculation well is also an important design criterion. Many design engineers have arbitrarily set this at a value equal to approximately one-half of the tank depth at the location of the baffle. With a sloped floor, this would be a little deeper than one-half of the side water depth. In more recent designs supported by results from computational fluid dynamic modeling, shallower flocculation baffle penetrations have been used. Some of these are less than one-half of the side water depth. If the baffle is too shallow, however, it is possible that some residual jets from the EDI would fall below the bottom of the flocculation baffle and enter the quiescence zone of settling. This can lead to excessive solids carry-over to the effluent.

In some early designs, several slow-moving, pitch blade vertical turbines were provided to obtain floc formation. Parallel operation of such systems has shown that equivalent results can be obtained with the mixers on or off. In recent years, EDIs have been used to obtain adequate mixing within the flocculation zone, and mechanical mixers are rarely, if ever, used.

Early designs of the EDI were similar to the one shown in Figure 8.8, using the simple hinged gate alternate. It was the designer's intent to allow operators to set the adjustment chain to increase or decrease the velocity entering the flocculation zone. Tightening down the chain would increase head loss, inlet velocities, and stirring. In practice, many operators simply set the hinged gate at one location (for example 1/2 or 2/3 open) and did not make further adjustments.

The diameter of the EDI is often set at approximately 10 to 13% of the tank diameter. A detention time of 8 to 10 seconds is used by some design engineers. Making the EDI too large subtracts from the volume of the flocculation zone and increases downward velocities through it.

The use of EDIs with tangential release of flow has been supported by data such as that shown in Figure 8.9. Stirring was shown to be the best method of forming floc and delivering low effluent turbidities. In view of the fact that many operators did not adjust the hinged gates, some designers have tried to improve performance by replacing the hinged gates with curved chutes, such as that shown in Figure 8.8. Indeed, many tanks were designed and constructed with this feature. It commonly appeared in major equipment supplier brochures and catalogs.

In some recent comparative evaluations, it was discovered that the curved chutes resulted in excessive jetting into the flocculation zone. Studies by Esler (1998), Haug et al. (1999), and others at Hyperion (City of Los Angeles) demonstrated that the provision of an EDI with such chutes actually performed worse than adjacent tanks with no EDI at all. A similar side-by-side comparison for a trickling filter final clarifier was performed at Central Weber, Utah. An EDI with curved chutes and its associated flocculation baffle did not perform as well as an old large simple inlet well with a bottom

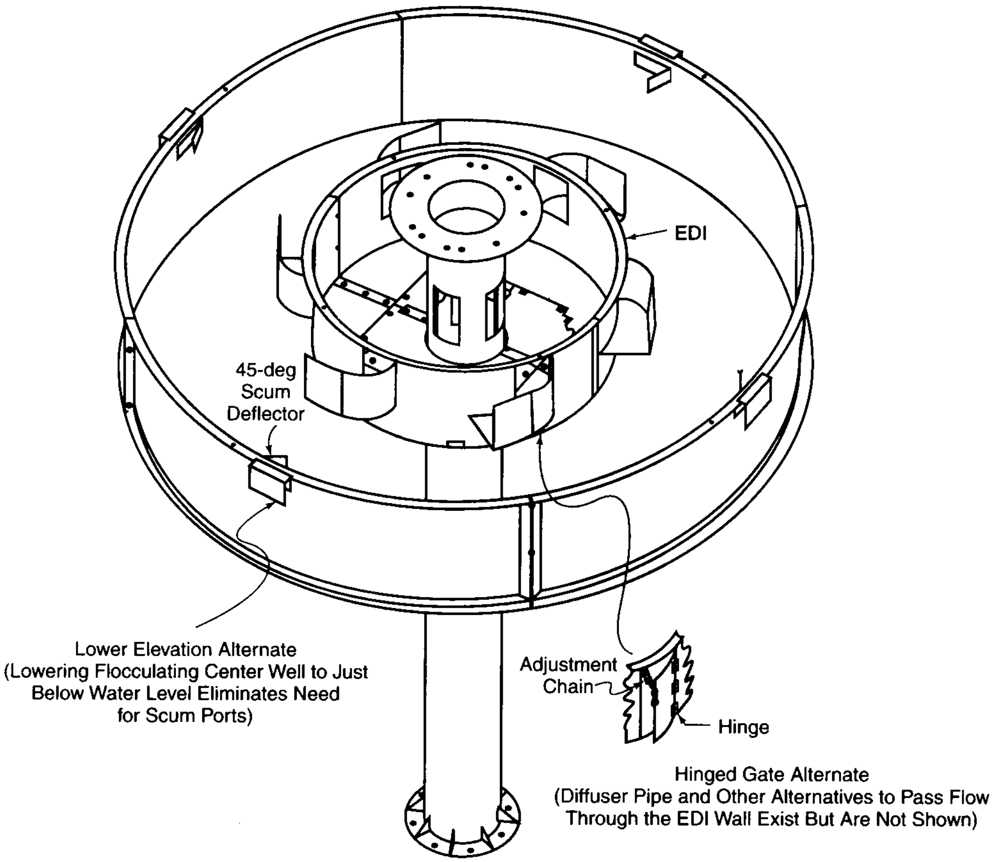


**FIGURE 8.8** Center-column EDI and flocculation baffle.

and a number of diffusers containing a lattice structure around its lower perimeter (Tekippe, 2002).

These findings led to modifications of the curved chutes. The result was the development of the double-gated EDI shown in Figure 8.10. For this design, flow leaving the EDI goes through eight ports, which are immediately followed by a baffle equipped with an adjustable gate at each side. A bottom is also provided between this vertical baffle and the bottom of the feed ports. Thus, flow leaving each port is divided and deflected two ways in the direction of its neighboring ports. This creates impingement of flow at eight locations. Because double gates are provided, one can

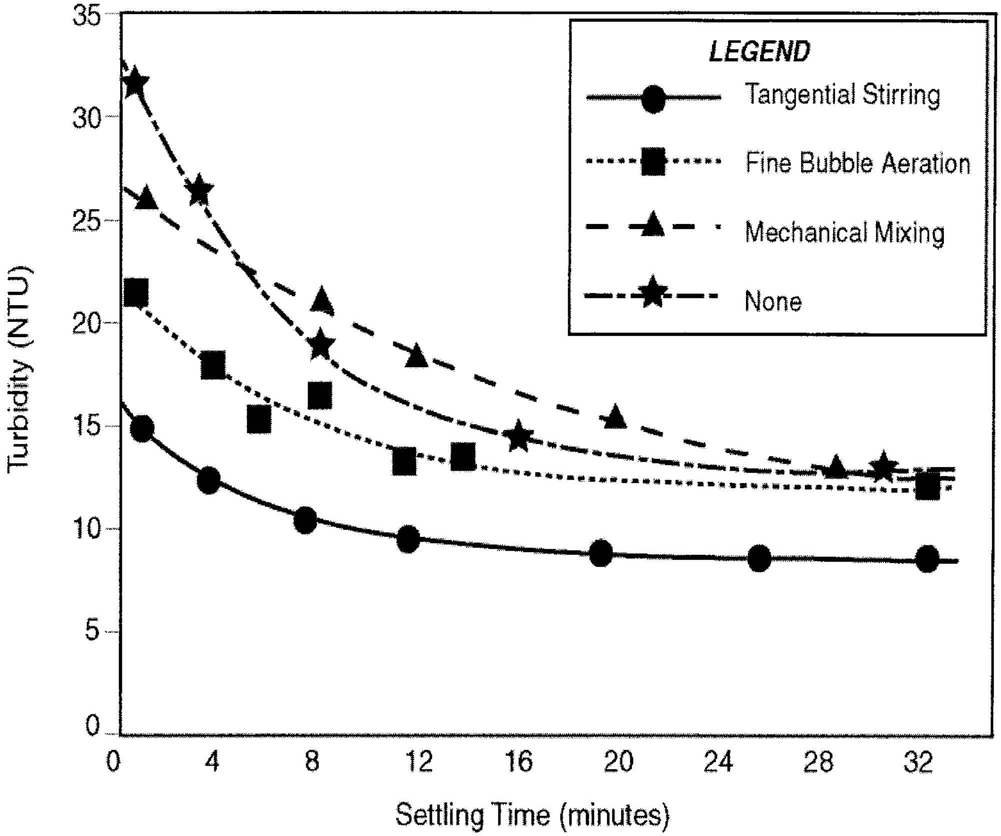


**FIGURE 8.9** Reported results of different flocculation methods.

be opened more than the other to create adjustable degrees of rotation within the flocculation baffle.

This arrangement performed better than the curved chutes at Central Weber and was used to retrofit all four of its trickling filter secondary clarifiers. Diagnostic tests (WERF, 2001) were provided at that plant, and some of the results, shown in Figure 8.11, quantify the improvements gained. The values shown in this figure represent influent suspended solids (ISS), dispersed suspended solids (DSS), flocculated suspended solids (FSS), and effluent suspended solids (ESS).

The side-by-side full-scale studies (Haug et al., 1999) conducted at the Hyperion Wastewater Treatment Plant, serving the City of Los Angeles, led to an innovative

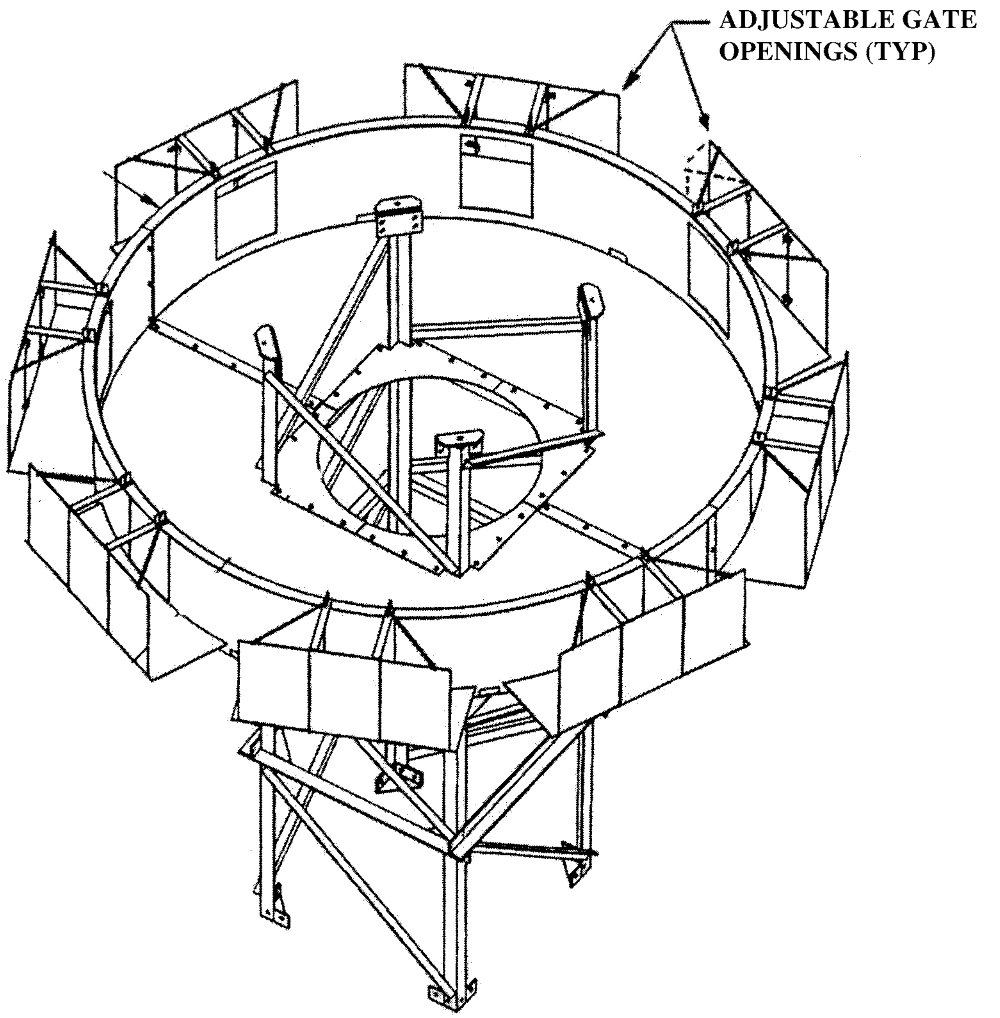


**Figure 8.10** Double-gated EDI used successfully at Central Weber, Utah.

design involving multiple diffusers (similar to those used for many years in rectangular clarifiers at many plants) located around the perimeter at the bottom of the EDI (Figure 8.12). In this arrangement, EDI effluent was conducted downward through eight 0.6-m (24-in.) openings, that, in turn, had 32 small 0.35-m (14-in.) diameter diffuser pipes that were paired off to impinge against each other. Small openings at the

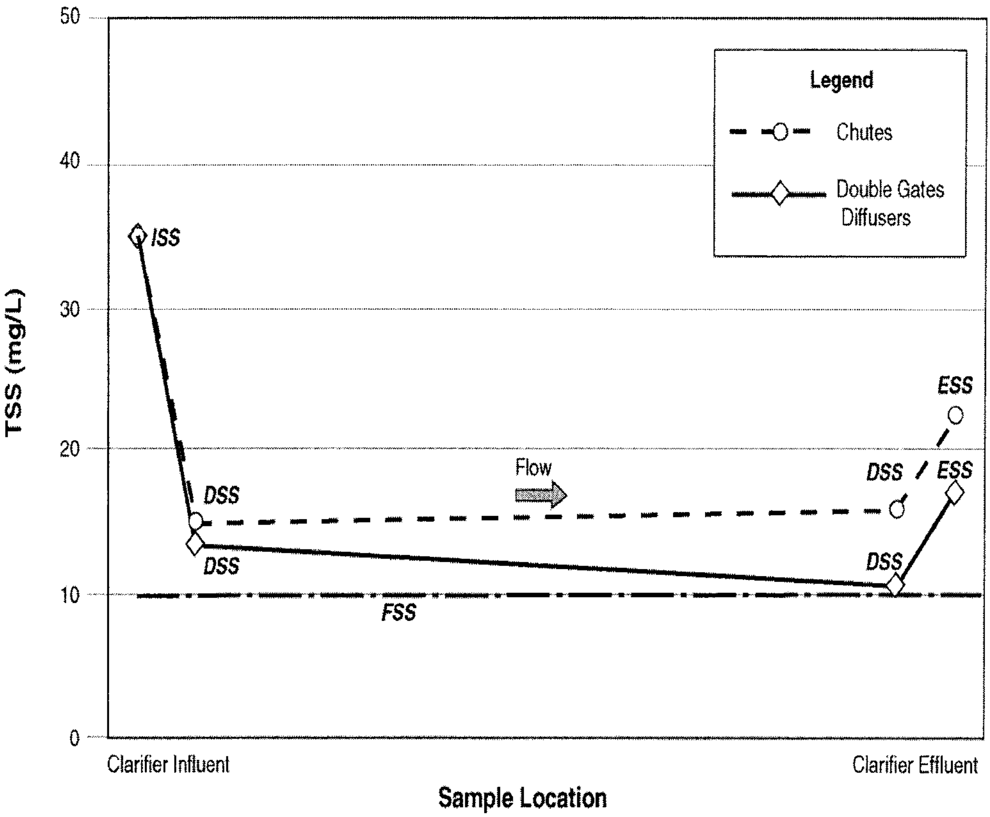


**FIGURE 8.11** Diagnostic test results of different EDI designs at Central Weber, Utah (overflow rate [OFR] = 1.4 m/h [825 gpd/sq ft]) (ISS = influent suspended solids; DSS = dispersed suspended solids; FSS = flocculated suspended solids; and ESS = effluent suspended solids).

surface for passage of scum are provided. This design was found to be superior to any other tested at that plant and was used to retrofit all existing 36 clarifiers at this new facility. Additional details regarding studies leading to this design are presented later in this chapter under the heading Case Studies.

Another innovative EDI design that has just been marketed during the last few years is illustrated in Figure 8.13. In this arrangement, flow enters through four ports from the feed pipe. Opposite each port is a pair of vertical baffles that form a corner. An opening is left midway between these four corners, and flow from adjacent corners impinges as it goes through the openings. Upon leaving this opening, it is split

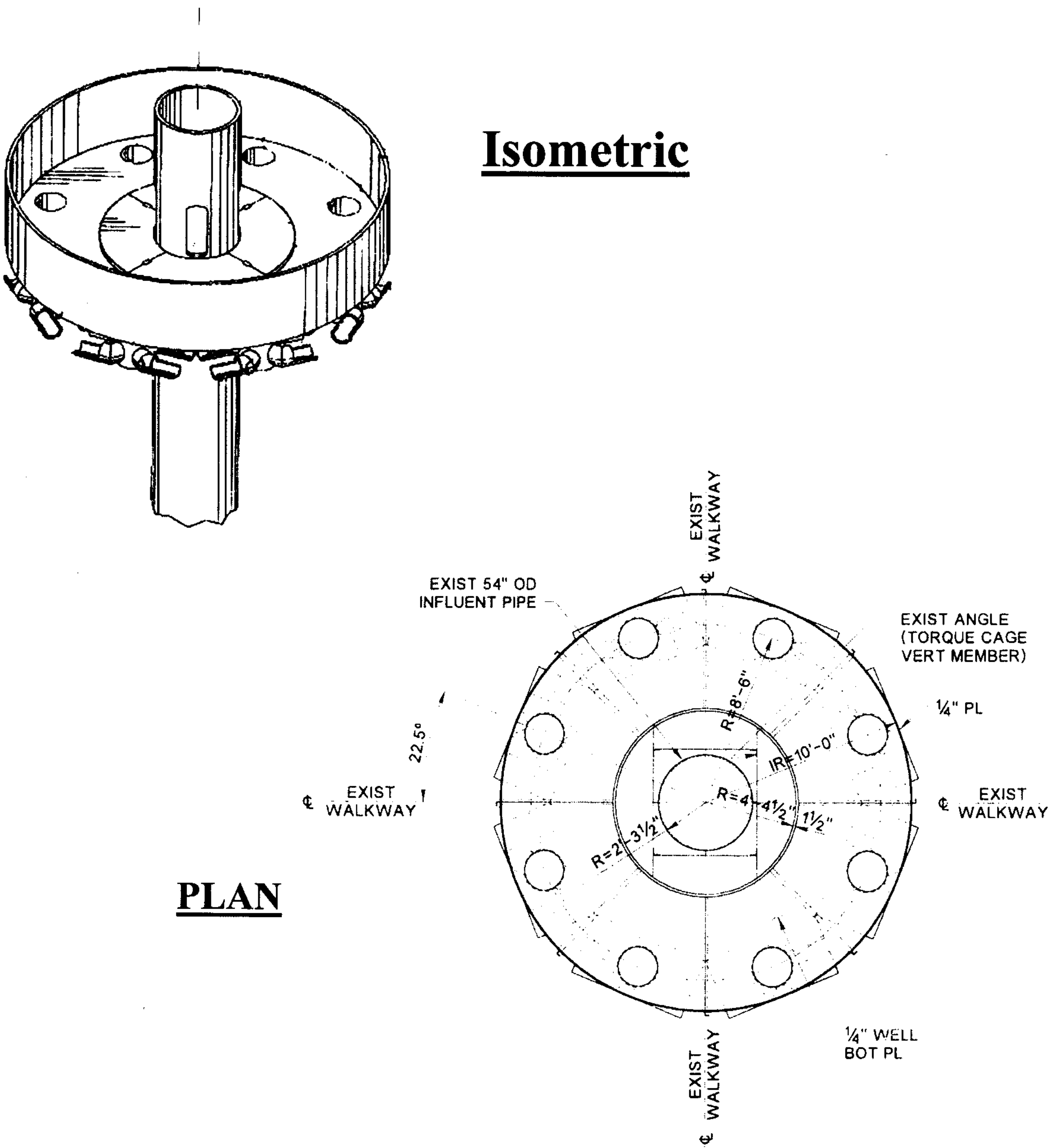


**Figure 8.12** Los Angeles, California, EDI patent drawing and plan view.

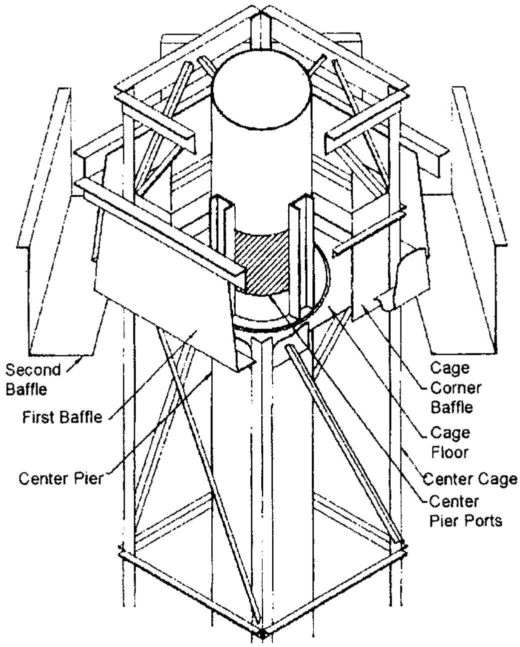


**FIGURE 8.13** Flocculating energy dissipating feedwell (FEDWA).

at 90% and again forced to impinge on flow from adjacent openings. This process is repeated one more time before the mixed liquor is discharged into the flocculation zone. Developers of the FEDWA inlet report good results; however, full-scale, side-by-side tests have not been conducted next to the EDIs with curved chutes, double gates, or Los Angeles (LA)-type diffusers. Nevertheless, its developers are confident of success. Consideration has also been given to elimination of the flocculation baffle with this EDI.

To prevent odors and unsightliness (Figure 8.14), it is important to move floatables out of the flocculation zone. In early years of design, the top elevation of the flocculation baffle was set to project above the water surface at all flowrates. This design resulted in the confinement of foam and other floatables, even though scum ports were provided. At other sites, the top elevation was lowered to equal that of the bottom of the v-notch effluent weirs. This allowed floatables to pass over the top of the flocculation baffle, but still directed most of the flow downward on the inside. However, at high flows, supernatant would actually flow into the flocculation zone

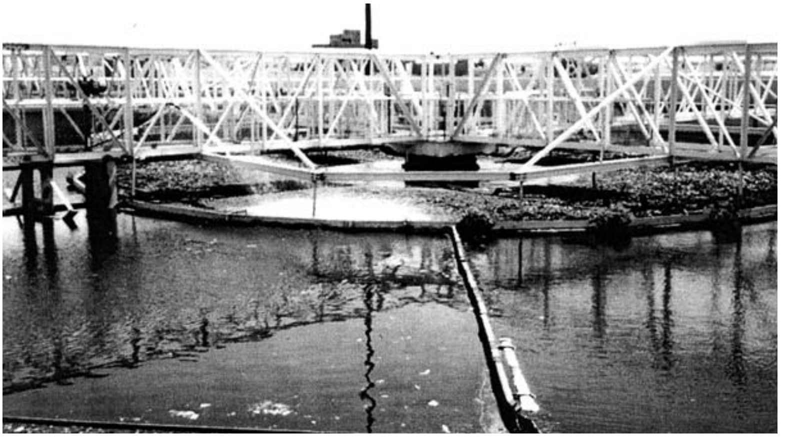

**FIGURE 8.14** A flocculation baffle that traps floatables creates odors.

over the baffle. This, of course, would dilute the contents of the flocculation zone and shorten the detention time of the incoming flow.

To avoid this problem, some have designed the flocculation baffle to be adjustable upward. This allows an operator to raise its level so that it typically projects above the water surface but, at high flows, could be topped to flush the floatables out into the tank proper. In some designs, it was found to be most cost-effective to mount the flocculation baffle itself in a rigid position and bolt an adjustable plate at the top. Careful adjustment of this plate would allow the flocculation baffle to overflow only at the desirable peak-flow periods.

***Center Feed Bottom Release Clarifiers.*** There have been several designs of center-feed clarifiers that release flow into a zone near the bottom of a tank. In some designs, a baffle with vertical slots has been used. In another, rotating arms with several portal openings each have been used to distribute the incoming flow just above the sludge zone. However, these designs have been rarely, if ever, used in the United States, and are therefore not discussed in further detail in this text.

***Tertiary Treatment Clarifier Inlets.*** Designs such as those shown in Figures 8.15 and 8.16 are used for tertiary treatment and water treatment where chemical coagulants are added. They feature mechanical mixing of the contents in the flocculation chamber. The configuration in Figure 8.15b provides for high levels of mixing energy near the tank bottom to prevent any settling from occurring in the central flocculating

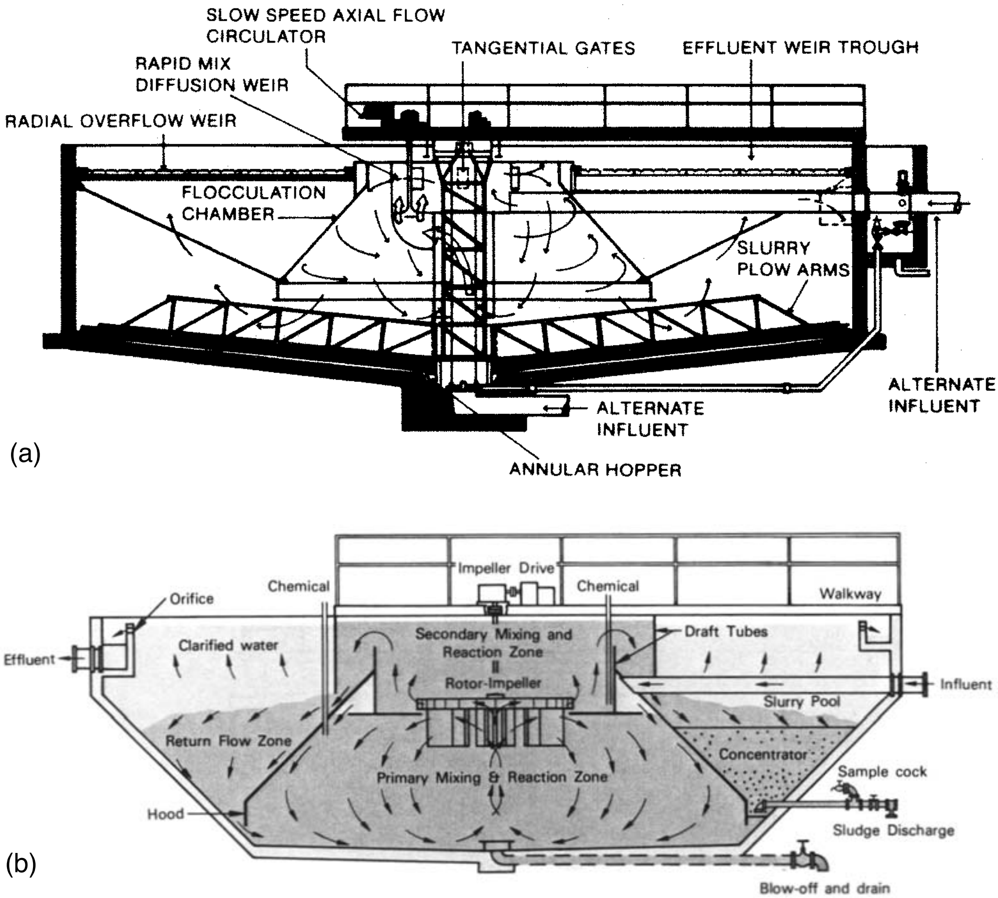


**FIGURE 8.15** Solids contact clarifier design features: (a) mechanically mixed flocculating center feedwell in circular tank with vertical walls, and (b) conical-shaped walls and baffles with two-stage mixing and flocculation.

zone. It is therefore necessary to relocate the hopper to approximately mid-radius of the tank. In some other designs, the central area of solids would be plowed outward to this hopper, whereas the outer portions of the tank would be plowed inward. Vacuum collectors could also be used for this design.

The conical-shaped flocculation baffle creates additional stability for the blanket in the tank because it offers a decreasing vertical velocity as the flow ascends and approaches the launder area. The process capitalizes on using recirculated sludge to

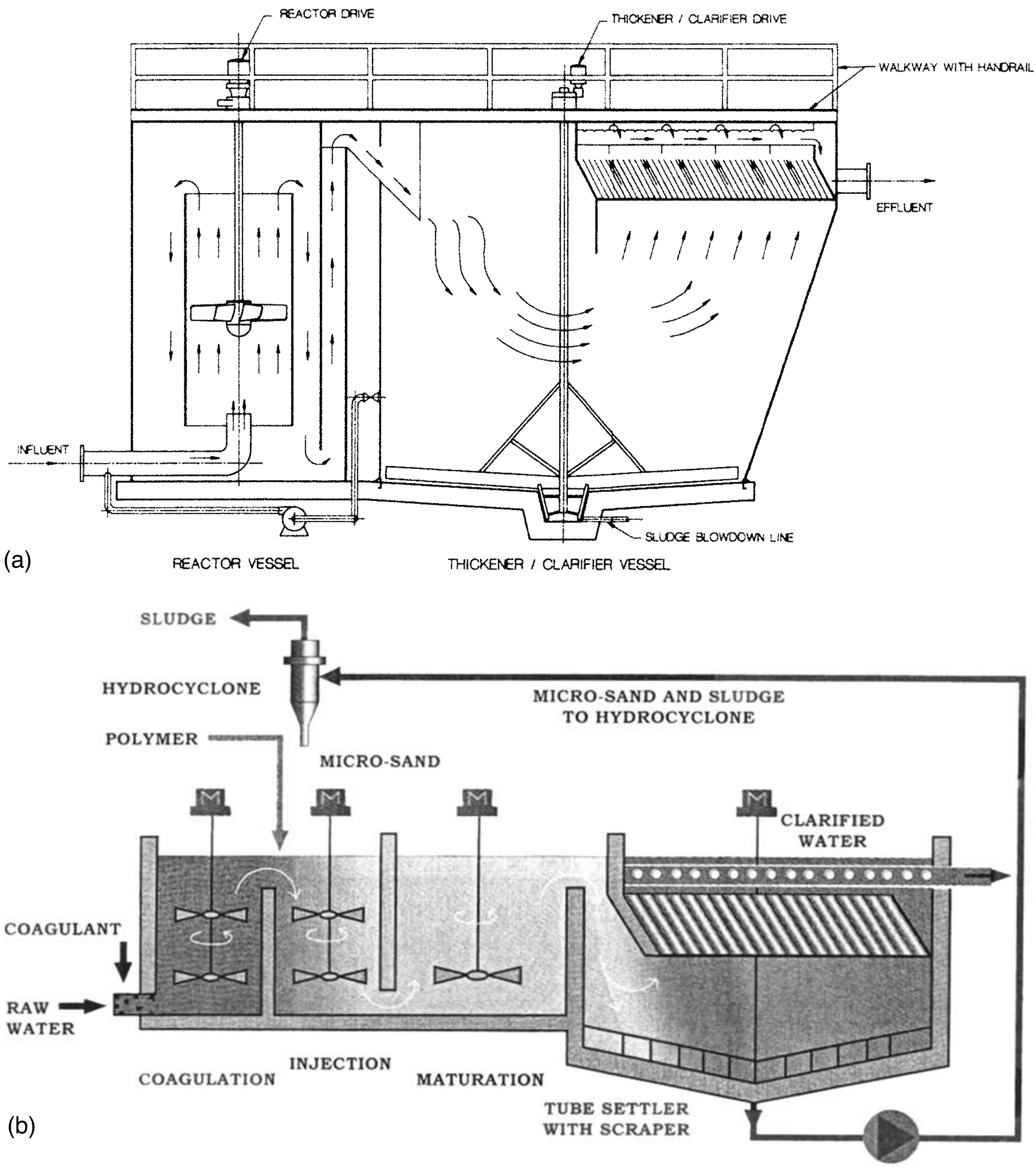


**FIGURE 8.16** High-performance clarifiers with sludge recirculation ballasting: (a) high-rate clarification with sludge recycle, and (b) ballasted clarifier design.

serve as a nucleus on which incoming solids may attach and form a heavier and stronger flocculated particle. This is advantageous in treating dilute quantities of solids when increasing the number of collisions is needed for floc growth.

In solids-contact-type clarifiers, where there is appreciable flow upward through the blanket of settling solids, the stability of the blanket is uncertain if the flowrates fluctuate widely on a diurnal basis.

Providing the appropriate level of mechanical mixing is important. Variable speed mixers are common. They should provide values of energy gradient ($G$), as defined by Camp and Stein (1943), in the range of 10 to 50 $s^{-1}$. Energy gradient is defined as follows:

$$G = \left(\frac{w}{\mu}\right)^{1/2} \tag{8.1}$$

Where

$G$ = root mean square velocity gradient ($T^{-1}$),
$\mu$ = absolute viscosity of the fluid ($FT/L^2$), and
$w$ = rate of power dissipation per unit volume ($F/TL^2$).

For mechanical mixing, the power can be determined from the brake horsepower of the mixer. If water is introduced at high velocity to induce mixing and floc formation, the dissipation can be calculated as follows:

$$w = \frac{Qh_f}{V} \tag{8.2}$$

Where

$w$ = dissipation number function ($F/TL^2$),
$Q$ = discharge ($L^3/T$),
$h_f$ = head loss by friction taken as the velocity head at the inlet ($F/L^2$), and
$V$ = volume of liquid in the basin ($L^3$).

The designs shown in Figures 8.16a and b provide for sludge recycle and ballast addition, respectively. These added features allow considerably higher hydraulic loading and excellent suspended solids removal, but these advantages are partially, at least, offset by higher costs for polymers, ballast microsand, and recirculation energy. They use tube or plate settlers separated from the inlet by an underflow baffle. A relatively low level of inlet geometry sophistication has been found necessary for these designs.

***Peripheral Feed.*** In the 1960s, the concept of spreading inlet energy over a large fraction of the tank volume led to the development of peripheral feed circular clarifiers. As shown in Figures 8.17 and 8.18 and discussed above, the influent is distributed around the perimeter by use of a channel with bottom ports or by means of creating a spiral roll pattern. Most peripheral feed tanks are used for activated sludge secondary settling. However, the spiral flow design option, without inlet ports, has also been used for primary clarification.

Several model and full-scale dye tests have been conducted on peripheral feed clarifiers (Dague, 1960). These have indicated that peripheral feed tanks have a higher hydraulic efficiency than center feed models used in those tests. Specifically, full-scale activated sludge tests conducted at Sioux Falls, South Dakota, showed that, in addition to better hydraulic efficiency, peripheral feed tanks also achieved higher suspended solids removal than the existing center feed design. However, the latter did not use the flocculation centerwell concept developed in more recent years.

In some designs, a head loss across the orifices of approximately 25 mm (1 in.) at average flow was used to obtain reasonably uniform distribution of flow around the perimeter of the tank. For plants with large peaking factors, some maldistribution of flow and solids occurred. Design criteria were changed to provide more head loss (approximately 60 mm, or 2.5 in.) for better distribution at average flow. Peaks of more than 3 to 1 accommodated the higher loss. At low flows, head losses across the orifices can be very low and do not achieve a quality of distribution. However, under these conditions, overflow rates are low and clarifier performance is satisfactory. Minimum flow distribution therefore is not generally considered to be a limiting design criterion. For plants with extreme peaking factors, a special overflow provision in the battle wall or tank wall can be added (Figure 8.17b).

For these inlet designs, the feed channel/zone is baffled off from the body of the settling liquid. As such, floatables can accumulate on the inlet zone surface and create odors and objectionable aesthetics if not removed. Provisions for this are discussed below in the section on skimming systems.

A third type of peripheral feed tank was developed but has received very little use. This consists of transferring flow from a peripheral feed channel down into the lower levels of the clarifier by use of downcomer pipes. The large secondary clarifiers at Detroit, Michigan, are of this design. The performance was considered inferior to other peripheral feed designs, and the downcomer pipes were modified with different types and levels of diffusers. Because of some of these early problems, this concept of peripheral feed has not caught on, and, as a result, it is not discussed further in this text.

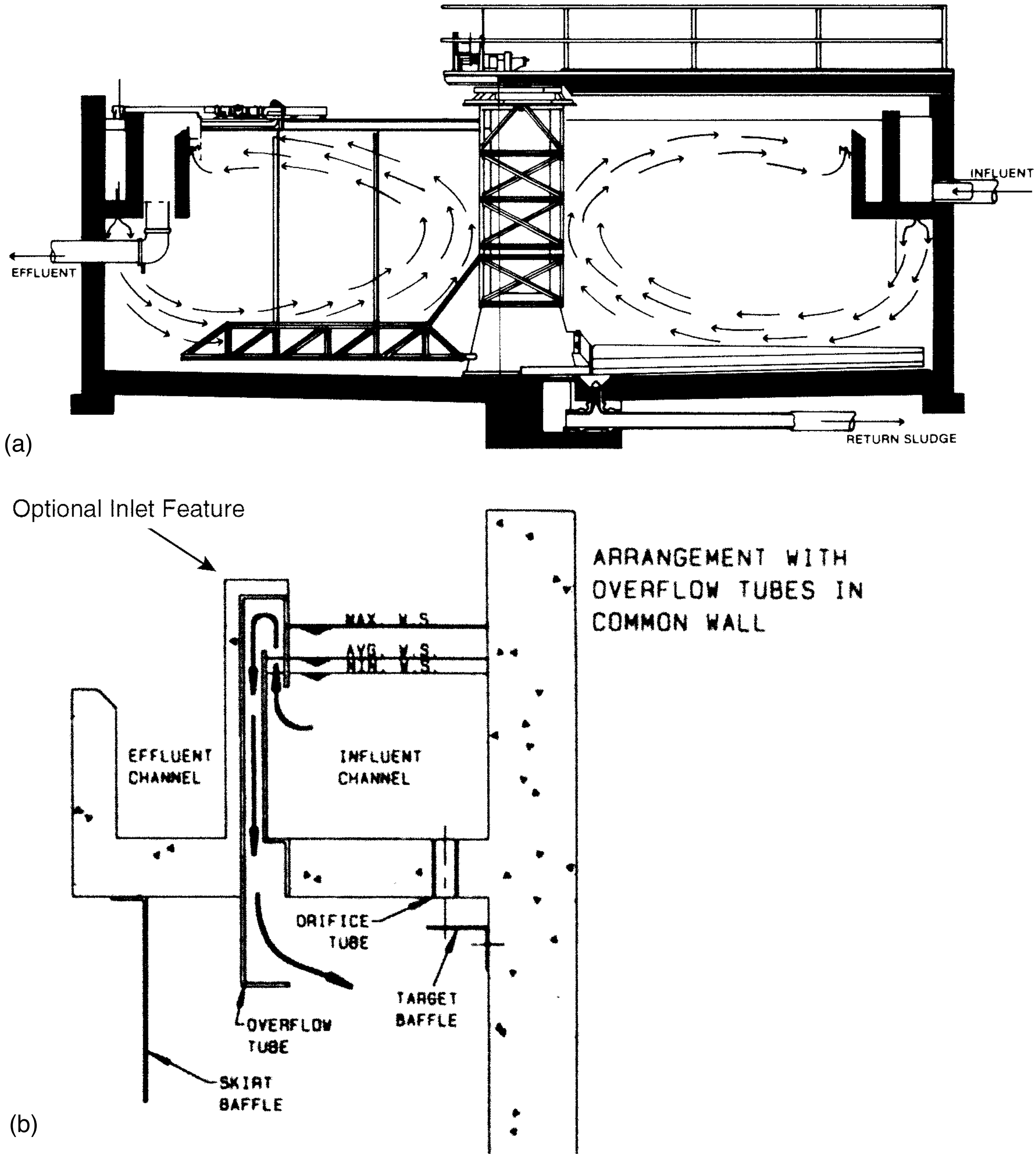


**FIGURE 8.17** Peripheral feed clarifier flow pattern.

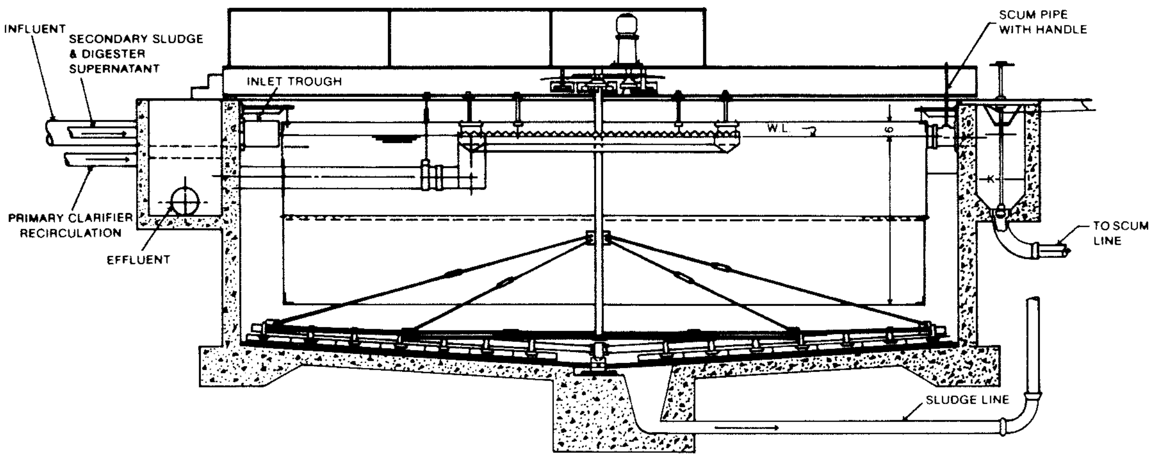


**FIGURE 8.18** Peripheral feed clarifier with spiral roll pattern of flow distribution.

**DIAMETER.** Although area requirements are covered in Chapter 4, circular tanks have some practical sizing limits to be considered in detailed design. Wind on the surface can be sufficient to create surface currents that upset the radial flow balance. Even though tanks up to 100 m (300 ft) in diameter have been built, most engineers counter the wind effect by keeping tank diameters to approximately 50 m (150 ft) or less. Exceptions are taken more often for primary clarifiers, especially those with covers. The performance of primaries is not considered by some designers to be so critical if they are followed by secondary and/or tertiary processes anyway. Even if covers are not used, wind effects are countered somewhat by tanks that have high solid walls that serve to shield the wind, even though they may have been provided for other purposes.

**DEPTH.** ***General.*** Before the early 1980s, circular clarifiers used for primary and secondary settling often had depths of 2.4 to 3.0 m (8 to 10 ft). A shift to deeper circular tanks settling activated sludge was supported by performance data that showed such change to result in lower effluent suspended solids and more resistance to upset from hydraulic peaking. Increased RAS concentrations also occurred. Guidelines, such as those of Table 8.1a, resulted.

In a 1984 survey (Tekippe, 1984), many of the largest environmental engineering consulting firms in the United States were surveyed and reported using design depths of 4 to 5 m (12 to 15 ft) for activated sludge clarifiers. For circular primary

**TABLE 8.1a** Minimum and suggested side water depths for activated sludge clarifiers.

| Tank diameter, m (ft*) | Side water depth, m (ft) Minimum | Suggested |
|---|---|---|
| Up to 12 (40) | 3 (10) | 3.7 (12) |
| 12 – 21 (40 – 70) | 3.3 (11) | 3.7 (12) |
| 21 – 30 (70 – 100) | 3.7 (12) | 4 (13) |
| 30 – 43 (100 – 140) | 4 (13) | 4.3 (14) |
| >43 (140) | 4.3 (14) | 4.6 (15) |

*Note: ft × 0.304 8 = m

clarifiers and those serving fixed-film biological secondary plants, depths of approximately 3 to 4 m (10 to 12 ft) were considered most common.

***Definition of Tank Depth.*** There is a need to be clear about the definition of tank depth in the field of circular tank design. Most professionals think of depths at the side water. Others, however, compare depths at the tank center. In Figure 8.19, a simple graphic is used to illustrate differences between clarifiers with the common center depth but different floor slopes. If these tanks were 50 m (150 ft) in diameter and the side water depths were 5 m (15 ft) for the tank with a bottom slope of 1 on 12, the center depth would be just over 7 m (21 ft). A tank with equal center depth and a flat floor would have more storage volume and, as reported by Parker (WERF, 2001), would be able to keep the blanket further away from the effluent weir at peak loading conditions. The relative cost of the two clarifiers in this figure could be substantial because the tank with the flat bottom would require deep excavation at the walls and greater wall thickness.

***Better Performance with Deeper Tanks.*** During the past two decades or so, several publications have been issued to quantify the effect of deeper tanks. Figures 8.20 and 8.21 compare effluent quality from a variety of final activated sludge clarifiers serving different plants in the United States and Europe. The results suggest that, to obtain effluent suspended solids of 10 mg/L, depths of over 5 m (15 ft) may be required for overflow rates higher than 0.85 m/h (500 gpd/sq ft). The data of Figure 8.22 also show that, for a given overflow rate, one can obtain better effluent quality

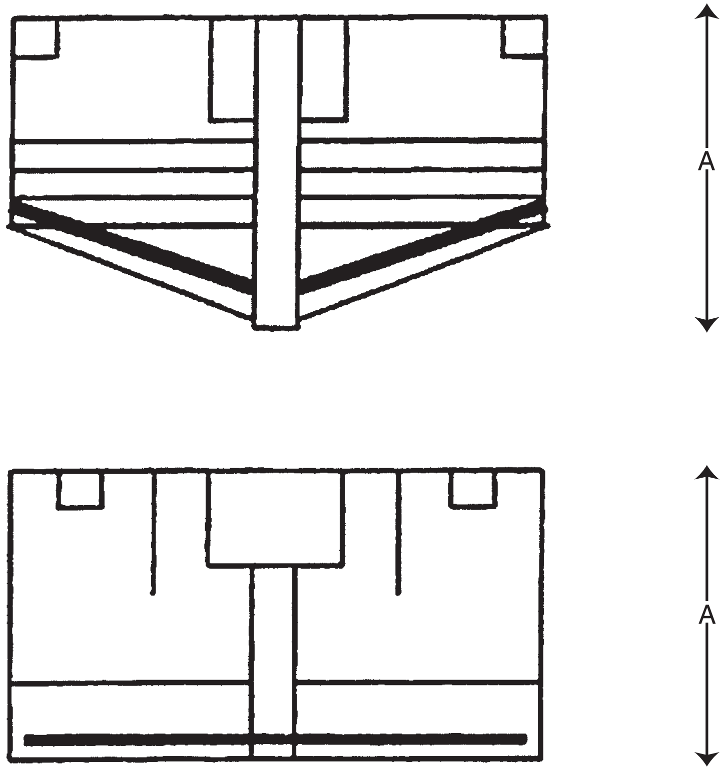


**FIGURE 8.19** Flat bottom tanks with comparable center depth are considered to be deeper and more expensive to construct than those with sloped bottoms, but they offer more storage volume for sludge.

for higher sludge volume index (SVI) values with deeper tanks. Likewise, Figure 8.23 shows limiting overflow rates for various MLSS concentrations, an SVI of 100 mL/g, and different tank depths to obtain an effluent solids of 10 mg/L. For a given mixed liquor concentration, higher overflow rates can be obtained with deeper tanks (Parker, 1983; Voutchkov, 1992).

For tertiary clarifiers, adequate depth is required for an additional reason. The conical-shaped flocculating clarifiers are often equipped with more elaborate launders to remove effluent from over most of the tank's surface. Such tanks may require greater depths to provide true upflow into a radial launder pattern.

***Depth Determination.*** Most United States design engineers, in recent years, select tank depth as a function of diameter. Table 8.1a has served as a useful guideline.

There are more sophisticated ways of determining depth requirements for activated sludge clarifiers. One of the most sophisticated is the Abwassertechnische Vereiningung (ATV) (ATV, 1973, 1975, 1976, 1988, and 1991) approach that has been developed and used in Germany. In this approach, four functional depths are defined and added together to obtain the recommended minimum tank depth. For larger tanks, the common results often lead to depths of 4 m (12 ft) or more. Additional details and design example are presented in Chapter 4 of the International Association of Water Quality (IAWQ) Scientific and Technical Report No. 6 (IAWQ, 1997).

***Free Board.*** It has been common practice for United States engineers to use 0.5 to 0.7 m (1.5 to 2.0 ft) of freeboard for most clarifier designs. The origin of this range is uncertain, but this practice has been common for many years. This much freeboard

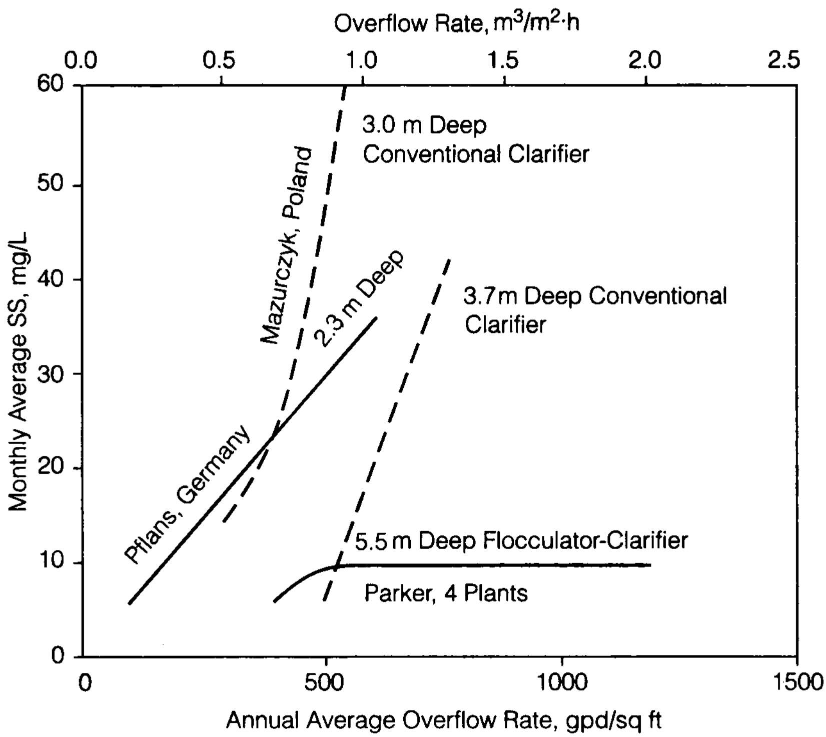


**FIGURE 8.20** Performance response curves for conventional clarifiers and flocculator clarifiers (gpd/sq ft × 0.001 698 4 = $m^3/m^2 \cdot h$).

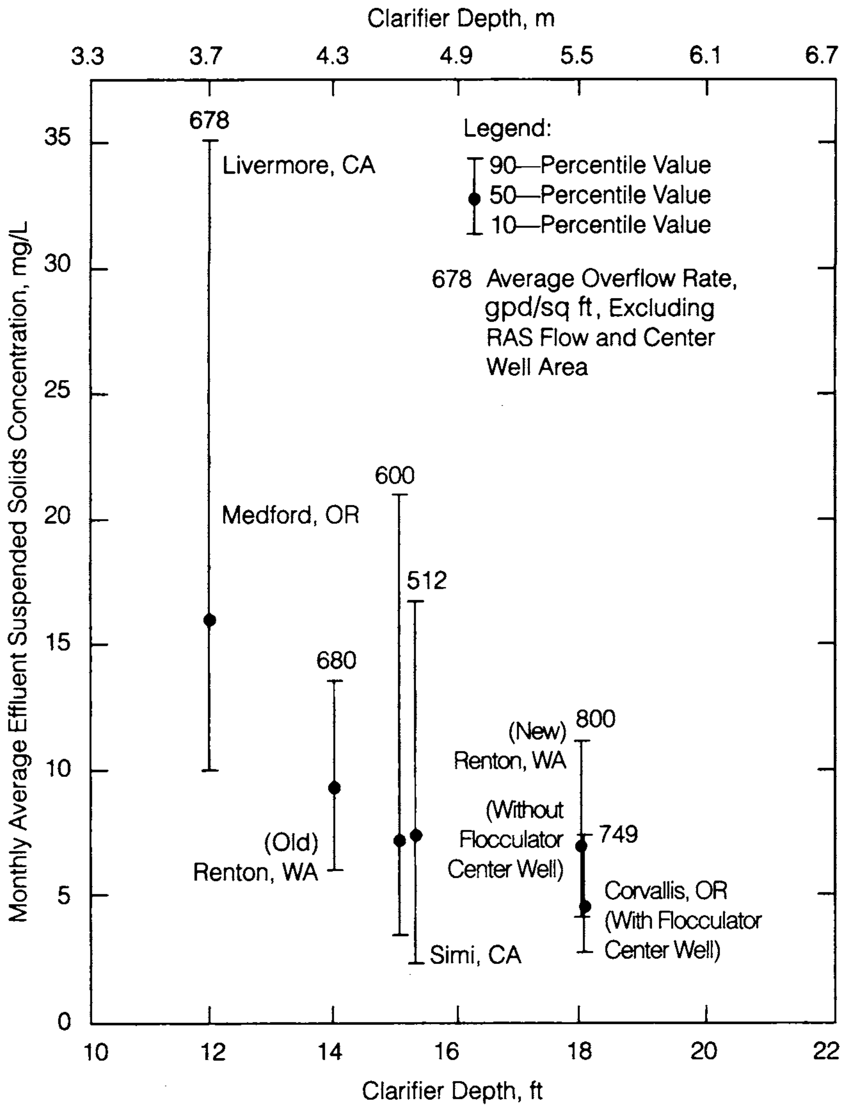


FIGURE 8.21 Effect of clarifier's depth and flocculator center well on effluent suspended solids (gpd/sq ft × 0.001 698 4 = $m^3/m^2 \cdot h$).

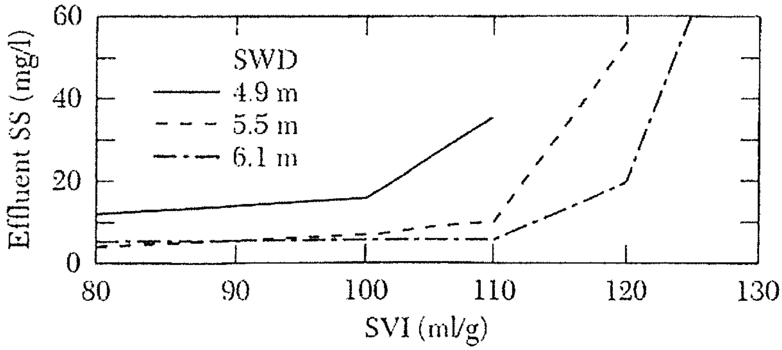


**Figure 8.22** Effluent suspended solids as a function of secondary settling tank depth and SVI, based on a two-dimensional hydrodynamic model.

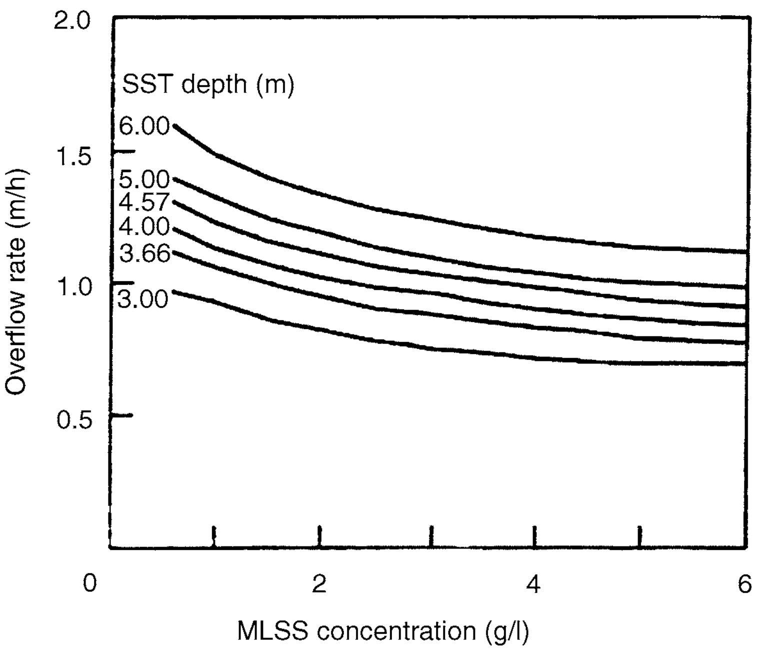


**Figure 8.23** Limiting MLSS concentration and solids overflow rate (SOR) to obtain an ESS of 10 mg/L for a sludge with an SVI of 10 mL/g. The figure shows that deeper tanks can accommodate higher SORs and MLSS levels. The interaction of MLSS and SOR is akin to a relationship of ESS to solids loading rate or sludge volume loading rate.

allows for downstream hydraulic problems (such as a pump failure or partial pipe inlet blockage) that may transmit upstream, possibly resulting in flooding of clarifier launders, without allowing overtopping of the tank walls. This amount of freeboard is generally sufficient to back flow up into the splitting structure and possibly even into the aeration basins of activated sludge systems.

***Integration of Walls and Handrails.*** Many circular clarifiers are equipped with handrails around the walls. Occupational Safety and Health Administration (OSHA) (Washington, D.C.) guidelines require handrails along walkways higher than 460 mm (18 in.) to be 1.07 m (3.5 ft) high. Careful design of clarifiers should enable final grading of the site topography to be coordinated with design of the concrete walls, in a way that eliminates the construction of handrails on top of the walls. If this can be done, the resulting design is more favorable toward operators leaning over the walls to hose down weirs. Extension of the walls, in some cases, also reduces the effects of wind on the clarifier water surface. This, in turn, can affect the movement of floatables and tank performance.

**OUTLETS.** Outlets for most circular center-feed clarifiers consist of a single perimeter v-notch weir that overflows into an effluent trough. Alternatives to this include cantilevered or suspended double weir troughs and submerged-orifice collector tubes. Open-trough options are shown in Figure 8.24.

For peripheral-feed designs, a singular perimeter weir is used in one concept. Another includes provisions of a square, octagonal, or circular double-sided launder suspended from the bridge or other structural support near the center of the tank.

In many states, regulations allow the weir loading that results from simply building a perimeter weir. In others, regulations include a weir loading limit expressed in flowrate divided by length of weir. For example, the Ten State Standards (Great Lakes-Upper Mississippi River Board of State and Provincial Public Health and Environmental Managers, 1997) limit weir loading to 250 $m^3/m^2 \cdot d$ (20 000 gpd/ft) for plants with average flows less than 0.04 $m^3/s$ (1 mgd) and to 375 $m^3/m^2 \cdot d$ (30 000 gpd/ft) for larger plants.

***Peripheral Weir.*** There are two common designs for peripheral weir outlets for circular tanks. In the first, a concrete trough is constructed on the inside of the tank wall. The weir plate is then bolted at the top of the inward face of the trough wall (Figure 8.24d). The other common arrangement for a perimeter weir is shown in Figure 8.24a.

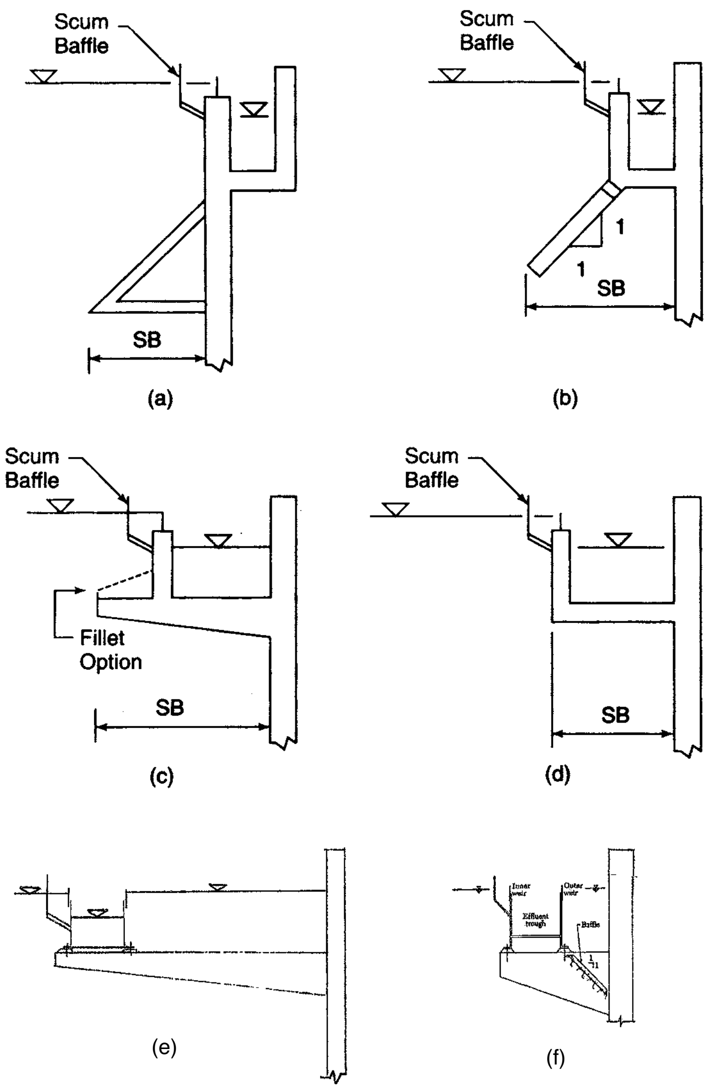


FIGURE 8.24 Alternative peripheral baffle arrangements: (a) Stamford, (b) unnamed, (c) McKinney (Lincoln), (d) interior trough, (e) cantilevered, and (f) cantilever with deflectors.

The weir plate is bolted to the inside of the tank wall. A concrete effluent trough is then constructed outside of the tank wall.

From a construction standpoint, the most cost-effective design between these two options is still a subject of debate. Most contractors like to construct the launder on the inside, allowing a smooth, vertical face on the outside to compact backfill. Other contractors prefer to compact backfill up to the tank wall and then build the external launder on top of the backfill without construction of as much formwork.

From a performance standpoint, it is generally preferred to have the trough extend inward. This helps deflect some of the "wall effect" solids updraft inward near the surface and prevents the loss of suspended solids over the weir.

The most common type of weir plate involves the placement of 90-degree v-notches at 152- or 304-mm (6- or 12-in.) intervals. This design allows a balance of relatively low increases in tank elevation when flows increase as a result of diurnal and other changes and allowance for imperfect leveling with reasonably good distribution. In contrast, a flat weir plate is very susceptible to unbalanced withdrawal if the weir is not perfectly level or if wind effects on the surface are significant.

Square notches are also used by some design engineers. This design will result in a wider range of level changes with flow changes and is more prone to partial notch blockage resulting from leaves, algae strings, and other surface debris.

The proper sizing of troughs is an important aspect of design. Hydraulic formulas for doing this are outside the scope of the text. References such as that by Boyle (1974) and Fair and Geyer (1963) are recommended.

***Cantilevered Double or Multiple Launders.*** A typical cantilevered double launder is illustrated in Figure 8.24e. In early years and perhaps in some areas today, regulations that limit weir loading to sufficiently low values have enticed designers to use multiple weirs, serpentine weirs, and other ways to increase weir length for a given diameter of tank. Requirements, in recent years, have been relaxed in many design guidelines. Nevertheless, this cantilever double launder concept remains. For activated sludge, it offers the opportunity for solids moving up along to the wall to resettle before their inward flow takes them to the outer weir. Anderson (1945) and others recommend that the outer weir be at least 25% of the tank radius from the wall.

***Launders Suspended from the Bridge.*** For some small, circular tanks, the double-sided launder design discussed in the previous subsection are suspended from the bridge. Necessary structural trusses are constructed to stabilize this form of outlet.

This concept has received most widespread use with peripheral feed clarifiers that use the spiral influent design. Peripheral feed tanks that use orifices for feeding often have an inward projecting trough that is constructed with a wall common to the feed trough.

***Full-Surface Radial Launders.*** For tertiary clarifiers, full-surface radial launders are often provided. These help create a more uniform updraft above the solids blanket. A similar, although generally rectangular, arrangement is often made for the withdrawal zone above tube and plate settlers.

***Submerged Orifices.*** A relatively small number of circular tanks have been constructed with submerged orifices for effluent removal. A typical design has a circular pipe located near the wall with evenly spaced circular orifices cut into the top. A downstream hydraulic control device is required to maintain a level within the clarifier. This design has the advantage of full-surface skimming. It also allows greater fluctuation in clarifier water level elevation with minimal performance effects. Another advantage is less splashing and odor release.

Some designs in Europe have included radial orifice pipes cantilevered from the wall. These have not received significant use within the United States and are not discussed further in this text.

***Safety Concerns and Provisions.*** Many operators take various measures to keep the appearance of the clarifier effluent troughs neat and tidy. For primary clarifiers, stringy material may be caught in the weirs, and, in secondary clarifiers, algae growth can become problematic. Therefore, to clean the weirs and launders, operators often lean over the tank wall and hose these areas clean or even put on waders and walk into the troughs themselves. It is important that design engineers provide features to minimize accidents and injuries to operators that perform these chores. To this effect, the troughs should be made approximately 1 m (3 ft) wide for most large tanks. Generally, portable ladders are provided for operators to access the troughs. Provision of hose bibs at approximately 30 m (100 ft) or less intervals around the tank is recommended to facilitate hose down and minimize the necessity for operators to enter the troughs themselves.

***Performance Comparison.*** The relative merits of the different types of outlet launder structures have been debated for years. Buttz (1992) conducted side-by-side, full-scale clarifier tests to compare baffled peripheral weirs with cantilevered inboard double launders. The results, exhibited as effluent suspended solids versus hydraulic loading rate, are shown in Figure 8.25. The data indicate no significant difference between these designs for overflow rates, up to approximately 2.5 m/h (1500 gpd/sq ft). Authors of the IAWQ technical report (IAWQ, 1997) point out that these studies were conducted with a shallow sludge blanket. It remains uncertain if these results represent comparable performance with deeper blankets.

**INTERIOR BAFFLES.** For many years, circular clarifiers were constructed without interior baffles, except for the inlet well. In the 1970s and early 1980s, research engineers, such as Crosby (1980), McKinney (1977), and others, found that the performance of activated sludge clarifiers could be significantly improved by the provision of strategically located interior baffles. The addition of a large flocculation baffle has been previously discussed in this chapter. Another baffle that was found effective to help confine the sludge blanket to central portion of the clarifier is illustrated in Figure 8.26. This baffle, extending up from the bottom, was initially developed by Crosby. It has been referred to as his "mid-radius" baffle. It has not been used often, if at all, in new tank design, but it has been used at times for modifications to improve performance of existing tanks.

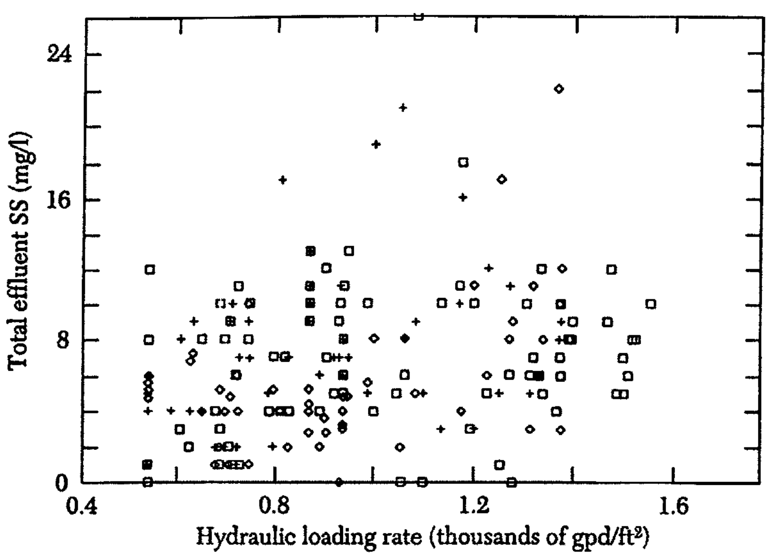


FIGURE 8.25 Comparison of performance of flocculator clarifiers with weir baffles and inboard weirs. Symbols: □, flocculator clarifiers with inboard weirs; +, ○, flocculator clarifier with baffled peripheral weirs (588 gpd/sq ft = 1 m/h).

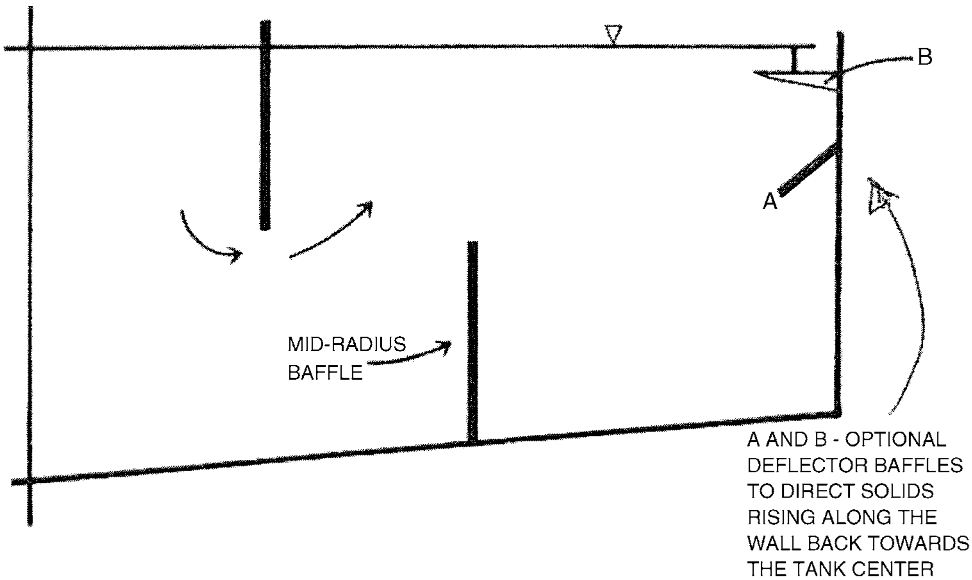


**FIGURE 8.26** Baffles provided to reduce effect of outer wall rebound and upflow (note: a tank typically would not have more than one such baffle).

Center-feed activated sludge clarifiers often create an updraft of suspended solids along the outer wall. Early studies by Anderson (1945) in Chicago, Illinois, revealed the presence of this movement. His response was to construct double launders sufficiently distant from the wall, allowing the updraft solids to resettle before the effluent reached a weir. Crosby (1980) and McKinney (1977) independently arrived at another solution, namely, constructing a perimeter baffle to deflect this flow back toward the center of the tank. The conceptual design of these two options is shown in Figure 8.26. Further refinements in this design are illustrated in Figure 8.24a, b, and c. For designs with the trough on the outside of the tank wall, the Crosby design shown in Figure 8.24a is most appropriate. For tanks with the trough on the inside of the tank wall, the three options shown in Figure 8.24b, c, and d have been used. The dimension SB on this figure needs to be several meters for deflection to be effective. The minimum size is a function of tank diameter and can be calculated by the following formula (WEF, 1998):

In metric units:

$$\text{Minimum SB} = 460\ \text{mm} + 25\ \text{mm/m}\ (D - 9.15)\text{m} \tag{8.3a}$$

Where

SB is in millimeters,
a = 16.7 (or 25) millimeters per meter, and
D = tank diameter is in meters.

Or in English units:

$$\text{Minimum SB} = 18 + a\,(D\text{-}30) \tag{8.3b}$$

Where

SB is in inches,
a = 0.2 (or 0.3) inches per foot, and
D = tank diameter is in feet.

This basic formula was recommended by Albertson (in preparation of Water Environment Federation (WEF) Manual of Practice (MOP) 8 [1998]) and initially had a = 16.7 mm/m (0.2 in/ft); however, in more recent (2002) personal communications, the higher value of 25 mm/m (0.3 in/ft) was recommended. Stukenberg et al. (1983) recommended the design of Figure 8.24c without the fillet. His recommendation was to extend the bottom baffle 0.6 m (2 ft) inside of the weir location, regardless of trough width. If the above MOP 8 formula with the larger "a" coefficient is used, the following values of minimum SB values result as shown in Table 8.1b.

For cantilevered double launders that are constructed too close to the tank wall, updraft solids commonly escape. Parker et al. (1993) developed a special baffle, oriented at 45 degrees from horizontal, to deflect the updraft solids away from the wall

**Table 8.1b** Minimum SB values.

| Tank diameter, m (ft) | Minimum SB, m (ft) |
|---|---|
| 20 (66) | 0.73 (2.4) |
| 30 (98) | 0.98 (3.2) |
| 40 (131) | 1.23 (4.0) |
| 50 (164) | 1.48 (4.9) |
| 60 (197) | 1.73 (5.7) |

and below the effluent trough for clarifiers of this design (Figure 8.24f). The baffle was constructed of strips of fiberglass roofing material that spanned from one support bracket to another. A spacing of approximately 35 to 50 mm (1.5 to 2 in.) was allowed to permit a small flow to rise and leave the outer weir. Most of the flow, however, was deflected to the inner weir, and the suspended solids were projected toward the tank center. At Lincoln, Nebraska, this arrangement reduced effluent suspended solids from 35 to 28 mg/L.

Others have attempted to minimize the updraft problem by reducing the number of notches or raise the outer weir of such a design to encourage most of the flow to leave the tank by way of the inner weir. Blocking the outer weir completely is not recommended because it then creates a dead space between the outer weir and the wall.

Convenient peripheral baffles are now available commercially. Two are shown in Figure 8.27. The upper configuration in this figure shows a molded fiberglass panel with intrical bracket stiffeners. These are used for tanks with external launder troughs. For those with internal troughs, a similar molded design is shown at the bottom of the figure. Note that each of these show molded vents that allow escaping gas bubbles to vent from below the baffle and eliminate or reduce the phenomenon of large bubbles coming from below these baffles.

**SLUDGE REMOVAL SYSTEMS.** Effective removal of sludge from circular tanks is vital to process performance. The rotating mechanisms of circular clarifiers are touted as one of the most important reasons to choose the circular shape. There are two basic types of sludge removal mechanisms, namely plows and hydraulic suction. Plows are used for all kinds of sludge encountered in wastewater treatment, whereas hydraulic suction is primarily limited to activated sludge secondary settling tanks.

For square tanks, spring- or counterweight-loaded, corner-sweep plows are used to gather from the corner areas outside the fixed-sweep circular area. If tanks are sufficiently deep, filleted tank walls can be used to fill in the corners and thereby use circular mechanisms without corner sweeps. This is preferable because corner sweeps are notorious for mechanical problems.

***Scrapers.*** There are several basic scraper designs used. Figure 8.28 shows four different types of scrapers. The multiblade plows, using straight scraper blades, have been used most extensively in the United States. The designs using curved blades are commonly referred to as *spiral plows* and have been used for decades in Europe. Based on encouragement from Albertson and Okey (1992) and others, United States engineers are now choosing spirals in an increasing percentage of clarifier designs.

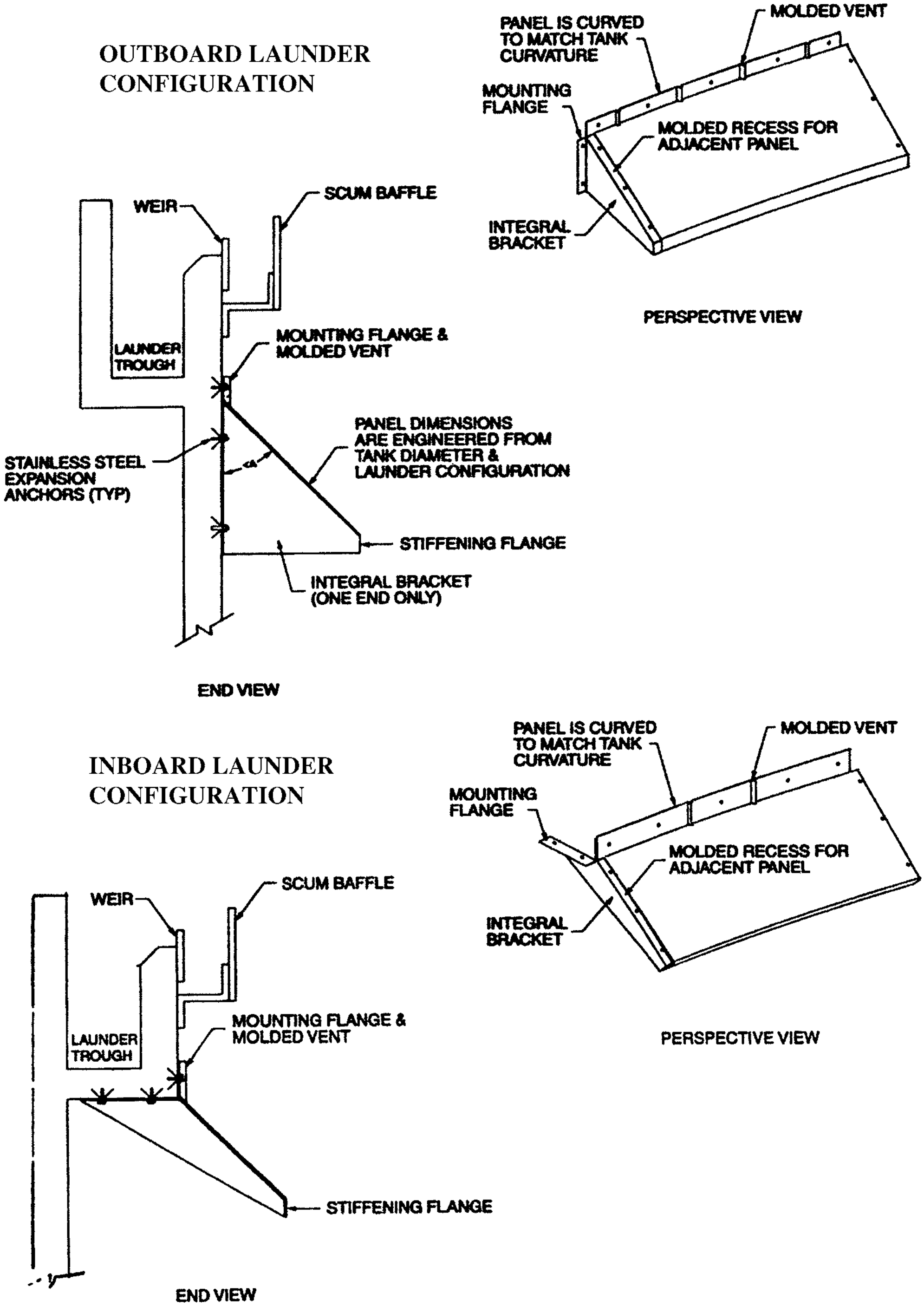


**Figure 8.27** Two types of commercially available peripheral baffles.

The straight, multiblade design (form C in Figure 8.28) remains the most widely used mechanism for primary and fixed film secondary tanks. For these heavy, relatively viscous sludges, the scrapers plow furrows of sludge progressively toward the centrally located sludge hopper. These mechanisms are commonly designed to rotate at a tip speed of approximately 3 m/min (10 ft/min).

For the lighter suspended growth sludges, spiral collectors (Figure 8.29) are becoming more dominant. Although blade angles of between 15 and 45 degrees have been used, 30 degrees has become popular in the United States.

In early years, the tip speed of spiral scrapers used in this capacity was also approximately 3 m/min (10 ft/min). Based on several plant improvement projects, Albertson and others have recommended values as high as 10 m/min (30 ft/min). These faster speeds, along with deepening the spiral blades closer to the center of the tank, give this system a relatively high sludge transport and removal capacity. The

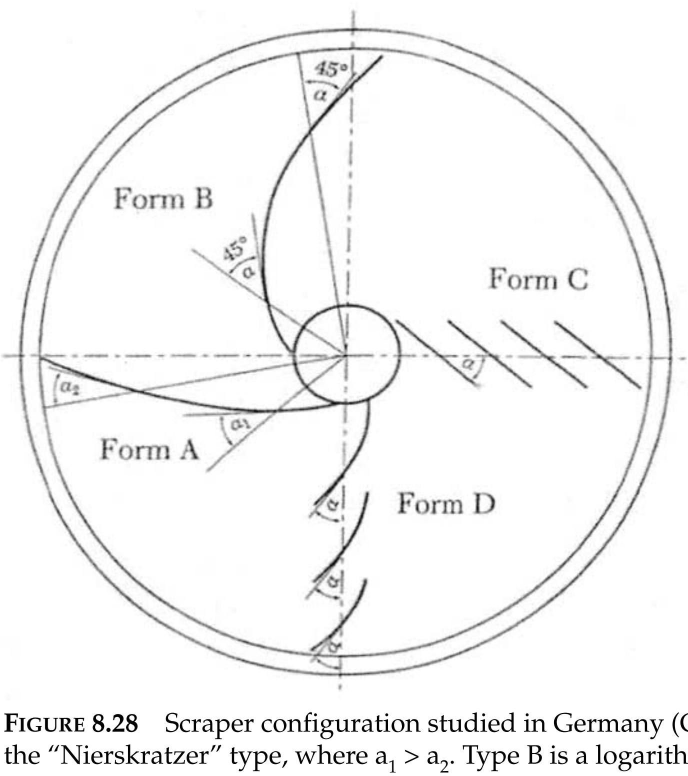


**FIGURE 8.28** Scraper configuration studied in Germany (Guenthert, 1984). Type A is the "Nierskratzer" type, where $a_1 > a_2$. Type B is a logarithmic spiral, with a constant at 45°; and types C and D are "window shade"-type scrapers.

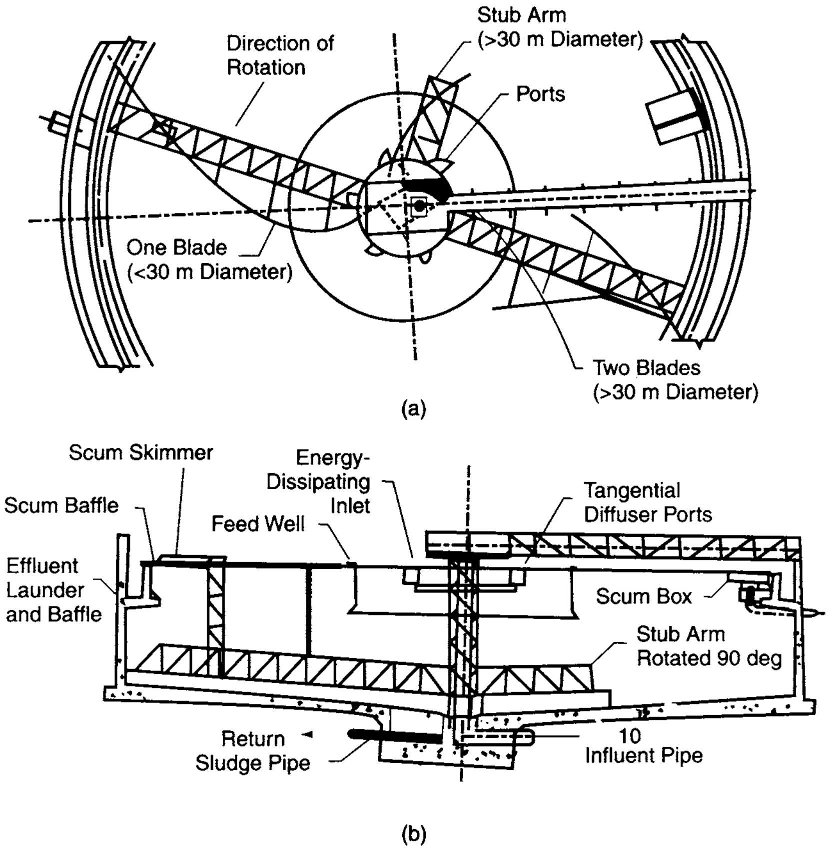


**Figure 8.29** Spiral sludge collector example: (a) plan view and (b) illustrative elevation.

higher speeds do induce some stirring of the tank contents, especially in smaller tanks, and may be detrimental to clarification. Some designs now provide variable speed drives to give operators the opportunity to increase or decrease tip speed as needed to balance sludge pumping capacity against effluent quality.

***Hydraulic Suction.*** For activated sludge treatment with partial or complete nitrification, the occurrence of denitrification in clarifiers can cause solids to float and effluent quality to degrade. In the 1960s, the concept of hydraulic suction was engineered to assist in removing sludge more rapidly, and it was believed to maintain lower sludge blankets. The data of Figure 8.30 shows that increasing sludge blanket

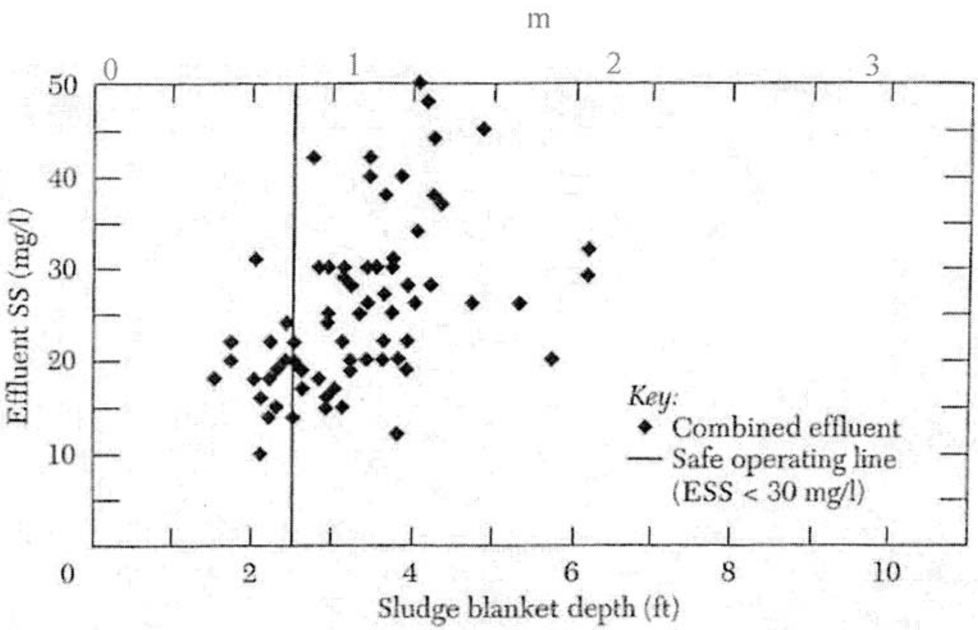


**FIGURE 8.30** Effect of sludge blanket depth on ESS at pure oxygen activated sludge plant; SVI = 51 to 166 mL/g, average 86 mL/g.

depth results in higher effluent solids in some cases. Some plants do obtain low effluent solids with deep blankets above the level of the bottom of the flocculation baffle by forcing inflow to rise up through the blanket enroute to the effluent weir. Increases in flowrate for such operations can make the blanket rise and lose solids, so most operators prefer to keep the blanket shallow.

The data of Figure 8.31 show that the sludge blanket depth of a pure oxygen activated sludge plant final clarifier is deeper at given flow and loading conditions when a scraper mechanism is used compared to a hydraulic suction sludge removal design. Similar findings were reported by Kinnear (2002), who compared hydraulic suction to spiral plows serving circular activated sludge tanks.

Hydraulic suction mechanisms lift solids from across the entire tank radius. The concept has been in use for well over 50 years. A hydraulic head deferential is established by use of pumps or adjustable valves to move solids into the collector arms. There are two fundamentally different types of hydraulic suction removal mechanisms. The first, commonly called an *organ pipe* or *riser pipe* type, has a sepa-

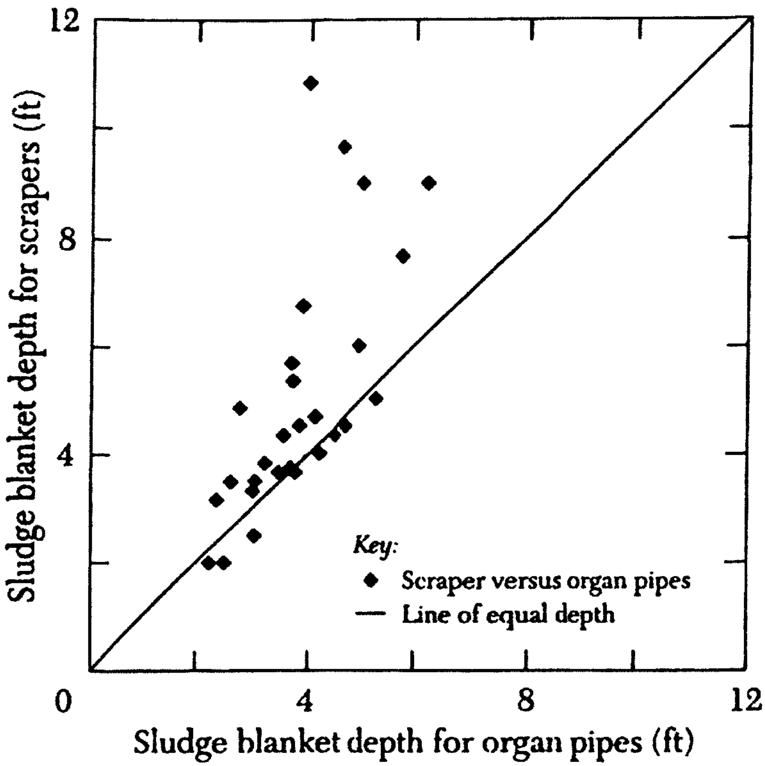


**FIGURE 8.31** Comparison of sludge blanket depths for scraper mechanisms at a pure oxygen activated sludge plant. Most of the data points fall above the line of equal depth, showing that the suction sludge removal SST maintains lower sludge blanket depths (1 m = 3.28 ft).

rate collector pipe for each suction inlet orifice. V-shaped plows direct the sludge to the multiple riser pipes.

The other type has a single or double arm extending across the full radius of the tank. The arm is tubular and has a number of orifice openings. It is commonly referred to as a manifold design but is also known as header, tubular or Tow-Bro, in recognition of Townsend and Brower, who developed it.

***Riser Pipe Mechanisms.*** Figure 8.32 shows a typical riser pipe clarifier design. The horizontal runs of the riser pipes are stacked vertically—an orientation that induces more tank stirring than the option of having these pipes in the same horizontal plane. Most designers prefer the latter. Each riser pipe of this mechanism is fitted with an

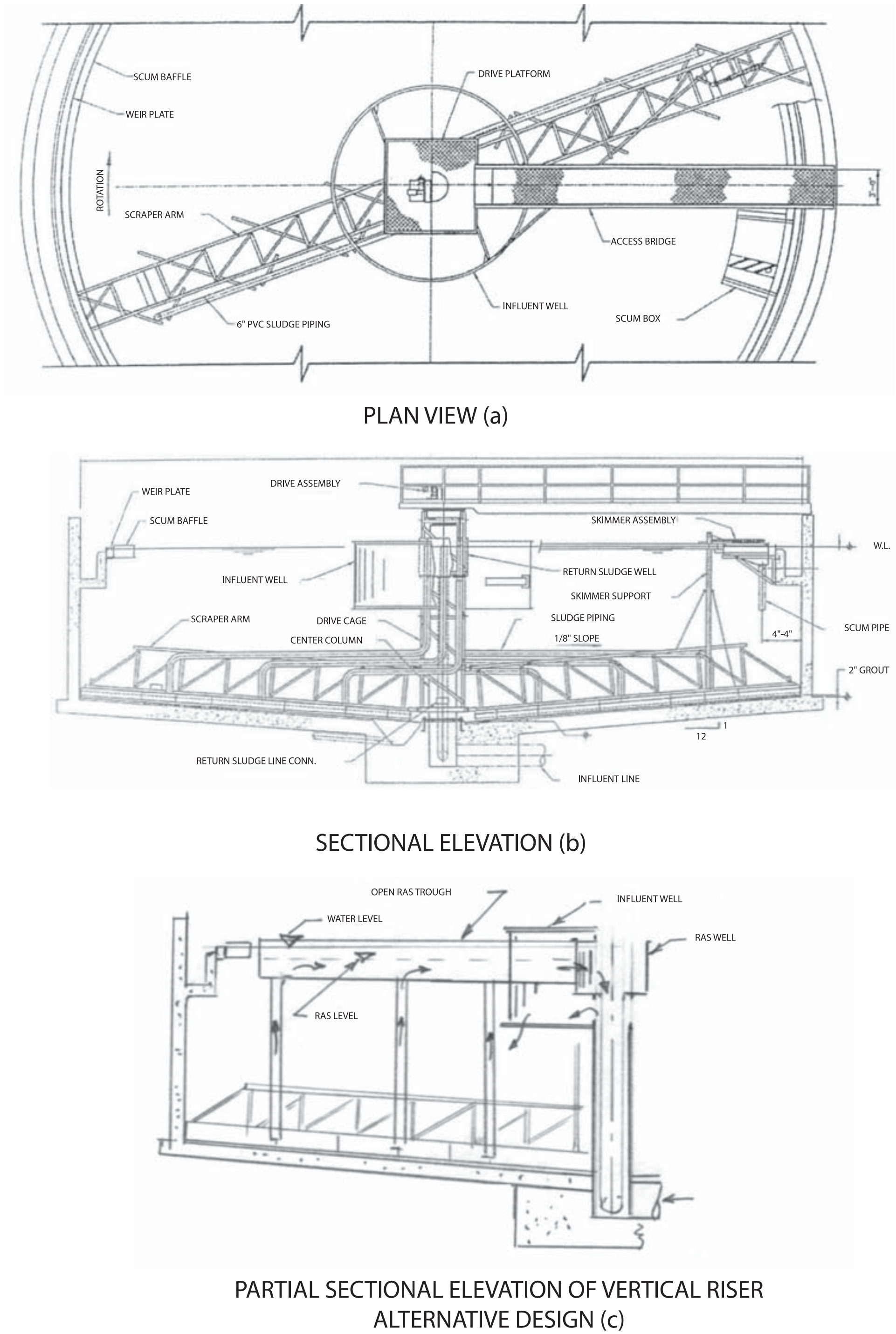


**FIGURE 8.32** Hydraulic sludge removal design with suction pipes.

adjustable telescoping weir, movable sleeve, or ring arrangement that allows the operator to adjust the flow independently for each suction inlet. If the telescoping valve discharges as a free fall, it is easier to take samples for suspended solids analyses. However, separate adjustments are needed to change the overall rate of sludge withdrawal from the tank. When tube discharges are submerged, the overall sludge flow through all tubes is a function of the deferential head between the tank surface and sludge level above the tube outlets in the collection box. A change in the level difference, such as by increasing the withdrawal pumping, increases flow through all tubes. The relative proportions of the total flow do not change except for friction loss differences resulting from variations in the length of the suction tubes. A change in RAS flowrate changes the velocity in the pipes, and relative head loss optimization would require adjustment of the individual tube flow control settings. At least one manufacturer has all tubes the same length. They all rise vertically into a trough instead of a box. This allows changes in RAS rate without changing the relative flows among risers.

The riser pipe design offers an operator the advantage of adjusting sludge flowrates from different points along the tank radius. Well-trained operators often argue that they can optimize tank performance by adjusting the relative flow from each tube. In some designs, adjusting these valves can be dangerous; it may require operators to leave the safety of the walkway and platform to properly adjust the valves. Visual observations are often inadequate to truly optimize the withdrawal pattern. Sampling of solids, followed by suspended solids measurement, is, in many cases, not convenient. Some operators have taken samples and centrifuged them to obtain quick measures of relative concentrations of sludges from the tube.

Some of the riser pipe designs allow operators to rapidly backflush each tube. Thus, if rags or other things clog them, they can be rapidly restored to full performance.

Riser pipe designs often result in the tubes that enter the collection box to pass in front of the central feed pipe orifices, which are located just below the box. This interference can deflect the inflow and may result in some jetting of flow into the EDI or central feedwell. The relative sizes of pipe, their orientation, and proximity to the orifices need to be considered in plant design and specifications.

Riser pipe clarifiers include a mechanical seal between the center column and each return sludge well. If the seal leaks, there is a loss of or decrease in water level differential between the tank water surface and the level in the well. Lower differentials can result in lower RAS rates or even no sludge removal.

***Manifold.*** The manifold type hydraulic suction mechanism contains multiple orifice openings along its radial length. Some clarifiers have a single tube, whereas larger ones have two tubes opposite each other. Figure 8.33 shows an example of this.

The orifice openings are sized and spaced in the factory to obtain a near-optimum pattern of collecting solids from the floor. The hydraulic formulas and orifice-spacing criteria are beyond the scope of this text. Some designs do have adjustable orifice openings, but it is necessary to take the tank out of service to make the required adjustments. In view of this, most operators do not change the settings, even though adjustments are possible.

With the manifold design, the plugging of any particular orifice can not be checked without dewatering the tank. Instrumentation that compares the flow to head loss can be used to determine if some plugging has occurred, and some RAS stations are designed to provide backflushing to solve a plugging problem.

The manifold type of hydraulic suction device has gained in popularity relative to the riser pipe alternative. The main advantage of the manifold is that it can be coupled directly to RAS pumps or to an RAS wet well with a substantial head deferential. This allows suction of relatively dense sludges, and supporters of the design that argue lower RAS flowrates are feasible. Results of a side-by-side test at a pulp and paper plant in the state of Washington has reportedly shown that the manifold type was able to obtain 1% RAS, whereas that from a parallel riser pipe was only 0.6% (IAWQ, 1997). At that plant, WAS is removed exclusively from the manifold equipped mechanisms.

A serious design issue with the manifold device is obtaining a good seal at the bottom. This device requires two seals, one at each side of the rotating collar that moves with suction tubes. In some early designs, the combination of silt or grit deposition and suction from the RAS pumps led to abrasion and wear of the seal materials. If wear is excessive or suction adequate, relatively low TSS water can be sucked through these seals, thereby defeating the purpose of hydraulic suction. This leaking water can dilute the RAS and eventually reduce flows through the orifices. Replacing the seals requires tank dewatering.

There are two or more successful modern seal designs. Figure 8.33 (detail A) shows a design with the seals located several inches above the floor to reduce the problem of grit abrasion.

***Hoppers.*** Until recent years, most United States clarifiers using scraper mechanisms were equipped with trapezoidal hoppers. An example of this is shown in Figure

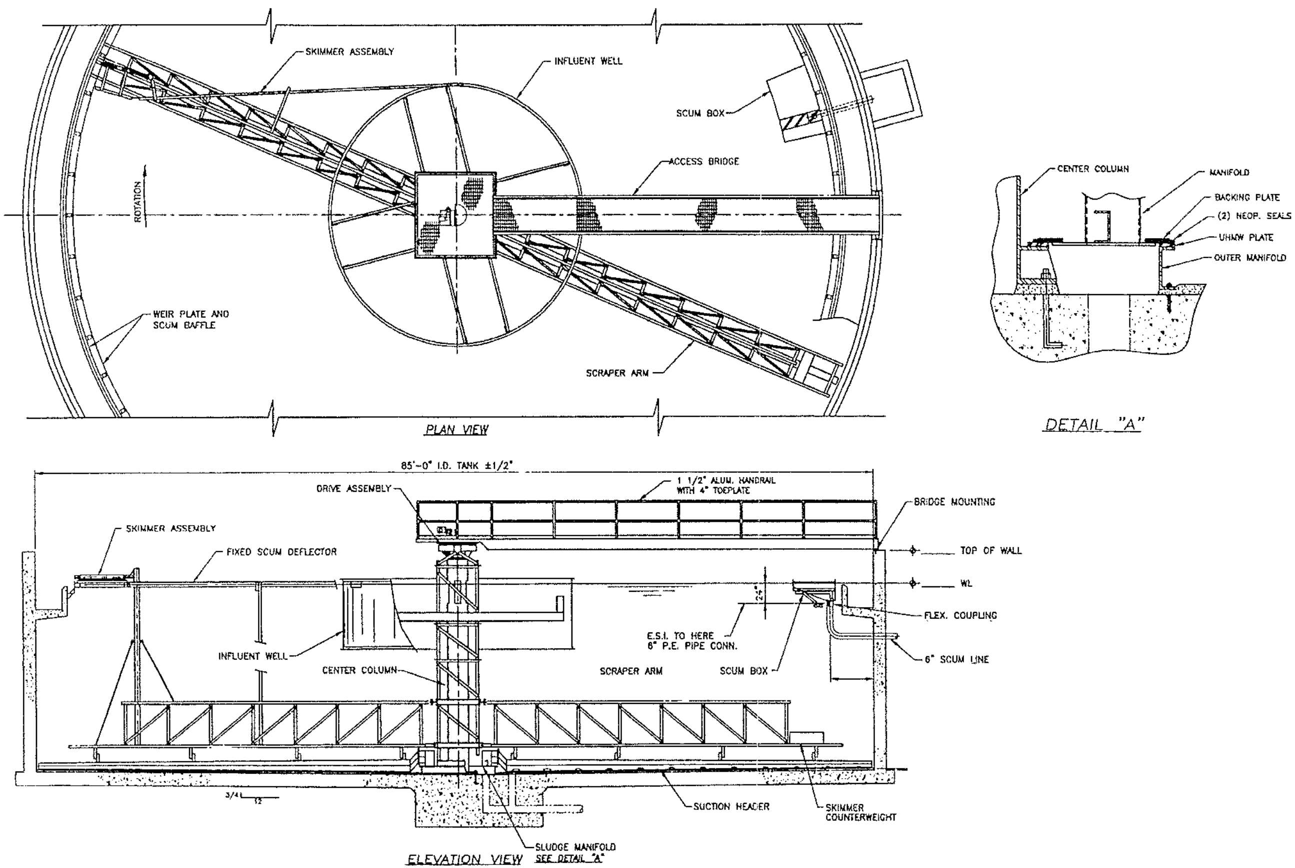


**FIGURE 8.33** Hydraulic sludge removal using typical suction header (or tube) design.

8.34a. Depending on tank size, these hoppers are typically several meters deep and have walls with slopes of at least 60 degrees above horizontal. This type of hopper is still the most prevalent for primary and fixed-film secondary clarifier design.

For suspended growth systems, other types of hoppers have been developed to prevent "ratholing" and dilution of the RAS. One type consists of deep conical or concentric sludge hoppers, as shown in Figure 8.34b. The rotating mechanism has stirrups that reach into the annular hopper to prevent bridging.

Another design concept was to make the sludge hopper longer and narrower and extend radially outward a distance of up to 25% of the tank radius. A place with several orifices was used to withdraw the sludge more uniformly over this larger radial distance. Details of this design are given by Albertson and Okey (1992).

***Collection Rings and Drums.*** Even for tanks with hydraulic suction, some engineers design a separate deep trapezoidal hopper at the bottom of the activated sludge clarifiers from which to waste sludge. They believe that thicker sludge can be achieved in this way.

In more recent years, the radial hopper discussed above has been replaced by sludge rings, sludge drums, and other such variations to assist in removal of sludges plowed to the center by spirals or multiple plows. Two such devices are shown in Figure 8.35a and b. In the sludge ring design, an annular area with multiple orifices is provided to remove sludge continuously from a full radius around the center column.

Some engineers have been concerned about the plugging potential of sludge ring orifices. In view of this, the sludge drum was developed. This design has only two large openings, one at the interior end point of each spiral blade. The opening is fixed relative to the blade end and the drum therefore rotates with the mechanism. There are other minor variations of this fundamental design. One is shown in detail A of Figure 8.33. In this design, only a flat "washer shaped" top plate of the drum rotates and two inverted U-tubes lift the RAS into the drum. Some of these devices are patented and are still relatively novel, but found to be effective in specific installations.

***Drive Location.*** Most clarifiers in the United States are driven off of torque applied to the center column or, for smaller tanks with side feed pipes, from a fixed bridge that spans the full width of the tank. In Europe, it is very common to have a drive located at the tank wall. This powers rubber-tired wheels that ride on the top of the tank wall and rotate the bridge that spans the tank diameter and is pivoted in the

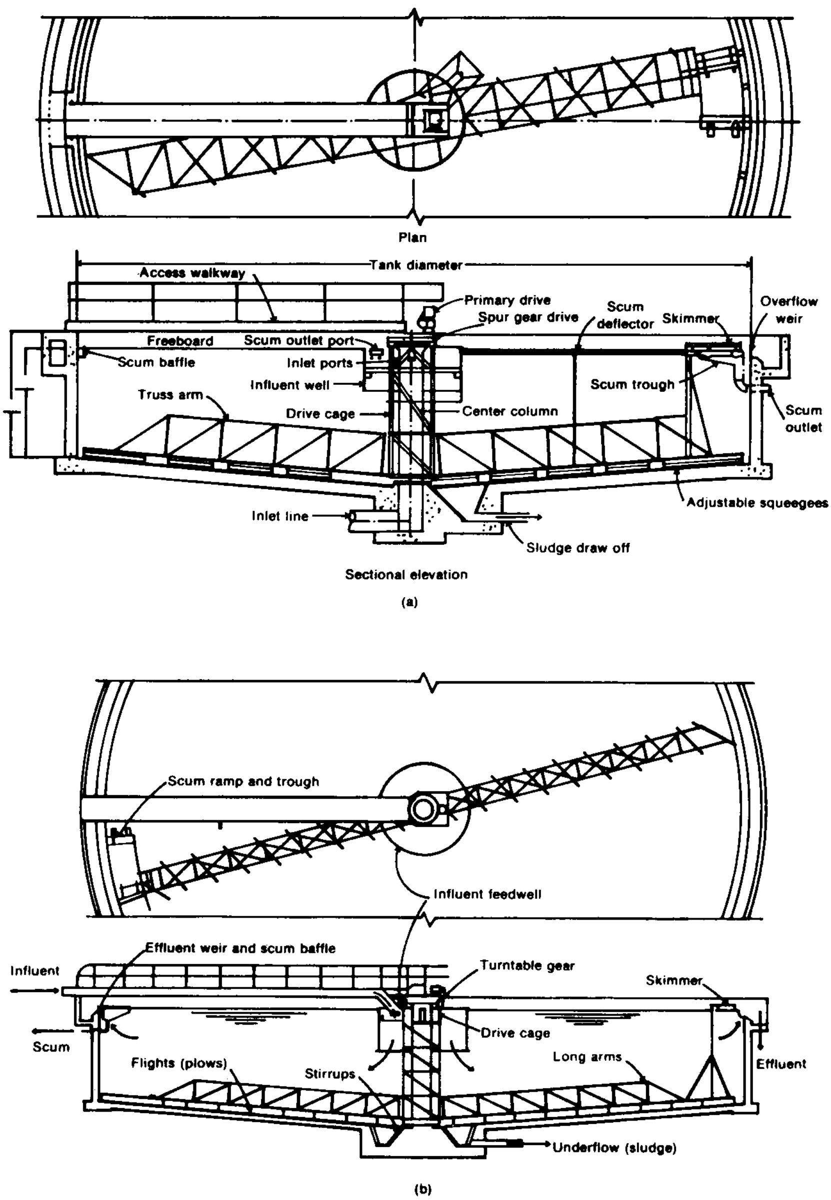


**FIGURE 8.34** Circular clarifier with (a) offset sludge hoppers and (b) concentric sludge hoppers.

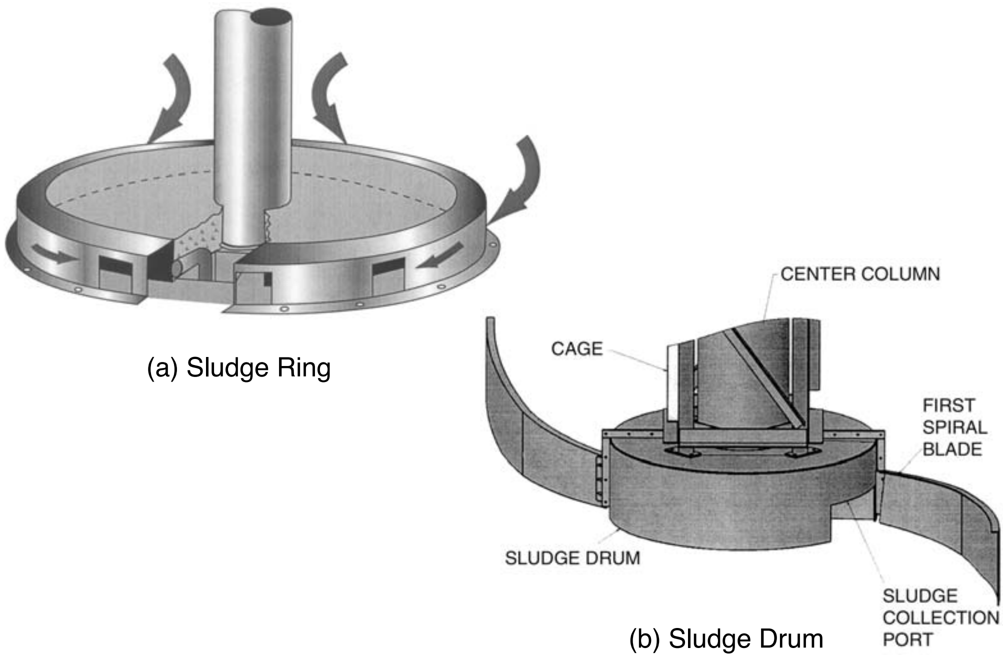


**FIGURE 8.35** The (a) sludge ring and (b) sludge drum to remove solids from activated sludge final clarifiers.

middle. There are some vendors in the United States that now carry these rim drive mechanisms, but they have not taken a major market share. Most United States design engineers have had concern about ice and snow interfering with the drive in cold climates, although these units do operate in Europe, where temperatures fall below freezing. Another difference is that the location of the bridge access stairs keeps moving and may not be in the most convenient location when someone wants to use it. It also makes collecting samples of the sludge blanket at right angles from the bridge difficult. In a recent study, it was necessary to tow a boat at this location to get a good cross-section of sludge disposition.

***Floor Slopes.*** Most clarifiers with plow or spiral mechanisms have a constant floor slope of 1 on 12. The true origin of this particular slope is uncertain, but it has received widespread use for decades.

In recent years, following the development of deep spiral collectors, Albertson et al. (1992) have promoted the use of a dual slope floor that provides a steeper slope in the central area for large tanks. This steeper slope provides for greater depth and sludge compaction.

For hydraulic suction clarifiers, a 1 on 12 slope may also be used. However, because it is not necessary to move the sludge across the floor, relatively flat floors are acceptable. Often a bottom slope of 1 or 2% is provided to facilitate tank drainage. Some design engineers actually prefer to reverse the slope and provide a gutter and mud drain valve at the perimeter of the tank to drain it.

***Return Activated Sludge Pumping Considerations.*** For activated sludge plants, a few comments are in order relative to RAS pumping. Many designers elect to couple the suction side of the RAS pump manifolds to the sludge removal hoppers or hydraulic suction mechanisms. These pump stations therefore do not have a wet well. They offer the advantage of not exposing the mixed liquor to air, where odors could be released or scum problems in the wet well could form.

It is important that a single pump be connected to each circular clarifier and not to more than one at a time. Such single, direct piping arrangements prevent the suction of dilute mixed liquor from one tank and reduced flows from another.

An alternative design provides for each clarifier sludge line to discharge into a wet well by way of a flow control valve. Such a valve allows independent discharge of sludges and separate control of each. The RAS pumps then operate on a level control signal to maintain the desired level in the wet well. In plants with many circular clarifiers, this arrangement offers the advantage of fewer RAS pumps. It does, however, generate the disadvantages of maintaining a wet well and its associated scum and odor problems.

It is vital to note that symmetry is never an acceptable principal to use to balance the withdrawal of sludge from clarifier hoppers. Independent control from each hopper is an absolute necessity.

**SKIMMING SYSTEMS.** The presence of scum and floatable material on the surface of clarifiers is a common problem in most municipal wastewater treatment plants. In primary clarifiers, the main contributors consist of grease and oils, plastics, leaves, rags, hair, and other materials. For clarifiers serving activated sludge and other suspended growth systems, scum formation is largely as a result of denitrifying sludges and foams (such as Nocardia filamentous bacteria) resulting from conditions in the biological treatment system. For secondary processes involving fixed-film biological treatment, the problem of scum and floatables is less severe, but, nevertheless, existing.

For tertiary treatment in which chemicals are added for phosphorous and suspended solids removal, scum formation is minimal, and some clarifiers at this stage do not provide for scum removal.

It has become common practice to remove floating materials from the surfaces of primary and secondary clarifiers in the United States. For circular tanks, a variety of skimming mechanisms have been designed and operate with varying degrees of capacity and success. The most common system used for center feed tanks is shown in Figure 8.36a. This figure illustrates the revolving skimmer with a fixed scum trough. This design has been used for many years and is considered by many to be the standard, especially for primary clarifiers. It features a rotating skimmer arm and wiper that travels around the outer edge of the tank next to a scum baffle. It moves the floatables onto a beach or egress ramp connected to a scum removal box (Figure 8.37). The skimmer blade is most effective if it is attached tangentially to the feed baffle, rather than perpendicular to it. The resulting pitch angle of the tangential design helps move floatables to the outer area of the tank. During each rotation, floating solids are pushed toward the egress ramp, where they leave the water surface, go up the ramp, and drop into the scum box. The skimmer blade then passes over the scum box and dips back into the water to repeat its rotation. Most primary clarifiers have one rotating blade per clarifier, whereas secondary clarifiers may have two or even four such rotating blades.

Some scum boxes are also equipped with an automatic flushing valve located on the centermost end of the box. The valve is mechanically actuated with each pass of the skimmer. It results in a water flush of the solids into the box hopper bottom and discharge pipe. The flush volume and duration are typically adjustable.

This scum trough often extends several meters (feet) from the scum baffle toward the center of the tank. Some designs extend this to the flocculating or center feedwell and thereby obtain full radius skimming. For shorter scum troughs, some system is generally provided to move the floatables toward the outer scum baffle. A fixed, flexible antirotation baffle, supported from the bridge and extended down to the surface of the tank, is sometimes used. The baffle is placed at an angle to the skimmer arm that intersects the tank water surface. The resulting scissorlike movement pushes the scum outward; this is illustrated in Figure 8.38.

Another method of moving floatables out toward the scum baffle is the use of water surface sprays. One effective design consists of spacing downcomer pipes a couple of meters (few feet) apart and placing a fan spray nozzle at the end of each. Figure 8.39 shows several positive features. The design uses two sets of

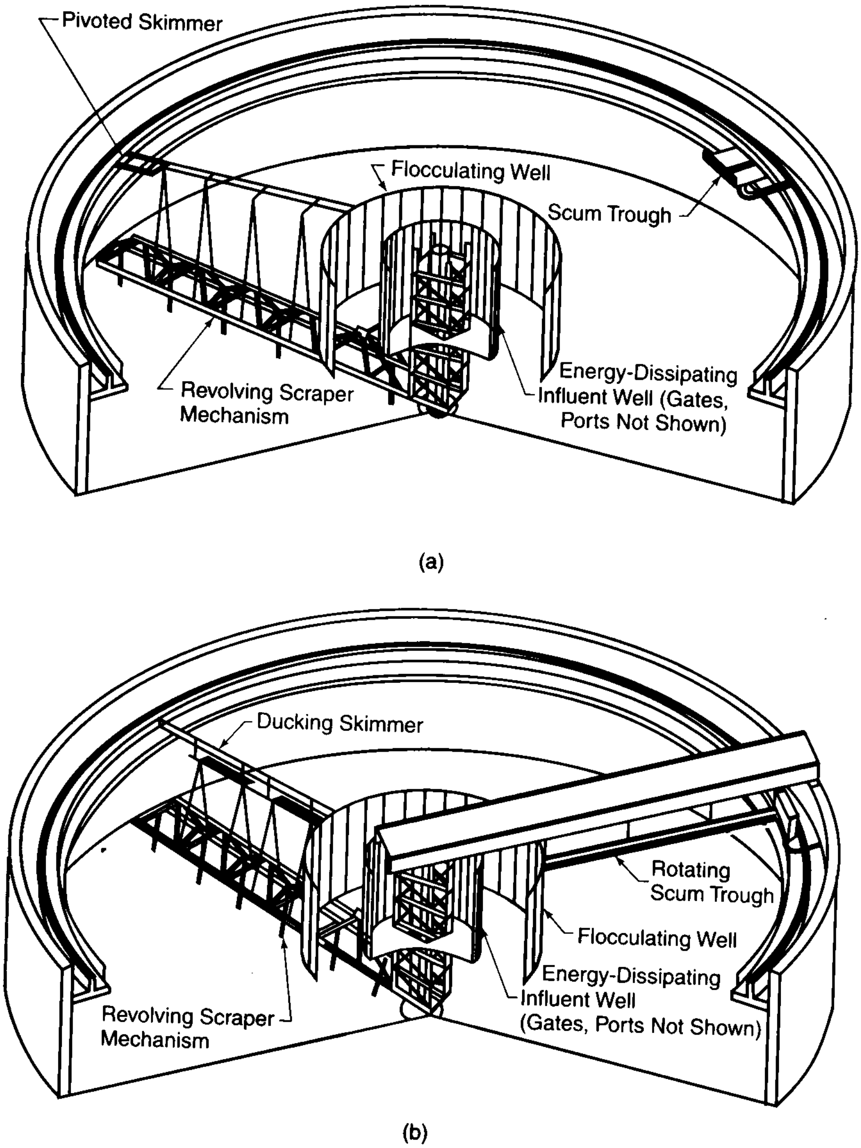


**FIGURE 8.36** Alternative skimming designs for circular clarifiers: (a) revolving skimmers and fixed scum trough and (b) rotary ducking skimmers.

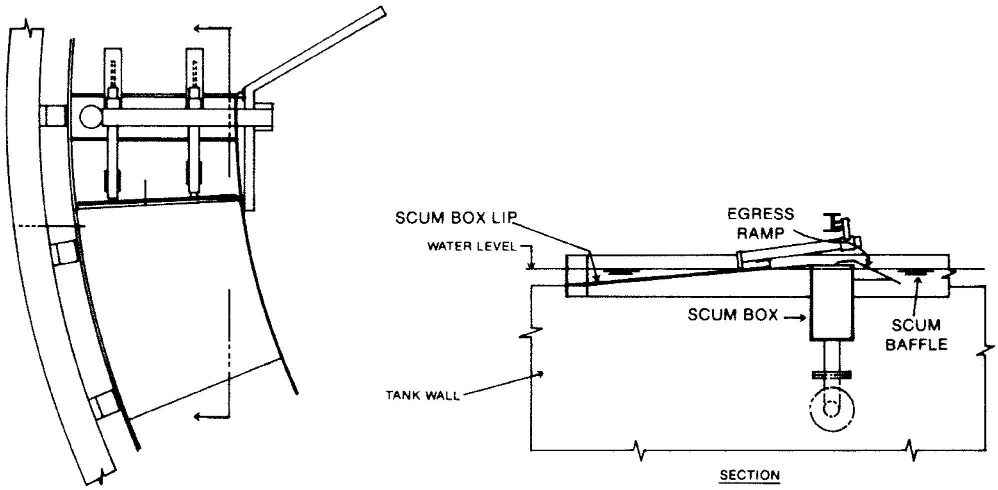


**FIGURE 8.37** Conventional skimming mechanisms for circular tanks.

double 90-degree threaded pipe elbows that permit field adjustments to optimize movement of floatable materials. The top of the pipe union is used to allow the arm to be removed for adjustment, cleaning, or replacement. The upper threaded 90-degree elbows allow the fan spray nozzle to be raised or lowered relative to the water surface. In addition, the lower pair of threaded elbows allows the fan spray angle to be field adjusted. It is necessary, of course, that the fan spray nozzle be placed close to the water surface but high enough to allow the skimming arm to pass below.

Blowers have also been used to push floatables to the outer edges of a tank, but installations are few in number. It is a good idea to locate the fixed scum beach on the downwind side of a tank. Yard piping arrangements may or may not make this option economically attractive.

Center feed tanks offer the problem of moving floatables from the feedwell out into the larger area of the tank served by skimming devices. For primary clarifiers, the small center wells can be designed with port openings. The turbulence within the well is generally effective in moving the floatables through these ports. For secondary clarifiers, ports or gates are generally adequate to allow the movement of floatable material from the EDI out into the flocculation well if tanks have this feature. However, the area within the flocculation well can be a significant

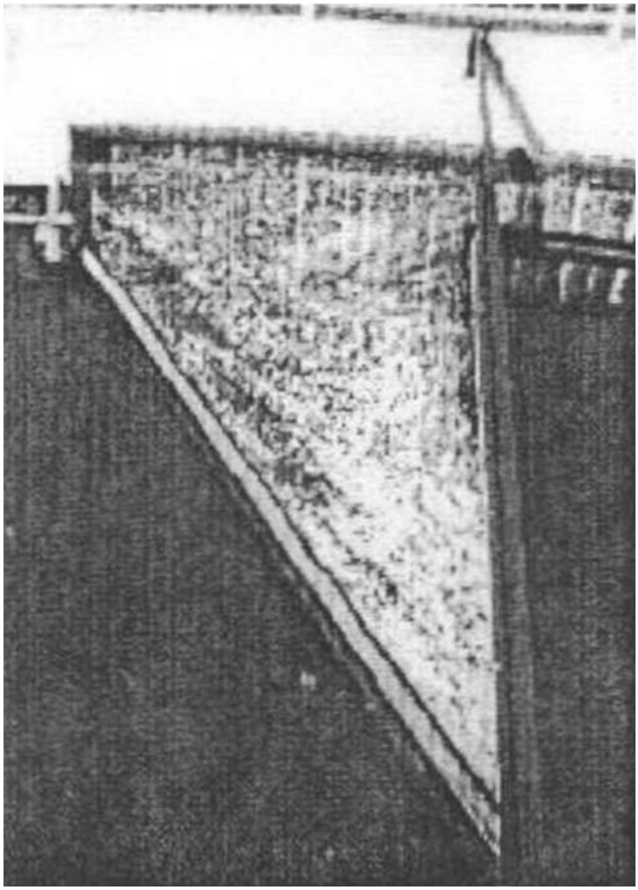

**FIGURE 8.38** Antirotation baffle working with the skimmer arm to "scissor push" scum to the tank perimeter.

problem. As discussed earlier in this chapter, the most effective means is to lower the flocculation well to an elevation that is near the water surface and provide adjustment to its elevation. This can be done by making the baffle adjustable or by adding an adjustable plate to its top.

Another skimming concept, known as the "ducking skimmer", is shown in Figure 8.36b; further details are shown in Figure 8.40. In this design, a skimmer board is connected to the sludge removal mechanism through a hinged, counter-weighted assembly. It pushes the floatables toward a fixed, rotating trough that turns into position as the skimmer board approaches and trips a trigger switch. When the board reaches the rotating trough, it ducks under the trough, and its counterweights return it to the surface to continue rotation around the tank. This device has an advantage of offering full-radius scum removal. Separate flushing is generally not required, but some designs feature a deeper cut opening at the inner end of the rotating trough to take on more water, which moves the floatables into the collector box at the other end of the trough.

Some installations have a reported high amount of maintenance associated with the ducking skimmers. Issues have included controls, bearings, actuators, and binding of the rotating trough. As manufacturers have experienced some of these problems, they have made subsequent designs more robust.

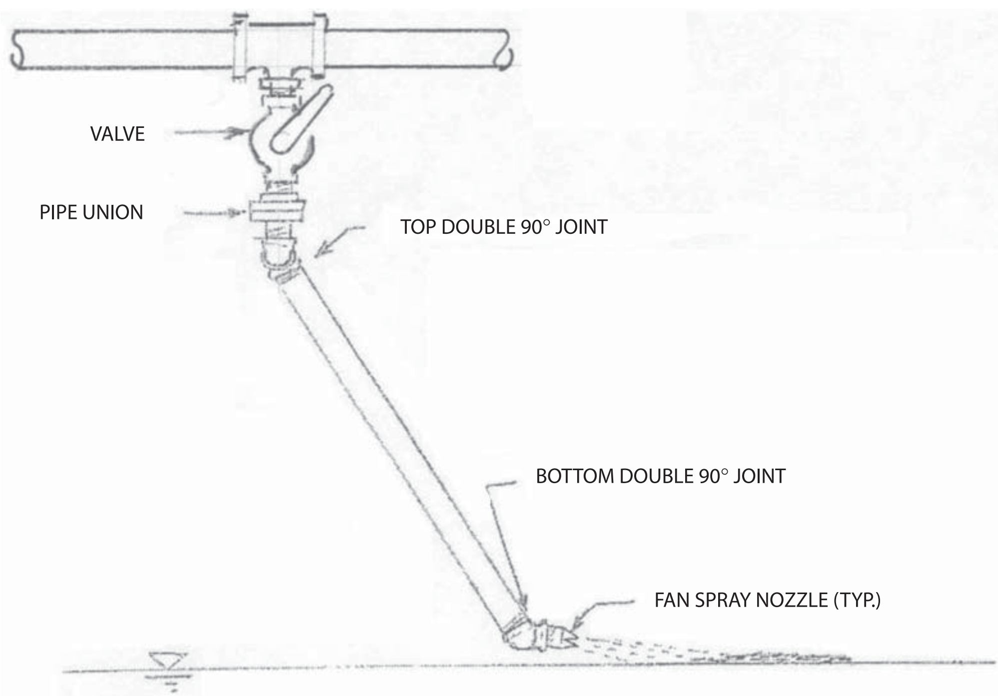


**FIGURE 8.39** Surface spray nozzle arrangement that offers height, vertical spray, and horizontal spray adjustments by using two sets of double 90-deg threaded pipe joints.

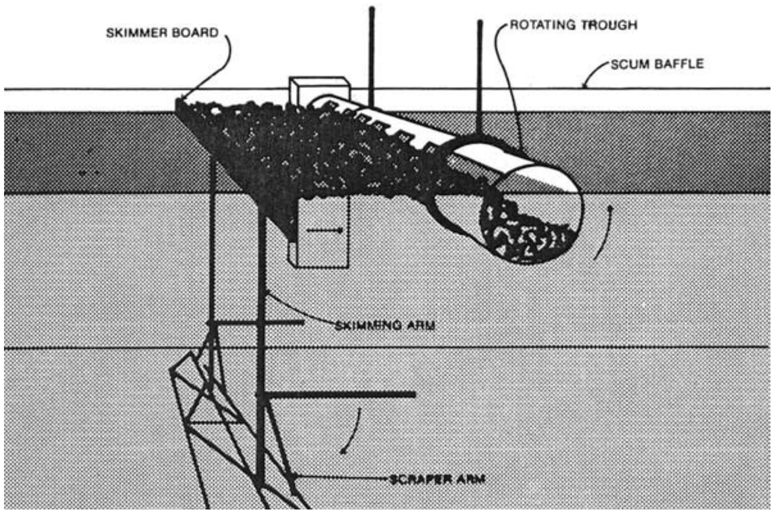


**FIGURE 8.40** Ducking skimmer (also called positive scum skimming device).

A third type of skimmer involves the use of a full radius traveling beach that rotates with the drive cage and discharges into a central, annular well, from which the scum is pumped out. A stationary, hanging flap that has its lower edge just below the water surface bends as needed to push the scum up the beach as it travels below.

Peripheral feed clarifiers must remove foam and floatables from the peripheral feed channel. One design is to feed the tank in a unidirectional manner and locate a small scraper and beach or overflow weir arrangement (described above) at the end of the feed channel. Scum removal is facilitated by having the channel fed in one direction and the cross-sectional area decrease with distance around the tank. This can be achieved by making the channel progressively narrower or decreasing its depth by sloping the floor upward. The latter design allows a fixed-width blade to fit the channel. If the channel becomes increasingly narrow, a narrow fixed, flexible, or hinged skimmer blade arrangement has been used to accommodate the decreasing width. Such a skimming system is shown in Figure 8.41. The weir gate can be carefully adjusted so that scum overflows only at peak flowrates. In other designs, the weir gate is motorized and mechanically lowered as the skimming arm approaches.

There have been incidents in which the feed channel foam problem has become so severe that it overflowed the wall, allowing foam to drop directly into the effluent channel. At Denver Metro, Colorado, this problem led to the conversion of 10 peripheral

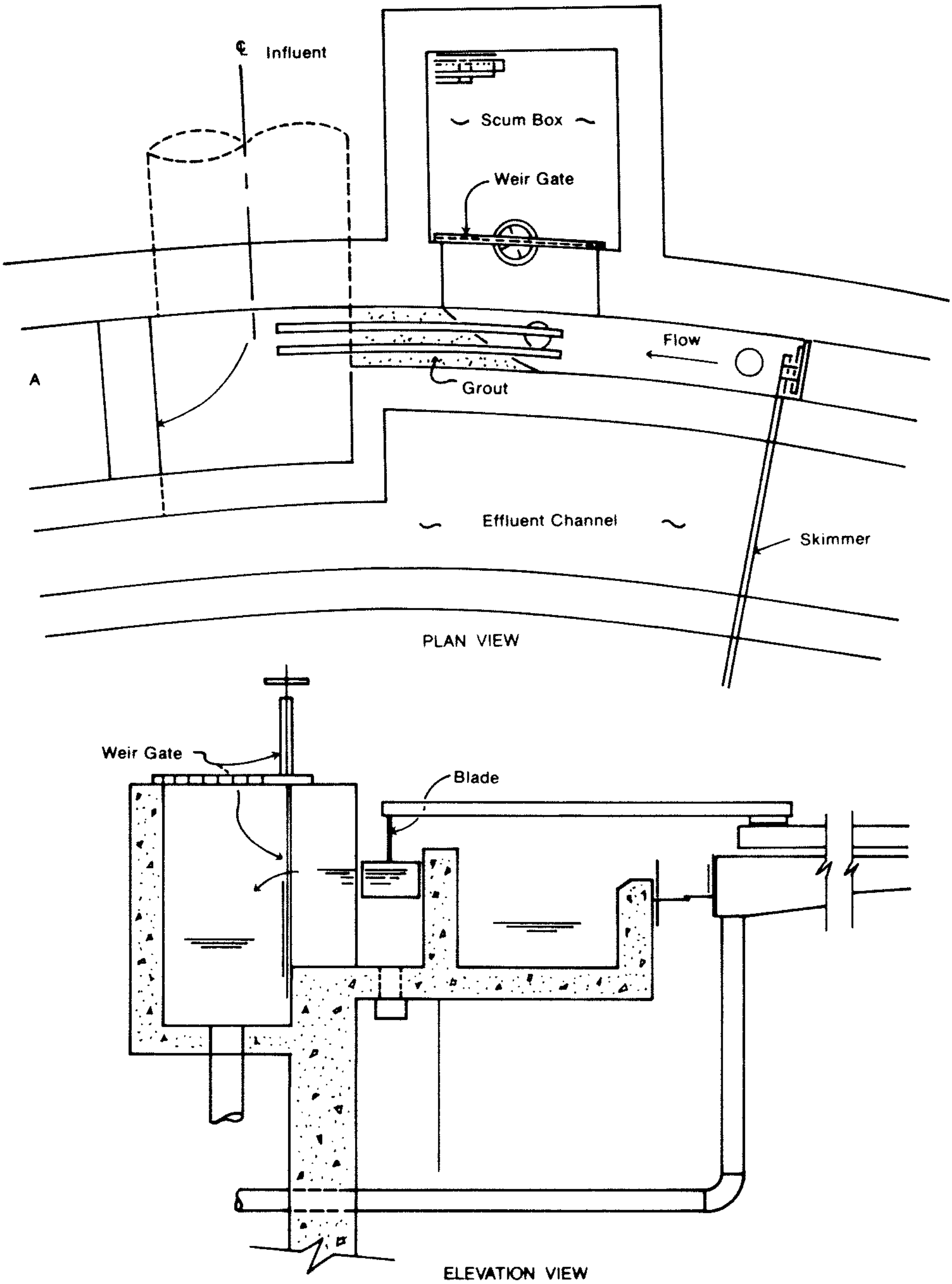


**FIGURE 8.41** Plan and elevation of effective variable width influent channel skimming design for peripheral feed clarifiers.

feed tanks to center feed. There are, however, hundreds of peripheral feed tanks, and most correctly designed units do not have this problem.

The ducking skimmer device offers the advantage of programming the rotating trough. Typically, one and no more than two skimming boards are provided per tank. The trough rotation can be programmed to trip every time the board approaches or can skip some of the cycles to reduce the amount of floatable material removed. This added flexibility is countered by the additional complexity of the system. Some of the models have had binding of the rotating troughs or failure of the motorized device used to rotate the trough. A robust mechanical design and equipment specification is important.

**BLANKET LEVEL DETECTION.** Knowledge of the sludge blanket level in a clarifier has always been essential to good operating practice. The use of a core sampler and other means of measuring this are explained in Chapter 7.

Automatic measurement of the level of sludge in clarifiers has become increasingly popular among operators and useful in automating control of sludge pumping and wasting. Several means are available to do this. An electronic sensor that is based on light transmittance has been used for years. Additional detail on instrumentation is given in Chapter 10.

**ALGAE CONTROL.** Algae growth is a problem with many clarifiers having weirs and open troughs. Many plants do not have provisions to control this growth, and the operators are left to deal with it as a maintenance chore. This involves occasional hose downs and even putting waders on and walking in the troughs to clean them. This is a tedious task and can also be dangerous.

For plants with gaseous chlorine, some designs have provided for a diffuser pipe just upstream of the weir. A periodic release of chlorine solution has served to kill the algae and keep the troughs from excessive growths. In recent years, most plants in the United States have abandoned gaseous chlorine and shifted to hypochlorite or UV light for disinfection. Hypochlorite has a tendency to form a chemical deposit and clog the orifices of a chlorine diffusion line. Thus, chlorine control of algae at clarifiers is declining.

Another method used in some new and retrofitted circular tanks is the addition of spring-loaded brushes to the rotating skimmer arms. Figure 8.42 shows such an installation. This has been successful in many applications but does require periodic adjustment and replacement of the brushes.

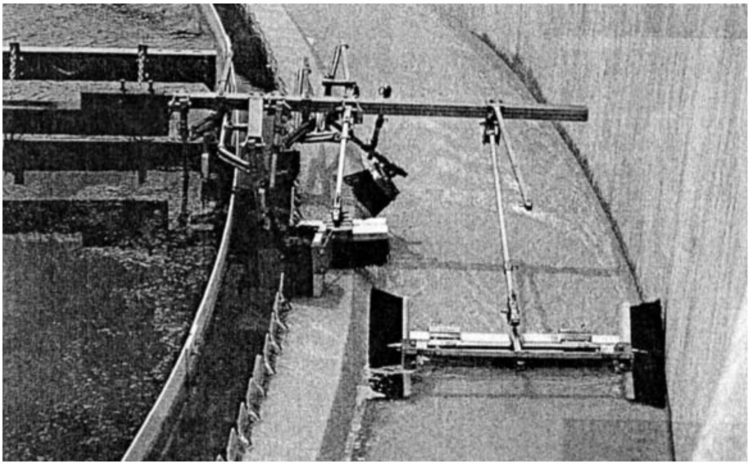

**FIGURE 8.42** Spring-loaded brushes can be strategically placed to keep the effluent trough clean.

Water jet spray systems represent yet another concept to remove algae. In the past several years, an automated arrangement that can be tied into a supervisory control and data acquisition system of a plant has emerged and successfully applied in a few dozen installations. One example is illustrated in Figure 8.43. Supply water is piped to the tank center, where it transfers through a submerged slip ring to a traveling pipe attached to the rotating sludge/scum truss structure. Fixed or rotating branch pipes, with nozzles at their ends, emit an intense spray that can be directed at the walls, weirs, baffles, and troughs. This concept has the ability to clean irregular shapes of tanks and weirs, including the cantilevered double-weir launders.

The concept of adding covers to keep the launder areas dark has been practiced for many years. The covers were not used extensively in circular tanks because of the relatively large area involved and the curved shape.

Commercially available covers made of fiberglass have become more popular in recent years and have been found effective (Figure 8.44). These can be designed to have hinges at the weir or at the wall. Multiple doors can be opened to expose the trough for access and to visibly inspect the quality of effluent. Locating hinges at the wall allows the operators access to the weir and scum baffle, whereas locating them at the weir facilitates observation of the troughs. Covering of the troughs of circular

**FIGURE 8.43** Water spray jet system removing algae from serpentine weirs.

tanks can also be done by constructing a concrete deck over the trough and providing multiple access hatches.

**WALKWAYS AND PLATFORMS.** Most circular clarifiers are equipped with a single walkway that extends from the perimeter of the tank to the center area. For some plants, it is advisable to extend this walkway across the full diameter. In some designs, two full diameter walkways, at 90 degrees to each other, are constructed to facilitate operator traffic, guest tours, sampling, and structural support for the flocculation baffles, skimming devices, and other facilities.

Walkways are typically a minimum of 1 m (3 ft) in width; 1.4 m (4 ft) is preferred by many. A life buoy and a least one hose bib should be provided along the walkway.

The center platforms for clarifiers with center drives are sized to give operators a minimum of 0.7 m (2 ft), and preferably 1 m (3 ft), clearance around the mecha-

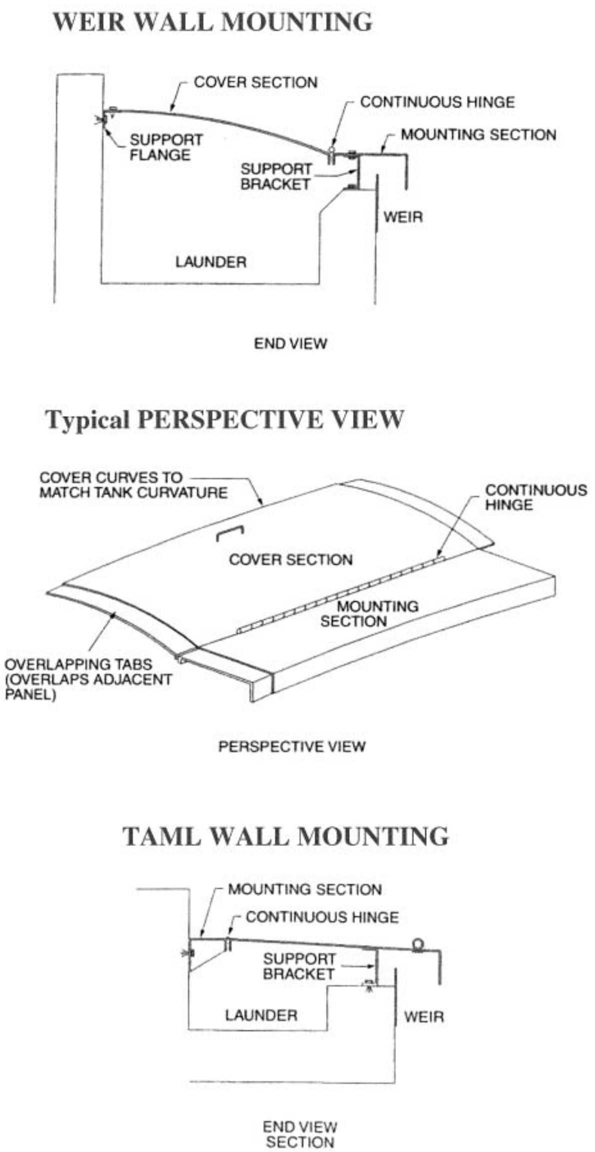


**FIGURE 8.44** Launder covers are available to reduce algae growth by keeping out sunlight.

nism for the operator to work. Platforms that are 2.4 × 2.4 m (8 × 8 ft) or 3 × 3m (10 × 10 ft) often result from this. Access holes and trap doors are often provided at convenient locations to facilitate maintenance, operation, and observation activities of the operators.

**RAILINGS AND SAFETY MEASURES.** ***Railings.*** Federal and state OSHA guidelines require handrails around the clarifier facilities and along walkways to be 1.07 m (3.5 ft) or more in height. In addition, walkways along the bridge require a minimum of three horizontal bars and a kickplate at the bottom. Some designers prefer to use chainlink fencing to reduce cost. This limits the operator's ability to reach through between the bars to take samples and perform maintenance chores.

The outer walls of circular tanks can be designed to serve as safety barriers, protecting people from falling into the tank. In many cases, the backfill grading around the clarifier can be left at 1.07 m (3.5 ft) below the wall. As shown in Figure 8.45, this can eliminate the need for guardrails and facilitate maintenance.

***Lighting.*** Lighting for clarifiers is often tailored to fit the general lighting philosophy of the plant design. A light at the center of the tank to facilitate observation, maintenance, and repairs of the drive is desirable. Lights around the perimeter of the tank to observe and hose down weirs are not considered essential because such operations are generally scheduled for daytime hours. A light over the scum hopper is convenient, but not essential. The philosophy of some plants is to provide only low-level lighting for routine operation and permanent or portable, separate, high-intensity lighting for special or emergency periods.

**DRAINS.** Provisions to drain a clarifier by gravity into a plant drain system are convenient, but not essential. Portable, submersible, or self-priming pumps have been found adequate, especially at small plants where there are few tanks and the frequency of draining one is low.

For clarifiers with scrapers, the sludge line leaving the hopper may serve the dual purpose of providing for tank drainage. If hydraulic suction is used, a mud valve, with a short line connecting to the sludge removal pipe, can enable an operator to drain the last several centimeters of the tank that will not drain through the hydraulic suction device.

For some hydraulic suction clarifiers, the bottom slab can be sloped slightly from the center to the perimeter. A mud valve can then be placed at the wall. A perimeter

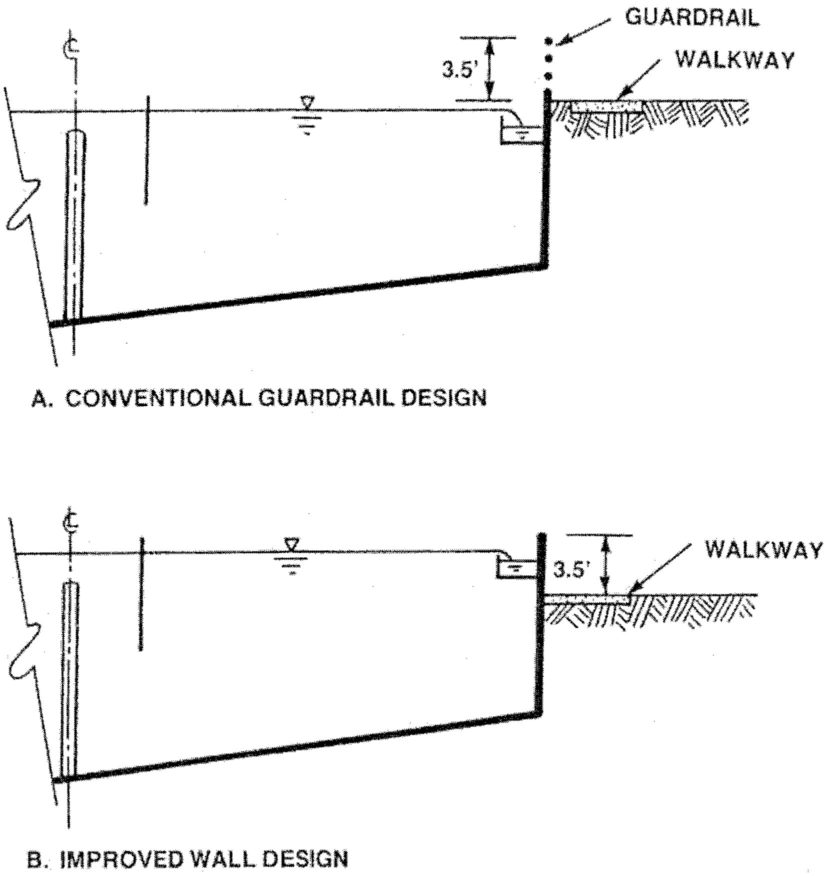


**FIGURE 8.45** Integrating final grading elevations around the tank can eliminate guardrails and give better access to maintenance areas (3.5 ft = 1 m).

gutter should then be constructed to remove the remaining water of the tank to the mud valve location. This design allows a short pipe to take drainage from the valve to a sump or plant drain line.

## EQUIPMENT SELECTION

The most costly item of equipment for a circular clarifier is the rotating mechanism and drives. Once the design drawings are complete, the specifications to get a complete, workable system are extremely important. The experience record and level of satisfaction with installed equipment is important in listing acceptable manufacturers to bid the design. Specifications for circular tank equipment are often categorized according to functional performance, structural loading of the equipment, mechanical design of the components, electric motors, controls and alarms, materials, and coatings for corrosion protection.

**DRIVES.** The drive units for circular tanks typically consist of three sets of reducers that transition speeds from the motor to the rotating mechanism. Worm gears, cycloidal speed reducers, and cogged gears have been used by the different manufacturers. Bearings are extremely important components of the drive mechanisms. The principal types include one in which steel balls run on hardened strip liners set in cast iron, and the second involve forged steel raceways. The latter are commonly called *precision drives*.

In the United States, there are two common forms of clarifier drives: bridge supported styles and center pier (column) supported styles. Bridge supported drives are used in full span bridge clarifiers, typically less than 15 m (50 ft) in diameter. The access bridge supports the center drive. The output flange of the drive attaches to the rotating torque tube (drive shaft), which rotates the collector mechanism (Figure 8.46).

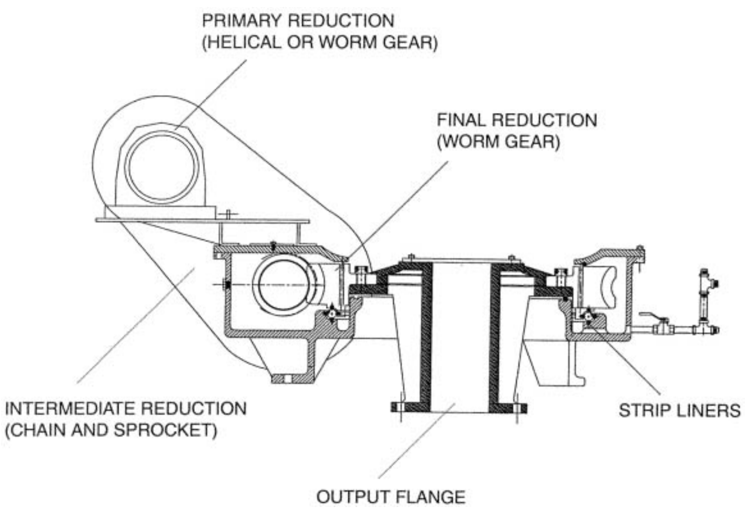


**FIGURE 8.46** Bridge supported style worm gear drive, with replacement strip liners.

Typical drive configurations include a primary and final gear reduction unit (Table 8.2). Selection of the drive size and operating torque and rotational speed are dependent on the application.

Center pier supported styles are used on half span bridge, center column support clarifiers, typically larger than 15 m (50 ft) in diameter. The center column supports the center drive, and the rotating spur gear attaches to the drive cage, which rotates the collector mechanism (Figures 8.47 and 8.48). Typical drive configurations include a primary, intermediate, and final gear reduction unit (Table 8.2). Selection of the drive size, operating torque, and rotational speed is dependent on the application.

Another type of drive used more commonly in Europe is the rim-drive mechanism. It features a motor, gear-box, and drive wheel that runs on the top of a circular tank wall. There are a few units operating in the United States; however, because of the small number, it is not discussed further in this text.

The loading for drives is important to properly size them and the structural members of the collector arms and/or trusses. The load applied to the rotating rake arm is the continuous operating torque, or running torque. This value must be derived from data relative to the actual sludge being removed or derived from sludge of similar characteristics.

Calculation of torque for a circular drive unit is based on the simple cantilever beam-type of equation, with a uniform load ($W$) applied. Torque required to turn a rake arm with radius, $r$, would equal the resultant force of the uniform load ($W \times r$) multiplied by the moment arm ($r/2$). Because most circular clarifiers have two arms, the resulting equation will be as follows:

$$T = (2)\,(Wr)\,\frac{(r)}{2} = Wr^2 \qquad (8.4)$$

Where

$W$ = Units of force, and
$r$ = Units of length.

If only one arm is considered, the previous calculations would be divided by two.

Torque value specified should be tested in the field by means of tiedown tests.

A summary of some uniform loading criteria is presented in Table 8.3. For precise calculations of loadings, variables such as material density, sludge depth, and repose angle of the solids also need to be considered.

**Table 8.2** Clarifier drive comparison.

| Common drive combinations | Primary reduction | Intermediate reduction | Final reduction gear | Final reduction gear—advantages and disadvantages |
|---|---|---|---|---|
| A – Worm gear with replaceable strip liners and ball bearings (bridge-supported style) | 1. Helical gear motor<br>2. Motor and worm gear reducer | 1. Sprocket and chain<br>2. Variable pitch (Reeves type) unit | Worm gear<br>• Hardened alloy worm<br>• Cast iron or cast bronze worm gear<br>• Cast iron housing<br>• Replaceable ball, race, strip liners, and seals<br>• Oil lubricated | • Final reduction worm gear is the least efficient of all combinations for power transmissions.<br>• Worm gear provides steady speed output when subjected to shock loading.<br>• Bearings with strip liners generally have shorter bearing life because of their lower contact angle and lower Rockwell hardness values of the strip liner.<br>• Replacement of strip liner is difficult to perform in field and expensive.<br>• Generally strip liner replacement is being done in the shop.<br>• Cast iron gear housing is more rigid than fabricated steel. There is less problem of warping as a result of welding as in the case of fabricated steel housing. Cast iron is also less damaged because of corrosion. |
| B – Spur gear with replaceable strip liners and ball bearings (pier-supported style) | 1. Helical gear motor<br>2. Motor and worm gear reducer<br>3. Hydraulic system | 1. Sprocket and chain<br>2. Variable pitch (Reeves type) unit<br>3. Cycloidal speed reducer<br>4. Worm gear<br>5. Helical gear<br>6. Planetary gear | Spur gear<br>• Internal pinion<br>• Ring gear<br>• Ductile iron, cast iron, or steel housing<br>• Replaceable ball, race, strip liners, and seals<br>• Oil lubricated | • Final reduction spur gear is the most efficient of all combinations for power transmission.<br>• Spur gear provides unsteady speed output when subjected to shock loading.<br>• Bearings with strip liners generally have shorter bearing life because of their lower contact angle and lower Rockwell hardness values of the strip liner. Replacement of strip liner is difficult to perform in field and expensive. |

| | | | | |
|---|---|---|---|---|
| | | | | • Generally strip liner replacement is done in the shop.<br>• Cast iron gear housing is more rigid than fabricated steel. There is less problem of warping as a result of welding as in the case of fabricated steel housing. Cast iron is also less damaged because of corrosion. |
| C – Spur gear with precision main bearing (pier-supported style) | 1. Helical gear motor<br>2. Motor and worm gear reducer<br>3. Hydraulic system | 1. Sprocket and chain<br>2. Variable pitch (Reeves type) unit<br>3. Cycloidal speed reducer<br>4. Worm gear<br>5. Helical gear<br>6. Planetary gear | Spur gear<br>• Alloy steel internal pinion<br>• Alloy steel ring gear<br>• Fabricated steel housing<br>• Forged alloy steel precision four point contact main bearing<br>• Oil lubricated | • Final reduction spur gear is the most efficient of all combinations for power transmission<br>• Alloy steel precision bearing capable of overturning load capacity<br>• High load bearing capacities<br>• Longer bearing life<br>• Less bearing maintenance<br>• Spur gear provides unsteady speed output when subjected to shock.<br>• Gear tolerances of precision bearings are less accurate because of induction-hardened heat treatment after machining.<br>• Gear machinery quality of precision bearing is lower because of induction-hardened heat treatment after machining.<br>• Nylon spaces in precision bearings increase loading per ball.<br>• Precision bearing cannot be refurbished in the field.<br>• Precision bearings face a possible problem due to the fabricated housing if the plate thickness is too small. |

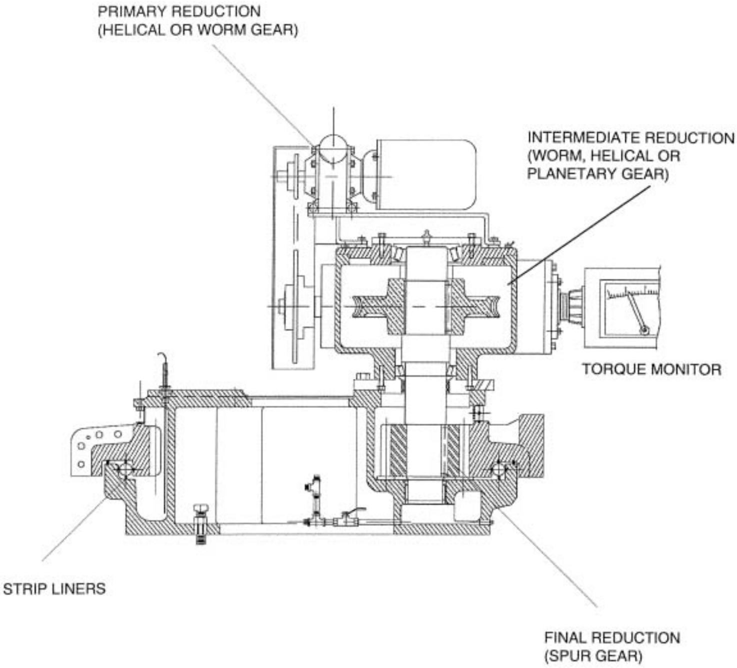


**FIGURE 8.47** Pier supported style cast iron drive, with replaceable strip liners.

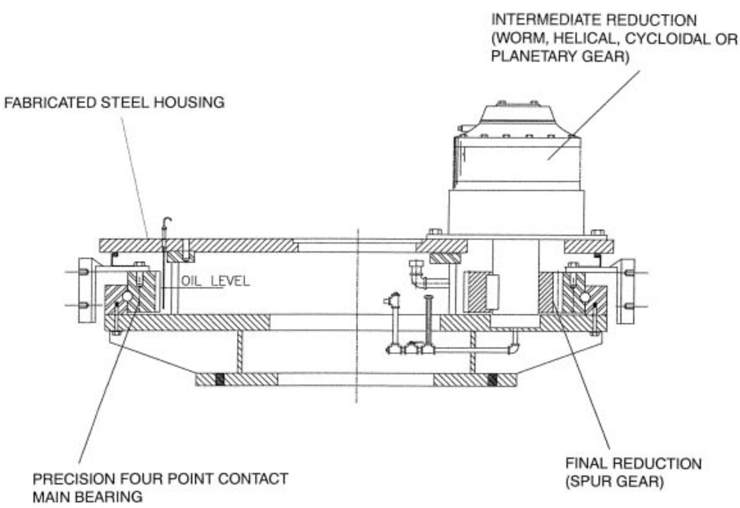


**FIGURE 8.48** Pier supported style fabricated steel drive, with precision bearing.

Structural design should be based on torque values that are at least twice the running torque. The arms are generally of a steel-truss design with rake blades attached to the underside to sweep the sludge toward the center well. The rake blades are often fitted with adjustable brass squeegees for clearance adjustment with the floor.

Some manufacturers recommend a rotation speed limit of 1 to 2 revolutions per hour. Others recommend tip velocities that are converted to rational speed. Typical tip velocities for circular units are shown in Table 8.3

Specifications for clarifier drives should also include reference to the American Gear Manufacturers Association (Alexandria, Virginia). This organization publishes standards that should be followed by gear manufacturers.

**MATERIALS OF CONSTRUCTION.** The primary material for construction of circular clarifier mechanisms has been coated carbon steel. A common coating specified in the United States is coal tar epoxy. When properly applied, this coating provides excellent corrosion protection. Nevertheless, some pinholes in application or scratches during construction occur. Further protection with this coating can be achieved by the installation of galvanic or impressed current cathodic protection systems. These are actually used in a very small percentage of clarifiers, but have been used for industrial applications or waters with high salinity or low pH values.

Galvanized steel can also be used. This has greater application in Europe than in the United States. A disadvantage of this coating is that it can be scratched or damaged in transport and in installation. However, if adequately protected and installed, it does offer good resistance to corrosion.

The use of stainless steel has increased in recent years to provide additional corrosion protection for circular clarifier mechanisms. Either stainless steel 304 or 316 can be used for this purpose. The latter is more expensive but offers better corrosion protection in some installations.

Fiberglass and plastics have also increased in popularity as a construction material for circular tanks. Fiberglass is commonly used for weir plates and scum baffles. It is also used widely now for construction of flocculation baffles, but rarely used for construction of EDIs. The walkways and center platforms of clarifiers are sometimes made of fiberglass. Aluminum is also used in this area.

## TRENDS AND PROBLEMS

Some results from a 1984 survey (Tekippe, 1984) of 20 of the largest environmental engineering firms designing activated sludge clarifiers in the United States are

**Table 8.3** Circular collector drives load selection data.

| Application | Arm loading factor N/m | Arm loading factor lb/ft | Tank bottom slope[a] | Allowable tip velocity mm/s | Allowable tip velocity ft/min |
|---|---|---|---|---|---|
| Primary sludge | 120 to 150[b] | 8 to 10[b] | 1:12 | 50 | 10 |
| Secondary sludge, suction type | 14 to 100 | 3.6 to 7 | 1:50 (drain slope only) 1:12 | 30 to 40 | 6 to 8 |
| Secondary sludge, plow type | 90 to 100 | 6 to 7 | 1:12 | 30 to 40[c] | 6 to 8[c] |
| Alum floc (low turbidity) | 90 to 100 | 6 to 7 | 1:12 | 30 to 40 | 6 to 8 |
| Lime softening | 175 to 220 | 12 to 15 | 1.5:12 | 50 | 10 |
| Lime sludge rethickening | 580 | 40 | 3:12 | 50 | 10 |
| Raw plus secondary thickeners (wastewater) | 440 to 580 | 30 to 40 | 2.75:12 | 50 | 10 |
| Elutriation tanks | 730 | 50 | 3:12 | 30 to 40 | 6 to 8 |
| Blast furnace flue dust | 440 to 1020 | 30 to 70 | 3:12 | 30 to 40 | 6 to 8 |
| Heat treatment sludge | 1170 | 80 | 3:12 | 30 to 40 | 6 to 8 |
| Oxygen furnace dust | 1020 | 70 | 3:12 | 30 to 40 | 6 to 8 |

[a]Where 2:12 or greater slope is indicated, for tanks above approximately 25-m (85-ft) diameter, it is common to use a steep slope of approximately 18 m (60 ft) in diameter and to reduce the slope to 1:12 for the outer area. Exceptions are for blast furnace or oxygen furnace dust, where the outer area slope should be 1.5:12 minimum. This expedites sludge movement in an area likely to classify and permits carrying adequate sludge depth without carrying a big inventory of sludge over the whole tank bottom.

[b]Values will be smaller without grit.

[c]Values up to 150 mm/s (30 ft/min) have been used successfully in some installations with spiral collectors. Variable or multiple speed drives should be used in such cases so that speeds can be reduced in accordance with field observations and measurements. Note: This footnote has been added and was not included in the source reference (WPCF, 1985).

TABLE 8.4 Results from 1984 survey of twenty major United States consulting engineering firms.

| | | | | | | |
|---|---|---|---|---|---|---|
| **Feed** | Center (Simple) 11 | Center (Flocculation) 2 | Peripheral (Envirex) 6 | Peripheral (Lakeside) 1 | | |
| **Sludge** | Plows 4 | Manifold 6 | Riser pipes 10 | | | |
| **Skimmer** | Standard beach 18 | | Ducking skimmer 2 | | | |
| **Weir** | Single peripheral 14 | | Double launder 6 | | | |
| **Depth, m (ft)** | 3.7 (12) 7 | 4 (13) 0 | 4.3 (14) 3 | 4.6 (15) 4 | 4.9 (16) 5 | ≥5.2 (17) 1 |
| **Flow split** | Weir box 14 | Before aeration tank 3 | | Symmetry 1 | Gates 2 | |
| **Blanket level measurement** | Automatic 6 | | Manual 14 | | | |

shown in Table 8.4. A similar survey conducted by WEF® in 2004 is summarized in Table 8.5. The latter includes criteria used by one firm for plants in Canada and the United States. It is interesting to review the most common practices for United States plants in 1984 and compare them to present practices that have evolved over the last two decades.

**FEED.** In the early 1980s, most large consultants were providing conventional center feed inlets. This trend has clearly shifted to provision for flocculating center wells with EDIs for activated sludge clarifiers.

**SLUDGE REMOVAL.** Two decades ago, most firms were using hydraulic suction, using riser pipes for activated sludge. In recent discussions with equipment suppliers, the market share has clearly shifted away from this technology. As per Table 8.5, most new activated sludge clarifiers are now equipped with manifold hydraulic

**TABLE 8.5** Results from 2004 survey of major United States consulting engineering firms.

| Questions | Firm 1 | 2 | 3 | 4 | 5 | 2C* | UK* |
|---|---|---|---|---|---|---|---|
| Q. Tank depth for the diameter shown? (write in three numbers) | | | | | | | |
| A. 23 m (75 ft) | 13 | 14 | 14 | 15 | 14 to 16 | 12 | 6.5 |
| 30.4 m (100 ft) | 14 | 16 | 16 | 16 | 14 to 16 | 12 | 8.2 |
| 45.7 m (150 ft) | 15 | 18 | 20 | 16 to 17 | 14 to 16 | 12 | 8.2 |
| Q. Energy dissipating inlet for center feed tanks? | | | | | | | |
| A. Simple, small, single, cylindrical baffle with no flocculation baffle | | | | | | | |
| EDI tub with adjustable gates and flocculation baffle | X | | X | | | X | |
| EDI tub with fixed outlet chutes (like the clarifier optimization package) and flocculation baffle | | X | | X | X | | |
| Other (describe) | | B | | | | | |
| Q. Peripheral baffle for density current deflection? | | | | | | | |
| A. Crosby type | | | | | X | | |
| McKinney type | X | | X | X | | X | X |
| No baffle, but cantilvered double-weir launder set in from the wall | | X | | | | | |
| None | | | | | | | X |
| Q. Sludge removal mechanism? | | | | | | | |
| A. Spiral scraper | | Also used | C | X | C | | X |
| Mono-tube hydraulic section (Tow-Bro) type | X | X | D | | D | | Also used |
| Organ pipe hydraulic suction | | | | | | X | |
| Multi-blade scraper (like primary clarifiers have) | | | | | | | |

**Table 8.5** Results from 2004 survey of major United States consulting engineering firms *(continued)*.

| Questions | Firm 1 | 2 | 3 | 4 | 5 | 2C* | UK* |
|---|---|---|---|---|---|---|---|
| Q. Floor slope for hydraulic suction tanks? | | | | | | | |
| A. Flat or nearly flat and sloping to the center | X | X | E | | E | | |
| Nearly flat, but sloping to the outside | | | F | | F | X | |
| 1-on-12 slope towards the center | | | | | | | |
| Other (describe) | | | | | | | |
| Q. Sludge hopper for tanks with scraper removal for sludge? | | | | | | | |
| A. Traditional trapezoidal hopper | X | X | | X | | X | X |
| Sludge ring (annular ring with orifices) | | Also used | | | X | | |
| Sludge drum (similar to the sludge ring, but it rotates and picks sludge off at the innermost location of the scrappers or spirals | | | X | | X | | |
| Q. Scum removal devices? | | | | | | | |
| A. Normal beach type | | | | | | X | X |
| Ducking skimmer and rotating trough | | | | X | | | |
| Full radius beach-type skimmer | | X | X | | X | | |
| Other (describe) | A | | | | | | |

A. Wide beach, double arms.

B. Depends on diameter. Up to 20-m simple circular baffle of 30% tank diameter. Over 30 m needs flocculation chamber and horizontal central baffle (or increased depth at center). Over 45 m needs both flocculation chamber and horizontal center baffle and increased center depth, or use flat-bottomed, deep tank with full diameter sludge suction removal.

C. When plant does not have good primary treatment or at least good screening.

D. When plant has good primary treatment, use mono-tube hydraulic suction.

E. For Tow-Bro (mono-tube).

F. For spirals.

*Note: Firm no. 2 reported different values for plants that it designs in Canada (C) and in the United Kingdom (UK). No explanation of why the criteria vary were presented.

suction or spiral plows for activated sludge tanks. For primary and trickling filter clarifiers, plows have given way some to spirals. Hydraulic suction is generally not used for this application.

**SKIMMERS.** The conventional beach skimmer (full or partial radius) still remains the most popular. The ducking skimmer introduced in the early 1980s has been used by some design engineers, especially in plants with large quantities of Nocardia and other activated sludge foams. In recent years, use of the ducking skimmer design has declined somewhat because of its relative cost, concern about mechanical problems with pipe rotation mechanisms, and relatively large amount of water that is taken in with the scum.

**WEIRS.** The distribution of clarifiers using single weirs compared to double launders has probably shifted in favor of single weirs. Of these, the McKinney type is most popular. More regulatory agencies are now relaxing weir loading requirements to facilitate this.

**DEPTH.** The depth for activated sludge clarifiers has probably increased somewhat over the last two decades. Sidewall depths of 4.6 to 6.4 m (15 to 20 ft) are common for the larger tanks. Many engineers, in recent years, follow the WEF® MOP 8 guidelines for depth versus diameter (WEF, 1998). For primary clarifiers, wall depths of 2.7 to 4 m (9 to 12 ft) are quite common. The same would apply for fixed-film clarifiers.

**BLANKET LEVEL.** The degree of automation has continued to increase over the past two decades. More owners are attempting to reduce operator manpower and automate data compilation. Most plants provide clear plastic pipe graduated samplers for operators to use. For plants with automation, higher degrees of computerized data compilation and graphical output are provided to enable an operator to track and record sludge inventory shifting and solids accumulation patterns.

**INTERNAL BAFFLES.** The use of internal baffles to form a flocculation zone and to deflect rising currents at the wall of center feed tanks have become very common in recent years. For the latter, several different designs prevail.

**ALGAE COVERS.** As stated in a previous section, algae covers for secondary and tertiary clarifiers are becoming more popular. The use of chlorine solution has declined considerably because owners have chosen to avoid having gaseous chlorine on site and hypochlorite, often used to replace gaseous chlorine, causes scaling of the solution diffuser orifices. Owners are also becoming more concerned about operator safety while cleaning tanks to remove algae, so devices such as covers and automatic scrubbing are likely to increase.

## CASE STUDIES

There have been numerous case studies in the last several years that add to the body of knowledge regarding circular clarifier design. A few have been selected to share experience gained.

**HYPERION WASTEWATER TREATMENT PLANT (CITY OF LOS ANGELES, CALIFORNIA).** The new wastewater treatment plant at Hyperion has been touted as one of the major engineering accomplishments in recent wastewater treatment history. In the area of circular clarifier design, this project involved sequential replacement of rectangular final clarifiers with circular units. The plant presently contains 36 center feed circular activated sludge clarifiers. Each has a diameter of 45.7 m (150 ft) and a side water depth of 3.73 m (12.25 ft). Additional features are listed in Table 8.6. These clarifiers were initially equipped with large diameter flocculation wells and four vertical turbine mixers each. Upon startup, it was learned that the operation of the mixers did not influence clarifier performance. In the later units, these mixers were deleted.

Initially, the clarifiers had difficulty meeting the peak hydraulic capacity specified. To meet specifications, some side-by-side design modifications were tested. Four tanks in one module were chosen. Three were modified, and the fourth was used as a control. In two tanks, the flocculation baffles were supplemented with two different types of EDIs. The alternatives have been previously illustrated in Figures 8.8 and 8.12. In addition, an extended perimeter baffle was tested in the third tank that was modified. The geometries for the four tanks tested are shown schematically in Figure 8.49.

Comparative dye tracer isopleths after 15 minutes of dispersion time are also shown. The results show that a deeper, faster moving current is established for the control clarifier and the one using the LA EDI, which includes bottom diffusers. Figure 8.50 also shows the dye distribution curves and TSS average concentrations from these four tanks. The curves are similar, with the exception of clarifier 3B. This

**Table 8.6** Case history clarifier geometric features and loadings.

| Parameter | Hyperion, Los Angeles, California | Central treatment plant, Denver Metro Water Reclamation District | Kenosha, Wisconsin |
|---|---|---|---|
| Diameter, m (ft) | 45.7 (150) | 42.7 (140) | 42.7 (140) |
| Depth at wall, m (ft) | 3.7 (12.25) | 4.3 (14) | 4.3 (14.2) |
| Bottom slope | 1 on 12 | 2% | -3% |
| Feed pipe diameter, mm (ft) | 1370 (4.5) | a | a |
| Flocculation baffle diameter, m (ft) | 16.7 (55) | 19.5 (64) | c |
| Flocculation baffle depth, m (ft) | 2.1 (7 ft) (with 0.6 m (2 ft), 26-degree inclined lip) | 2.4 (8) | c |
| EDI diameter, m (ft) | 6.1 (20) | None | c |
| EDI depth, m (ft) | 1.37 (4.5) | None | c |
| Effluent weir and baffle type | McKinney | Trough, inset 2 m (6.5 ft) | Single, peripheral |
| Sludge removal mechanism | Plows | b | Tow-Bro |
| Mechanism drive type | Column | Column | Column |
| Collector tip speed, m/min (ft/min) | | | |
| initial | 2.87 (9.42) | | |
| increased later | 4.57 (15.0) | a | a |
| Surface overflow rates for test, m/h (gpd/sq ft) | 1.45 to 2.89 (850 to 1700) | 1.40 (825) | 1.36 to 1.87 (800 to 1100) |
| RAS (% of average flow) | 20 to 30 | a | 24 to 37 |
| SVI (mL/g) | 43 to 167 | 63 to 106 SSVI | 72 to 83 SSVI |
| MLSS (mg/L) | 920 to 1270 | 2 000 to 2 780 | 1880 to 2330 |

[a]Unknown

[b]Two tanks were tested. One had multiple plows and the other had organ-pipe, hydraulic suction.

[c]Peripheral feed

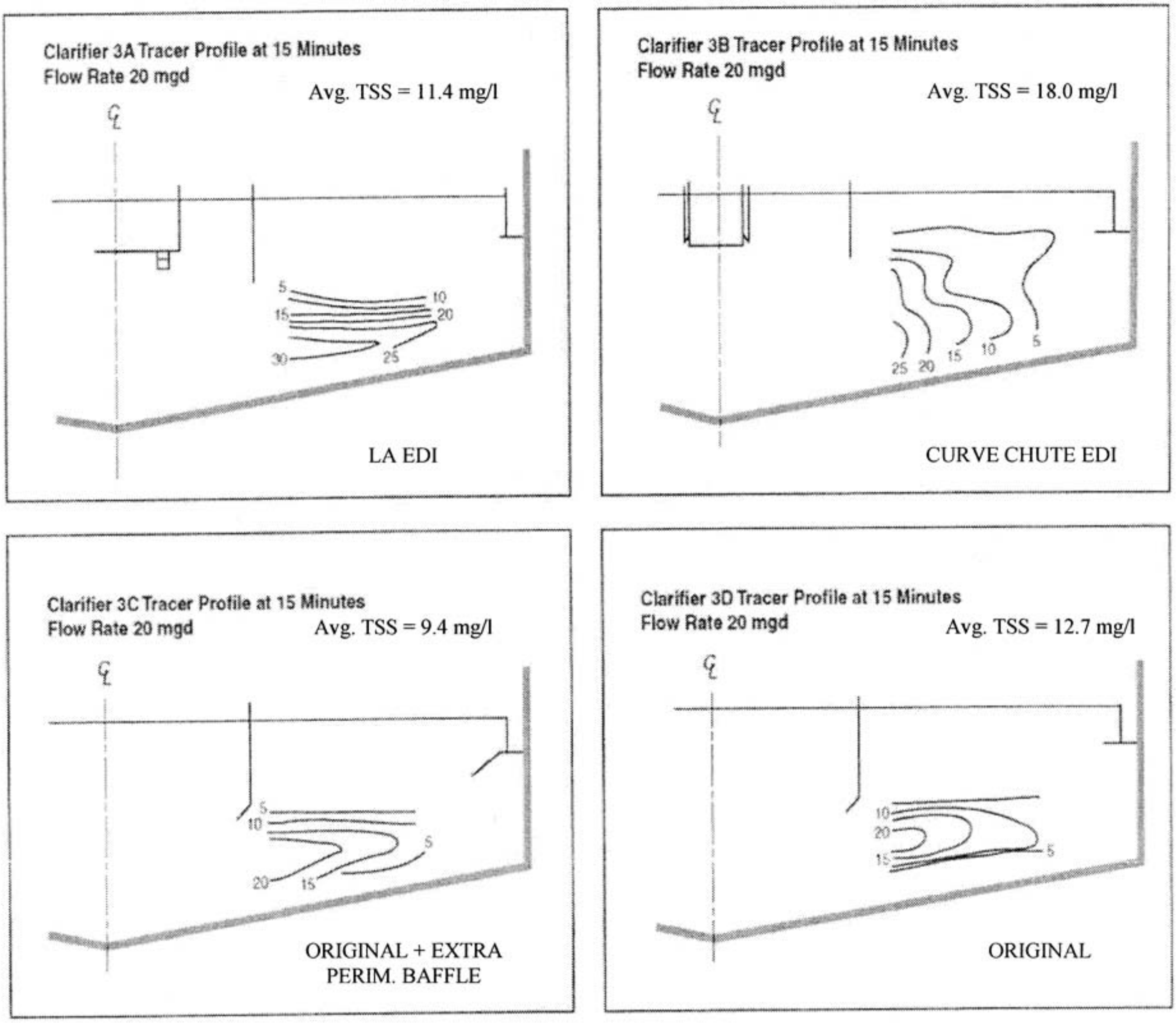


**FIGURE 8.49** Dye tracer movement in four test tanks at Hyperion wastewater treatment plant, Los Angeles, California.

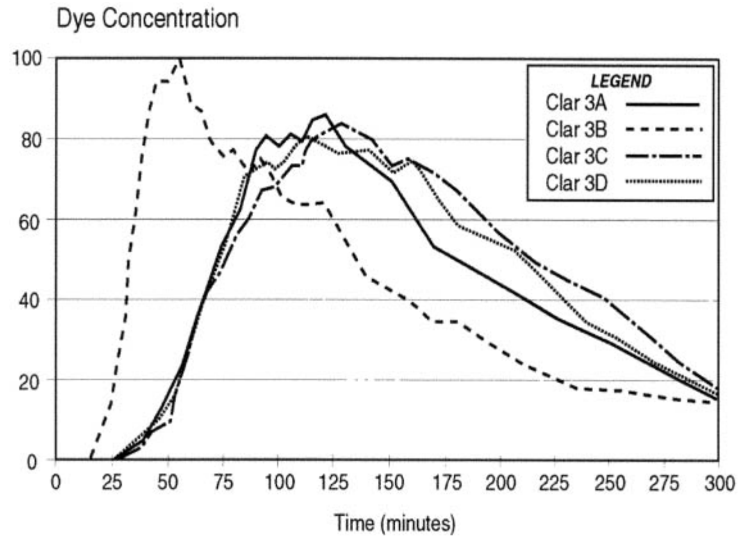


**FIGURE 8.50** Effluent dye concentration curves for different inlet geometries at Hyperion wastewater treatment plant, Los Angeles, California. Flowrate = 0.55 $m^3/s$ (12.5 mgd) per tank.

unit was the tank featuring an EDI with curved chutes. It shows a substantially earlier breakthrough of dye. Field performance results also show that this clarifier experienced severe blanket disruption and hydraulic turbulence at lower flowrates than the other three. Drogue tests were also performed on these tanks. Figure 8.51 shows results comparing the two tanks equipped with EDIs. The one with curved chutes shows substantially higher rotation velocity vectors.

Solids profiles at high loadings were also compared. Figure 8.52 shows solids profiles at 0.9 $m^3/s$ (20 mgd) per tank. The data show that the tank with an EDI featuring the bottom diffusers was able to maintain a shallow blanket, while the blanket rose substantially and resulted in effluent solids lost for the alternative EDI.

As a result of these and others tests, the city decided to modify the inlet design on all 36 tanks. Specifically, the large flocculation baffle was shortened somewhat, and new EDIs with bottom diffusers were constructed at all tanks. The plant was able to increase its hydraulic capacity by a factor of approximately two.

**DENVER METRO, COLORADO.** The Denver Metro Wastewater Treatment Plant, Colorado, has two large batteries of secondary facilities. The south plant consists of pure

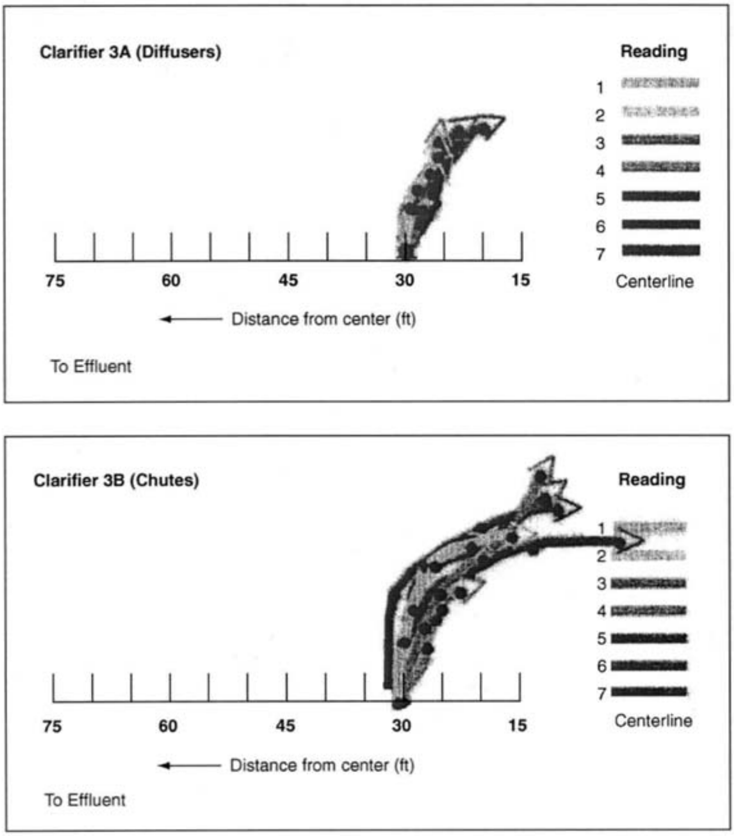


**FIGURE 8.51** Drogue movements resulting from two different EDI designs.

oxygen reactor followed by 10 circular secondary settling tanks. These tanks, originally constructed in the late 1970s, consisted of peripheral feed with ports at the bottom of the feed channels. Hydraulic suction using riser pipes was used for sludge removal. After over a decade of operation, the owner made a decision to increase the plant capacity to compensate for capacity lost resulting from nitrification requirements at the north plant. The secondary clarifier performance at the south plant was limited somewhat by the hydraulics and foam management capacity of the peripheral feed tanks. Therefore, these tanks were converted to center feed by construction of a large feed pipe and new center foundation. The internal mechanisms were replaced. The vertical wall between the inlet and outlet of the peripheral feed design was partially removed and the orifices were

Clarifier 3A (Diffusers)

| Depth | 12' | 24' | 30' | 50' | 67' |
|---|---|---|---|---|---|
| -1' | 180 | 150 | 60 | 80 | 110 |
| -2' | 220 | 170 | 60 | 80 | 120 |
| -3' | 240 | 170 | 60 | 110 | 140 |
| -4' | 320 | 190 | 140 | 130 | 140 |
| -5' | 650 | 210 | 150 | 150 | 140 |
| -6' | 590 | 220 | 160 | 160 | 150 |
| -7' | 640 | 230 | 160 | 160 | 150 |
| -8' | 610 | 240 | 170 | 170 | 150 |
| -9' | 460 | 350 | 170 | 170 | 150 |
| -10' | 510 | 370 | 260 | 180 | 190 |
| -11' | 610 | 380 | 410 | 300 | 170 |
| -12' | 580 | 580 | 400 | 480 | 200 |
| -13' | 570 | 500 | 490 | 530 | 7100 |
| -14' | 780 | 610 | 560 | 5800 | |
| -15' | 730 | 620 | 780 | | |
| -16' | 1060 | 4640 | 7790 | | |
| -17' | 8750 | 7410 | | | |
| -17.5' | 8700 | | | | |

Clarifier 3B (Chutes)

| Depth | 12' | 24' | 30' | 50' | 67' |
|---|---|---|---|---|---|
| -1' | 1460 | 1400 | 80 | 90 | 120 |
| -2' | 1300 | 1460 | 100 | 100 | 130 |
| -3' | 1430 | 1510 | 120 | 110 | 130 |
| -4' | 1450 | 1430 | 120 | 120 | 140 |
| -5' | 1500 | 1440 | 140 | 130 | 1170 |
| -6' | 1570 | 1480 | 250 | 1150 | 1150 |
| -7' | 1600 | 1340 | 1180 | 1360 | 1380 |
| -8' | 1590 | 1490 | 1340 | 1160 | 1580 |
| -9' | 2200 | 1370 | 1300 | 1500 | 1400 |
| -10' | 2020 | 1380 | 1410 | 1580 | 1450 |
| -11' | 2010 | 1370 | 1600 | 1300 | 1500 |
| -12' | 1740 | 1430 | 1450 | 1400 | 1660 |
| -13' | 1910 | 1410 | 1800 | 1400 | 8000 |
| -14' | 1660 | 1360 | 1730 | 6700 | |
| -15' | 1600 | 1820 | 1760 | | |
| -16' | 1560 | 1780 | 3870 | | |
| -17' | 2830 | 7710 | | | |
| -17.5' | 6400 | | | | |

**FIGURE 8.52** Solids profiles at high loading rates. Flowrate = 0.88 $m^3/s$ (20 mgd) per tank (1 ft = 0.304 8 m).

closed off. This enlarged the hydraulic capacity of the effluent trough. The clarifiers were then returned to service and their capacities were increased. Design data are given in Table 8.6.

An evaluation was also made relative to the sludge removal mechanisms. One plant was converted to multiple blade plows. It was then run side-by-side with another plant using the riser pipe hydraulic suction design. A clarifier research technical committee (CRTC) study was conducted onsite to compare performance of these units. It was concluded that hydraulic suction was more successful in maintaining a shallow sludge blanket. Because these tanks are considered relatively shallow by today's standards, blanket depth was judged to be important. The riser pipe option was therefore considered to be superior and the new mechanisms provided in all 10 tanks consisted of this design.

**KENOSHA, WISCONSIN.** The wastewater treatment plant serving Kenosha, Wisconsin, has four circular peripheral feed tanks with Tow-Bro hydraulic suction mechanisms for sludge removal. It is a conventional plug-flow activated sludge plant with fine-bubble diffused aeration. The plant was designed to treat an average flow of 1.23 $m^3/s$ (28 mgd) and a peak flow of 3.0 $m^3/s$ (68 mgd). The latter results in a surface overflow rate of 1.88 m/h (1104 gpd/sq ft).

To determine how the tank would perform under these peak conditions and to learn more about the tank hydraulics, a field-testing program was conducted in August 2000. The features of the tanks and the loading ranges tested are presented, in brief, in Table 8.6.

Tests were run at overflow rates of 1.36 to 1.88 m/h (800 to 1100 gpd/sq ft), producing solids loading rates of 88 and 127 $kg/m^2 \cdot d$ (18 and 26 lb/sq ft/day), respectively. Flow curve/detention time tests were run, and the data of Figure 8.53 were developed. The flow curve showed a slower tendency of the dye curve to taper off after the peak, compared to center-feed tanks (Esler, 2000). This was attributed to reflect less short-circuiting of flow.

The hydraulic movement within the tank was also studied by using an intensive dye tracer test, as described in WERF (2001). The results of profiles taken for the 1.88 m/h (1100 gpd/sq ft) loading are shown in Figure 8.54. The report authors described the current patterns as follows:

- The current initially developed as a plume, at approximately 2.4 m (8 ft) below the surface;
- The current continued as a concentrated plume toward the center at a rate of approximately 1.2 m/min (4 ft/min);.
- The current continued to propagate as a distinct plume to the center of the clarifier; and
- There was a slow expansion of the plume upward throughout the surface of the clarifier.

An interesting phenomenon about these tanks is that, as flows increase, they are driven further into the center of the tank and create a longer travel distance. The net effect is that there is a stabilizing effect at higher flows. A similar conclusion was reached by Crosby (1980), while testing peripheral-feed tanks at East Bay Metropolitan Utilities District.

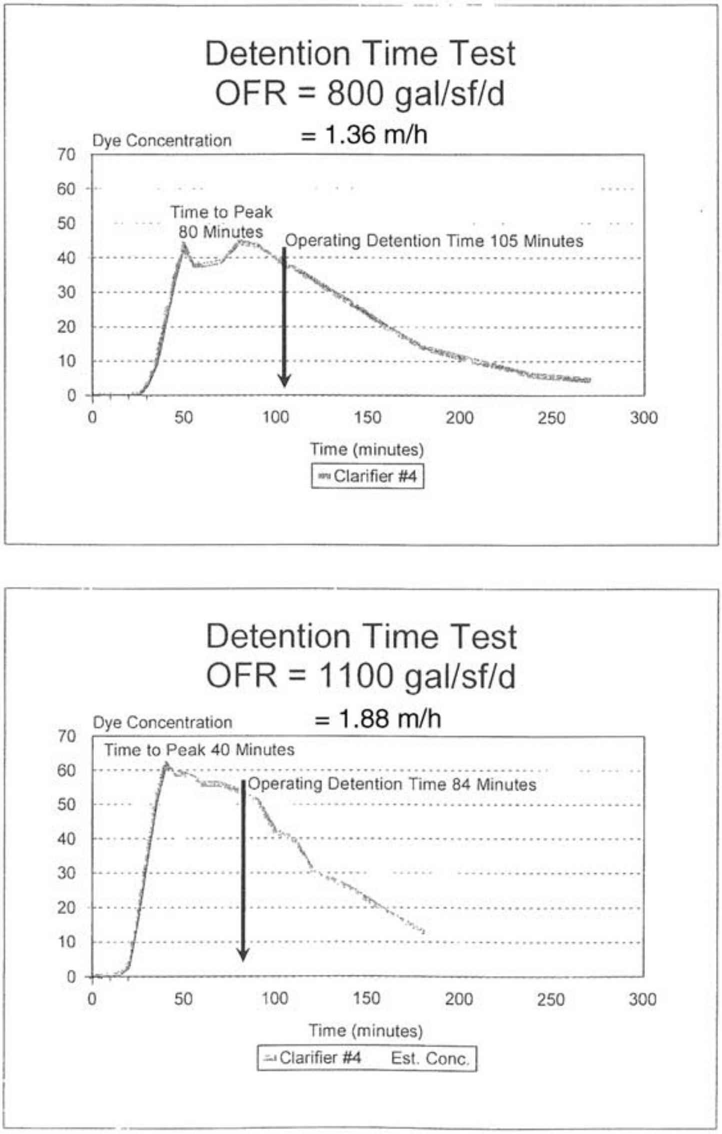


**FIGURE 8.53** Kenosha, Wisconsin, water pollution control plant secondary clarifier dye test results.

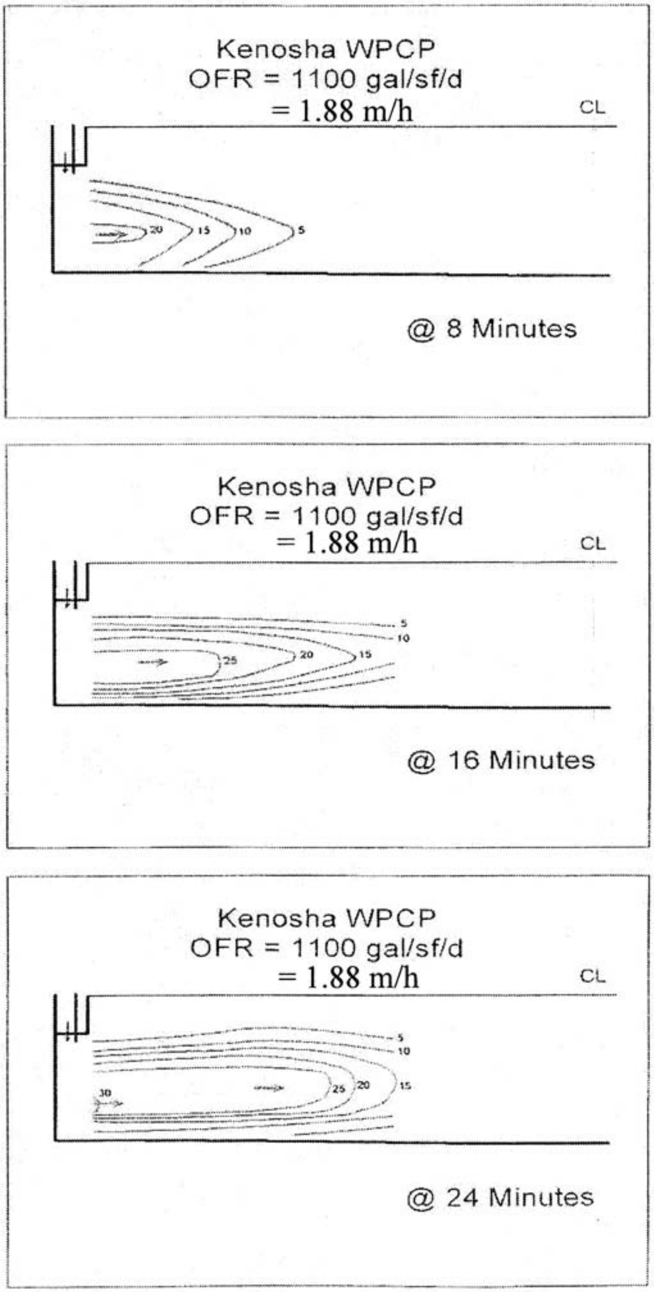


**FIGURE 8.54** Kenosha, Wisconsin, water pollution control plant secondary clarifier.

**TABLE 8.7** Summary of advantages and disadvantages for several clarifier configurations.

| Parameter | Advantages | Disadvantages |
|---|---|---|
| **Sludge collectors** | | |
| Plow scrapers | Low cost.<br>Commonly available.<br>Proven track record of performance.<br>Does not clog with raw wastewater sludge.<br>Capable of handling heavy sludges.<br>Compatible with any scum removal mechanism or feeding system. | Long time requirements to remove sludge from tank (denitrification potential). |
| Spiral scrapers | Low cost.<br>Maintenance from tank surfaces without dewatering the tank for access (some designs, not others).<br>Rapid sludge removal. | Incompatible with some scum removal mechanisms.<br>Sometimes used with circular sludge hopper that costs more. |
| Suction mechanisms | | |
| —Riser pipe type | Flat floor accommodation.<br>Adjustable relative sludge withdrawal rates.<br>Reduces denitrification flotation. | Often requires higher return activated sludge flowrates; therefore, more electrical energy.<br>Narrow tubes clog easily.<br>High cost.<br>Potential for safety problem.<br>More stirring of tank.<br>Difficult to obtain uniform withdrawal of sludge if dense.<br>Requires frequent adjustment of tube flowrate. |
| —Manifold type | Less stirring than organ pipe alternative.<br>Simplicity in design.<br>Rapid sludge removal.<br>Reduces denitrification flotation. | Inability to make field adjustments.<br>Orifice clogging is not easily detected.<br>Inability to make field observations.<br>Leaking of bottom seal could be a problem. |

*(continued on next page)*

**Table 8.7** Summary of advantages and disadvantages for several clarifier configurations *(continued)*.

| Parameter | Advantages | Disadvantages |
|---|---|---|
| Skimmers | | |
| Beach and scraper (conventional narrow types) | Low capital cost.<br>Simple to maintain and clean. | Limited capacity<br>Solids may rotate around tank before removal.<br>Ice buildup and damage in cold climates is possible. |
| Full radius beach and scraper | Positive scum removal.<br>No problem of solids rotating around tank.<br>High removal capacity | High capital cost (especially for large tanks).<br>Incompatible with spiral scrapers that are attached to rim drive bridges. |
| Ducking skimmer scum removal device | High removal capacity.<br>No problem of solids rotating around tank. | Remove large quantities of carrier water.<br>Rotating trough on some units have had binding problems.<br>Incompatible with spiral scrapers that are attached to rim drive bridges. |
| Submersible pump type | Compatible with sludge removal mechanism without skimmer.<br>Low capital cost. | Requires pumping through center bearing.<br>Requires slip-ring electrical contacts.<br>Solids may rotate around rank before removal. |
| **Shape** | | |
| Circular | Minimal construction materials for single unit.<br>Simplest mechanical equipment for solids and scum removal.<br>Low maintenance. | Requires most plant surface area.<br>No common-wall economy of construction.<br>Discrete inlet ports may cause jets.<br>Tanks are generally quite deep. |
| Square | Common-wall economy of construction.<br>Potential for common pipe galleries. | Corner sweep mechanism with high maintenance.<br>More flow in directions of the corners.<br>More construction materials required for single units.<br>Skimming problems. |

*(continued on next page)*

**Table 8.7** Summary of advantages and disadvantages for several clarifier configurations *(continued)*.

| Parameter | Advantages | Disadvantages |
|---|---|---|
| Hexagonal and octagonal | Common-wall economy of construction. | Filleted corners sloping to circular scraper mechanism.<br>More construction material for single units than for circular tanks. |
| **Tank depth** | | |
| Deep tanks (see Table 8.1a) | Improved performance over shallower tanks.<br>Increased sludge concentration. | Higher cost.<br>Potentially more difficult to construct because of dewatering and buoyancy. |
| **Feed point** | | |
| Center feed (simple) | Low cost. | Potentially lower efficiency.<br>Small volume to dissipate inlet energy. |
| Peripheral feed (compared to center feed) | Potential improved efficiency especially at high loadings. | Potential for orifice clogging.<br>Potential for poor inlet distribution if not well designed.<br>Influent channel skimming required.<br>Higher capital cost. |
| Flocculating center feed (compared to center feed) | Improved efficiency. | Higher capital cost. |
| **Outlet weirs** | | |
| Single peripheral weirs | Low capital cost.<br>Full-width skimming.<br>Permits easy hosing down and cleaning.<br>Easily kept level.<br>Cover or brush for algae control. | Potential for wall rebound phenomenon.<br>Short weir length and high loadings in large tanks. |
| Dual or multiweir cantilevered or suspended launders | May improve suspended solids removal somewhat (not always).<br>May help prevent wall rebound solids loss.<br>Lower weir loading rates to meet codes for large tanks. | Higher capital cost.<br>Can prevent full-width scum removal.<br>More difficult to keep level.<br>May not prevent wall rebound effect in some common designs.<br>Potential for oscillations of weir launders.<br>Difficult to hose down algae and keep clean. |

The effluent TSS concentrations at Kenosha average 17 to 19 mg/L at the 1.88 m/h (1100 gpd/sq ft) overflow rate. The sludge removal mechanism was very effective in keeping the blanket thickness to less than 0.6 m (2 ft) during this phase.

In summary, the tests demonstrated that the Kenosha clarifiers were able to perform well at their peak loading rates and produce an effluent with TSS values below 20 mg/L.

## SUMMARY OF ADVANTAGES AND DISADVANTAGES OF VARIOUS CIRCULAR CLARIFIER DESIGN FEATURES

The text of this chapter explains numerous circular tank design features and explains advantages and disadvantages of many. As a convenient point of reference, these are summarized in Table 8.7.

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# Chapter 9

# Rectangular Clarifiers


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## INTRODUCTION

Rectangular clarifiers are used most often in large wastewater treatment plants. Many engineers consider them less costly to construct in multiple units because of common-wall construction and the convenience of providing piping galleries and channels along either the influent or effluent end of the tanks. Design features of rectangular tanks include length, width, and depth, inlet gates, diffusers or ports, flocculation zones, internal baffles, skimmers, sludge flights, scrapers or suction mechanisms, location of sludge hoppers, type and location of weirs or submerged outlets, tank covers, and materials of construction. A large number of design features enable the engineer to develop a wide array of rectangular clarifier designs. A hypothetical rectangular clarifier containing a number of these design features is shown in Figure 9.1.

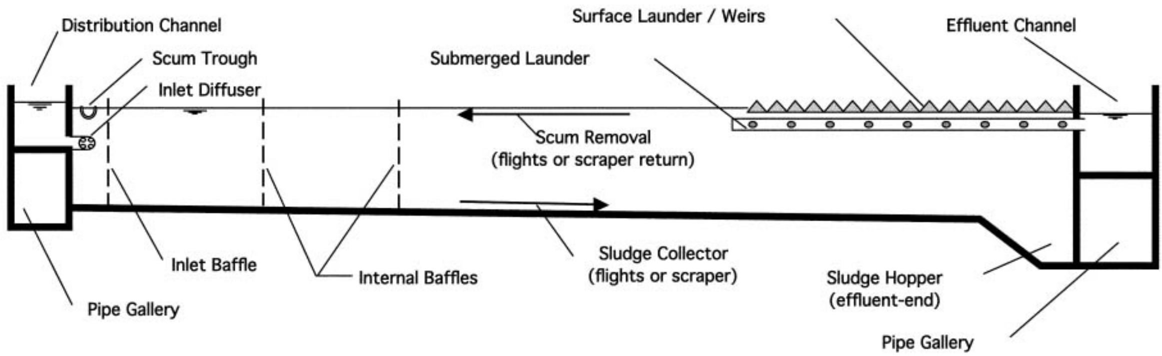


**FIGURE 9.1** Rectangular clarifier design features and nomenclature (note: hopper location may vary). (adapted from Secondary Settling Tanks, ISBN: 1900202035, with permission from the copyright holder, IWA).

Rectangular tanks are used mostly as primary and secondary clarifiers. They can also be used as tertiary clarifiers, although that use is becoming rarer. This is because treatment that was formerly reserved for tertiary systems, such as chemical addition for phosphorus removal and nitrification and denitrification, are being incorporated to primary and secondary treatment schemes. Additionally, the need for highly treated water for reuse and the success of other emerging treatment technologies have produced a trend such that tertiary clarification is being replaced with more advanced tertiary or polishing techniques that do not use clarifiers, such as membrane bioreactors and microfiltration/reverse osmosis systems.

Attention should be drawn to the fact that the design of any clarifier system should be carefully tied with the specific treatment processes that are provided and even the type of wastewater that is being treated. For instance, the peak flow a primary clarifier will experience depends on whether dedicated sanitary sewers or combined sewers are upstream. In addition, the amount of inflow and infiltration during wet and dry periods will have to be considered. Industrial wastewater contributions could have a large effect on the design. In secondary clarifiers, the type of biological treatment process will influence the design and the effluent regulations. Biological nutrient removal (BNR) systems can have specific clarifier requirements. Even the tradeoff between reactor size and sludge concentration will greatly influence a secondary clarifer design. Finally, the type of sludge produced by tertiary plants can vary greatly, depending on the process and chemicals used. In short, the design engineer needs to look at the big picture for the present and future, as emerging regulations can put more strain on operating systems. The reader is referred to Chapter 4 for more information on clarifier design strategies.

## TYPICAL HYDRAULIC FLOW PATTERNS

Rectangular clarifiers can be classified as having either longitudinal, transverse, or vertical flow patterns. For the sake of brevity, this chapter will not cover vertical flow clarifiers in any detail. Instead, this chapter will concentrate on the longitudinal flow type of rectangular clarifiers found in primary and secondary treatment systems and mention other rectangular clarifiers, such as tertiary and transverse clarifiers, where applicable. An important distinction is that longitudinal tanks can either have cocurrent, countercurrent, or crosscurrent sludge removal. When the clarified effluent flow stream reverses itself, it is referred to as a *folded flow pattern*. Rectangular clarifiers placed on top of each other are referred to as *stacked clarifiers*.

**LONGITUDINAL FLOW TANKS.** The simplest and most intuitive hydraulic flow pattern is realized by the longitudinal flow tank design. Longitudinal flow means that the influent flow is introduced at the narrow end of the tank and proceeds in a direction parallel to the long part, or longitudinal axis, of the tank. For a given surface area, this means that the hydraulic loading per unit width is high. The flow pattern of a longitudinal rectangular clarifier closely resembles a plug flow condition, although a sedimentation process is taking place along the vertical axis. Dye studies have shown that the rectangular design better approaches an ideal plug flow regime as compared to circular designs. This can translate into greater hydraulic stability and less variable residence time distribution patterns for both the effluent and the sludge solids (see Figure 9.2).

**COCURRENT, COUNTERCURRENT, AND CROSSCURRENT SLUDGE REMOVAL.** In the simplest longitudinal flow scheme, the clarified liquid flow and the sludge flow proceeds down the length of the tank in the same direction. This describes a cocurrent flow condition. Influent flow enters the narrow end of the clarifier and passes through multiple inlet devices located slightly above the existing blanket. These inlets are designed to distribute the flow across the width of the clarifier and dissipate the inlet energy. The bulk of the mixed liquor solids then separate rather quickly. These sludge solids begin to form a blanket interface, and, because of the difference in specific gravity of the water and the sludge solids, proceed to settle

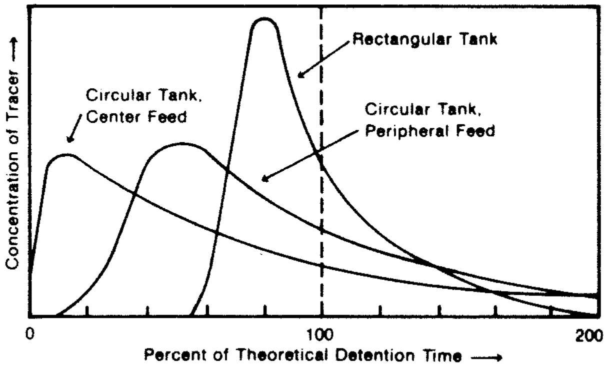


**FIGURE 9.2** Typical tracer study results showing residence time distribution

to the bottom. The settling action of the solids, along with the removal of the sludge flow stream, produces a density current along the bottom of the tank. The density current imparts a momentum such that the sludge is efficiently moved along the length of the clarifier to the sludge hopper on the downstream end. The sludge is generally assisted in its movement down the tank with flights or scrapers. The other part of the liquid not associated with the settled sludge becomes the clarified effluent. It traverses the length of the tank to effluent weirs or submerged outlet tube located at the opposite end of the tank. Scum and floatable materials will rise to the clarifier surface. Baffles and skimming devices are provided to prevent this undesirable material from exiting the clarifier along with the effluent. The scum is removed either where it collects in the clarifier or it can be transported and concentrated in a different part of the tank.

There are a number of variations to the cocurrent flow pattern for rectangular clarifiers, and these are generally associated with the sludge hopper location. The sludge hopper can be placed at the influent end so that the sludge is removed more quickly. In this case, the sludge flow reverses itself and is called a *countercurrent sludge removal flow pattern*. In most rectangular primary clarifiers, the sludge hoppers are at the influent end for quick removal of the heavier solids. The sludge hopper can also be placed at midlength (midtank), so that the sludge does not have to travel to the end of the tank for its removal. In most rectangular secondary clarifiers, the sludge hoppers are at the opposite effluent end or midtank. There can also be more than one sludge hopper location provided, using different combinations of any of the three sludge hopper locations mentioned above (influent, effluent, or midtank), although this is rare. Sometimes cross collectors are used to collect sludge from multiple tanks for removal from a more accessible location. This is common with multiple-tank systems with midtank hoppers. Finally, sludge hoppers are not totally necessary because rectangular tanks can be equipped with traveling hydraulic suction devices that remove sludge from the tank. See Figure 9.3 for typical flow schemes of longitudinal rectangular clarifiers.

**TRANSVERSE FLOW TANKS.** In the transverse flow design, the influent flow enters from a channel provided along the long side of the rectangular tank. Effluent weirs can be placed on the opposite long side of the tank to give a short plug-flow type of pattern. If the effluent weirs are located along the influent side of the tank, this becomes a folded flow pattern. In transverse clarifier designs, sludge withdrawal

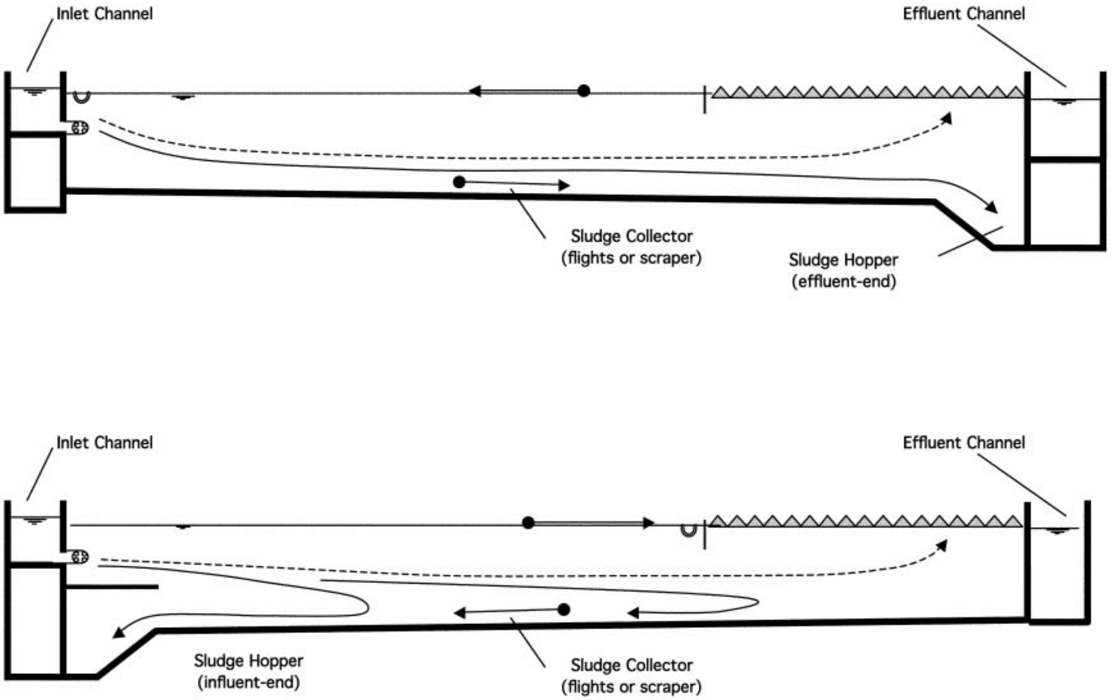


**FIGURE 9.3** Longitudinal section view of typical flow patterns in longitudinal flow rectangular clarifiers: top—cocurrent sludge removal and effluent pattern (effluent-end hopper); middle—countercurrent pattern (influent-end hopper); and bottom—mixed pattern (midtank hopper).

is generally accomplished by a traveling suction mechanism, making the provision of sludge hoppers unnecessary. Or, hoppers can be placed every so often (approximately every 10 m or 33 ft) along the short width of the tank where an embedded collection header with orifices can be placed. Figure 9.4 is a schematic of typical transverse tank flow patterns.

**VERTICAL FLOW TANKS.** In a vertical flow clarifier, the influent is introduced at a central location below the sludge blanket. Rather than using the usual sedimentation process, in this type of clarifier, the blanket is slightly fluidized and acts as a filtering agent, capturing the fine solids. A vertical flow clarifier is generally a circular or square design. Although the square configuration technically makes it a rectangular tank, this specialty clarifier is not very common and will not be covered further in this chapter.

**STACKED CLARIFIERS.** Longitudinal rectangular clarifiers also lend themselves well to stacking. Stacked clarifiers consist of hydraulically connected settling tanks, located one above the other, operating with a common water surface. In this sense, they become modular units. The stacking effect essentially increases the clarifier surface area without increasing clarifier facility footprint. They are also called *tray clarifiers* and can be double-decked or even triple-decked. The stacked clarifier design is similar to conventional rectangular clarifiers in terms of influent and effluent flow patterns, and in terms of solids collection and removal. Stacked clarifiers are covered in more detail later in this chapter.

## DIMENSIONS OF RECTANGULAR CLARIFIERS

**SURFACE AREA AND RELATIVE DIMENSIONS.** Recall that tank surface area, or the length times the width, is an important parameter in clarification. The surface area determines the overflow rate, which theoretically is equivalent to the settling velocity of the smallest particle to be removed and may be used as the basis of design. However, the design settling rate is often taken to be the blanket interface settling rate, as determined by batch settling tests. This velocity is generally greater than the velocity of the smallest particles. This means that a certain fraction of the lighter activated sludge particles that are not incorporated to larger flocs will not be removed.

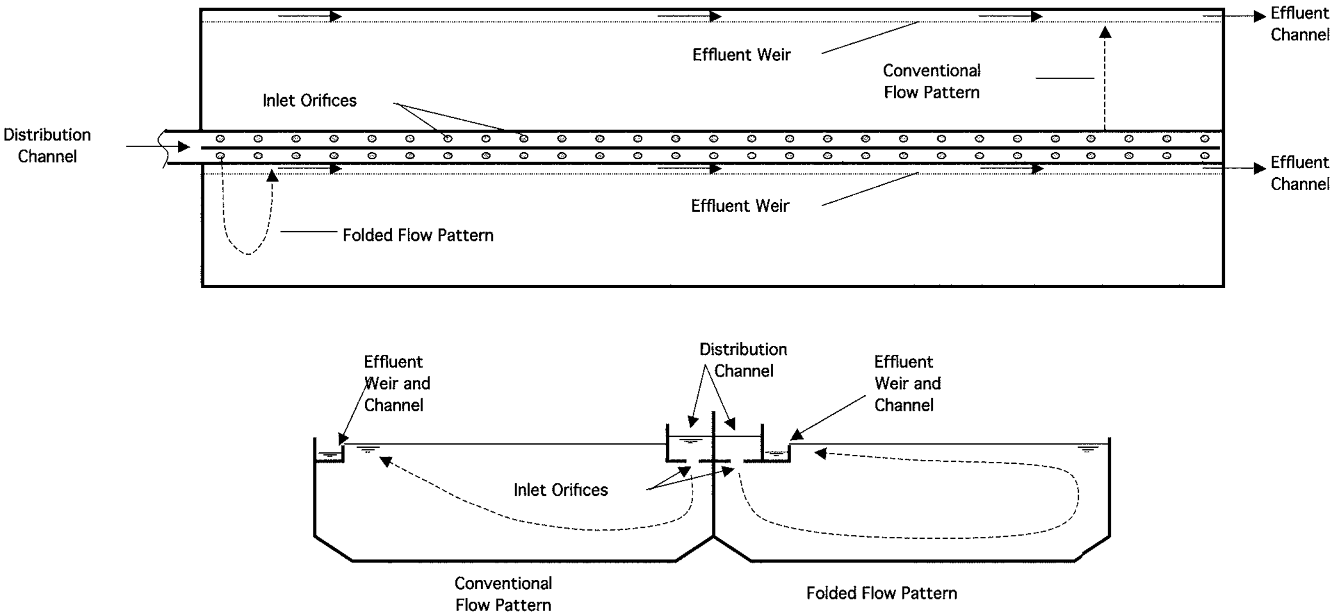


**FIGURE 9.4** Plan and section view of transverse tanks; conventional flow pattern top and left and folded flow pattern bottom and right. Note that sludge is either removed by suction device or by sludge headers embedded in a floor hopper.

There are acceptable minimum ratios of length-to-width that effectively limit the maximum size of rectangular tanks. The length-to-width ratios of longitudinal rectangular clarifiers may range from 1.5:1 to 15:1. A minimum length-to-width ratio of 3:1 is recommended to prevent short-circuiting (U.S. EPA, 1974) and, typically, the length-to-width ratio is greater than 5:1. Some references recommend that the length of the rectangular clarifier should not exceed 10 to 15 times the depth (Metcalf and Eddy, 2003). However, this length-to-depth ratio has been exceeded with success at larger plants. In addition, the design engineer should keep in mind, when determining the length, width, and depth, that these dimensions should be proportioned so that horizontal flow velocities are not excessive.

**LENGTH.** Rectangular clarifiers are seldom greater than 110 m (330 ft) in length and are typically 30 to 60 m (100 to 200 ft) long. As mentioned above, the longer lengths have been used with success at large plants. In very small clarifiers, such as those used in package plants, a minimum flow length of 3 m (10 ft) from inlet to outlet should be used to prevent short-circuiting (U.S. EPA, 1974). There is concern for possible suspended solids carryover with the increased hydraulic flow at the weir as the clarifier length is shortened. The ultimate length of the tank is limited by the flight collection system, because increasing the length of a chain-and-flight system also increases the mechanical stress on that system. Also, the longer the tank, the more difficult it is to transport sludge the entire length of the tank with one collector system. Sometimes, for long tanks and for tanks with midtank hoppers, multiple collector systems are used. This can increase the equipment cost, maintenance, and complexity of design.

**WIDTH.** For many years, wood was the standard material for collector flights. Because of a combination of the effects of deflection, buoyancy, and weight of the flights, the acceptable span of wooden flights restricted rectangular clarifiers with a single flight system in each tank to a nominal width of 6 m (20 ft). However, multiple parallel flights can be constructed in an extra wide tank with open side walls provided with columns supporting the collector sprockets. For instance, a 24-m- (80-ft-) wide tank could be fitted with four parallel sets of flights. It should be mentioned that recent innovations with fiberglass composite materials have allowed single flight systems to presently span clarifier widths of up to 10 m (33 ft).

The design option to not completely wall-off each rectangular tank is sometimes performed to save on concrete material costs, even though it slightly increases construction costs related to forming. For example, by alternately using columns instead of walls in a battery of clarifiers, a set of double clarifiers is created. One major disadvantage of this concept is that a larger percentage of tankage must be taken out of service and dewatered to repair only one of the chain-and-flight mechanisms. Also, it is not entirely clear if the hydraulic flow patterns are truly as stable as longitudinal tanks that are long and narrow. Without walls acting like longitudinal baffles to direct the flow, eddy currents could form to adversely affect hydraulic stability and suspended solids removal efficiency.

**DEPTH.** The design depth for any type of clarifier is often a contentious point of debate. Side water depth is generally measured at the effluent end wall for rectangular clarifiers. Although it is common for primary treatment rectangular clarifiers to be designed with a minimum depth of 2 m (7 ft), secondary clarifiers in activated sludge plants are generally deeper. The design trend, in recent years, has been towards increasing depths of secondary clarifiers for improved performance. Current practice is to provide a depth of approximately 4 to 5 m (approximately 12 to 16 ft), depending on the peak flows, sludge loading storage requirements, and available recycle. However, larger plants have reported success with rectangular clarifiers that were only 3 m (10 ft) deep (Stahl and Chen, 1996; Wahlberg et al., 1993 and 1994). Crosby (1984a and b) studied the effects of blankets and their maintenance. He indicated that it is the top of the blanket that determines the depth available for clarification. This means relatively shallow tanks with minimal blanket levels often perform as well as deeper tanks with thicker blankets. Where the overflow weirs are located where there is an upturn of the density current, it is good practice to provide a bottom depth below the weirs of at least 4 m (12 ft) (WPCF, 1959).

The design engineer must recognize that shallow clarifiers can limit the storage and thickening capability of secondary clarifiers in an activated sludge system. This, in turn, may decrease the return activated sludge (RAS) concentration and increase RAS pumping demands. Ample depth is recommended to provide for storage volume of solids and thickening during sustained peak flows and when solids loading exceeds recycle capacity (Boyle, 1975). If this storage is not provided, it is not uncommon for the sludge blanket of a heavily loaded clarifier at peak flow rates to fill the tank until a point is reached where the upflow velocity at the weir area starts to sweep the floc over the weirs. Increasing the RAS capability is helpful, but increasing the recycle rate increases the total hydraulic flow, horizontal velocities, and solids loading to the clarifier, each of which, in turn, can cause a clarifier to fail.

Finally, it is also important, from a biological treatment perspective, that unthickened solids are not recycled to the secondary process during peak loading.

## FLOW DISTRIBUTION TO MULTIPLE CLARIFIER UNITS

Flow distribution to rectangular clarifiers is straightforward, but somewhat challenging, in the sense that there are generally many more rectangular clarifiers for a given flow than circular clarifiers. Of course, equal flow distribution to a bank of clarifiers is essential for optimum performance. However, when thoroughly investigated, the hydraulic balance between clarifiers is often lacking. Flow imbalances can cause overloading of individual tanks, whereas other tanks can be underloaded and not contribute to their fair share of thickening or effluent production.

A covered distribution channel is generally provided ahead of primary clarifiers. For secondary clarifiers, an open distribution channel is generally provided to convey the mixed liquor from the reactor to multiple rectangular clarifiers. A long and fairly narrow channel can easily be constructed to run along the end of the tanks, with the channel flowing perpendicular to the flow of longitudinal clarifiers. For clarifiers of the same size or equal capacity, the flows should be distributed equally to each tank. In tanks of unequal dimensions, the flows are normally distributed in proportion to their respective surface areas.

Equal flow distribution is generally achieved by the use of inlet weirs, submerged orifices, or inlet gates. Weir inlets that discharge directly onto the tank surface should be avoided, although this type will better pass floatables. In primary tanks, this waterfall effect can contribute to odors. In secondary clarifiers, overflow weirs will exacerbate the effects of the density current. Submerged orifices are acceptable, but they can be restrictive during peak wet weather flows and may not be restrictive enough during low flows. Because the orifice size is fixed, submerged orifices may have to be accompanied by storm inlet gates. These may change the operation of the clarifier, as these gates would not necessarily have the same inlet characteristics as the orifices. Submerged inlet gates seem to provide the most flexibility, as their gate openings can be tailored for any extended flow condition that is encountered. They also allow a clarifier to be taken out of service easily.

Minimizing hydraulic losses in feed channels compared to inlet losses in the clarifiers will assure reasonably uniform flow to all tanks. Equal flow distribution to all tanks is also assured, not only at design flows, but also during significantly variable flowrates. For equal flow distribution, inlet gates can be throttled or orifices sized so that the head loss across the gate or orifice is approximately 10 times the total head loss of the channel at peak flow. For rectangular longitudinal tanks, head loss through the distribution channel itself can be kept fairly small because of the fact that the channel length spans the short side of the tanks.

Positive flow-splitting structures have also been used for distribution, although it is essential to design these facilities without extensive horizontal flow components after the split to keep individual flows equal. Open channel flumes have also been used and have the advantage of providing a means of flow measurement into each tank. Flow meters that are coupled with automatic valves have also been used, but have the disadvantage of high initial cost and high maintenance. It is not uncommon for a bank of clarifiers with an incorrectly tuned flow controller to experience flow disturbances because of the hunting action of the automatic valve. These flow perturbations can end up affecting the whole bank of clarifiers. See Figure 9.5 for typical flow splitting concepts.

Care should be taken in the sizing of distribution channels and clarifier inlets so as not to create excessive velocities that may shear the floc in secondary clarifiers. The inlet channel depth is generally set by the depth of the clarifier. Therefore, a wider channel will provide more flow area so only a small head loss is experienced down

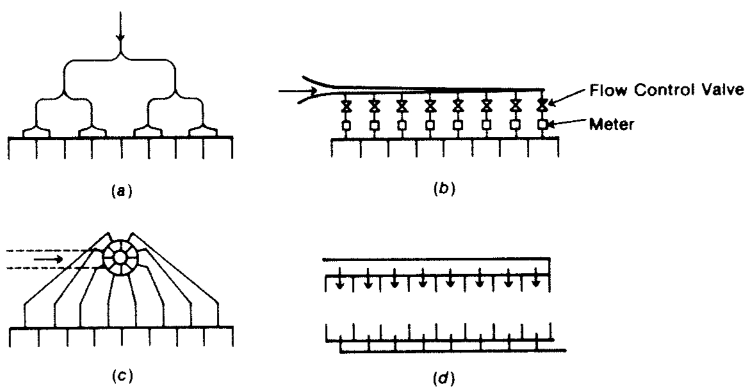


**FIGURE 9.5** Alternative concepts in flow splitting: (a) geometric symmetry, (b) inlet flowmeter and automatic control valve, (c) hydraulic weir splitting, and (d) feed gate throttling with effluent weirs at same elevation

the channel. This is to ensure that each tank will see essentially the same water surface elevation. Secondary clarifier distribution channels should be slightly aerated, unless the flow velocity is high enough to keep the floc in suspension. Velocities should be approximately 20 to 40 cm/s (0.7 to 1.3 fps). High aeration rates can also lead to floc breakup or exacerbate foam problems and should be avoided. It can be difficult to achieve the balance between the opposing requirements of keeping the floc in suspension without contributing to its breakup. However, even if the floc is broken up to some degree, it can be reflocculated as it enters the clarifier with the use of an inlet flocculator zone. The recommended G value of flocculation in an aerated channel is approximately 70/s (Parker et al., 1971). Lower G values are recommended to follow in the clarifier inlet/flocculator zone.

Floatables will generally collect in the inlet channel. Occasional aeration of a primary inlet channel will help to break up scum, but odor treatment may be necessary for this air. Providing an occasional downward opening gate in the distribution channel allows floatables to pass into a clarifier for subsequent removal. Floatables will generally concentrate on the downstream end of a distribution channel. Assuming the usual freeboard of approximately 0.5 m (1.5 to 2 ft) to the channel surface, the operator generally has a poorly leveraged position to lift large floating masses of rags that sometimes weave together while being aerated in the channel. A wench or boom truck is generally necessary to lift out such a mass. Sometimes, an unaerated inlet channel is provided with a scraper that helps to remove settled material. When this is done, the scraper can also be designed to assist with foam and floatable removal on the return pass. If the problem is severe, a skimming flight and beach mechanism analogous to primary tank scum removal can help with dead-ended floatables. Otherwise, partially submerged inclined bars can be situated at the end of a channel to allow some dewatering before removal.

## INLET CONDITIONS AND DESIGN

**GENERAL INLET CONDITIONS.** Inlets should be designed to dissipate the kinetic energy or velocity head of the mixed liquor. They should be able to distribute the flow equally in vertical and horizontal directions so that the whole cross-sectional area of the tank is used. Inlets should also be designed to prevent short circuiting (NEIWPCC, 1998), mitigate the effects of density currents, and minimize blanket disturbances. If possible, inlets should be designed to promote flocculation and prevent floc breakup. Inlets and baffles should be designed so as not to affect the sludge hoppers when these are placed at the inlet end of the tank. This problem convinced Gould (1943) to place hoppers for secondary clarifiers toward the downstream end of the tank.

Poor distribution and jetting of inlets result in short-circuiting that can be evidenced in dye studies (Crosby, 1984a). Poor inlet design can also exacerbate the effects of density currents and produce scouring of sludge solids that have previously been settled, especially in rapidly settling sludges, which produce higher density currents and turbulence. Finally, it should be noted that for the same length-to-width ratio for any side water depth, tanks of smaller area (shorter tanks) generally have more trouble with inlet and outlet energy dissipation than tanks of larger area (longer tanks), so that the inlets for smaller tanks may need to be designed for more energy dissipation to counter this effect. When there is too much inlet turbulence, the effective settling area of a clarifier is reduced (U.S. Army, 1988).

**FLOW DISTRIBUTION WITHIN THE CLARIFIER.** The introduction of flow to an individual tank is sometimes accomplished by spanning the width of the clarifier with a short open channel or by providing a manifold piping system. In both of these options, the flow is directed to multiple inlet openings in the tank. The multiple inlet ports are situated and sized to uniformly distribute flow over the width of the clarifier. For instance, in a 6-m- (20-ft-) wide tank, there are typically 3 to 4 inlet ports. Maximum horizontal spacing between inlets is generally approximately 2 m (6.5 ft), but less than 3 m (10 ft). Sometimes an inlet baffle is placed in the flow path of the inlet stream. It may be a solid target baffle to deflect the flow or a perforated (finger) baffle to break up any jetting action and disperse the flow. Pumped flow or any type of waterfall into the tanks should be avoided.

Although the inlet size varies considerably from one design to another, sufficient head loss should be provided to assure effective distribution. Head loss though a perforated inlet plate should be approximately 4 times the kinetic energy or velocity head of the approaching flow (WPCF, 1985). Each slot should not be less than 5 cm (2 in) wide. Smaller inlet openings are generally avoided because of possible fouling and formation of smaller jets of flow. Especially in primary sedimentation and depending on the type of pretreatment provided, rags or plastic bags can be present, which can plug smaller openings.

Equal distribution of flow through multiple inlet ports is a common, but not insurmountable, problem. An approximate equal distribution is generally accomplished by making the head loss across each inlet port relatively large in comparison to the head loss associated with the various flow paths (Rich, 1974). Some inlet design data from various sources are shown in Table 9.1. Detailed information regarding the hydraulics and design of multiple inlet ports is discussed by Benefield et al. (1984). Chao and Trussell (1980) have developed more sophisticated hydraulic methods to achieve favorable flow distribution.

Inlet design is somewhat more complicated for transverse tanks, because the inlet channel extends the length of the tank. Equal distribution can be compromised by the head loss associated with the length of the inlet channel. Equal head loss may be set for equal distribution at a certain design flow, but it may vary for other flow conditions. However, some designers, who are experienced with transverse clarifiers,

**TABLE 9.1** Inlet design for rectangular clarifiers.

| Source | Design data |
|---|---|
| (Rich, 1974) | $h_f = h_{f1}$ (1-r (1/n))<br>$h_f = (1/3)\, h_{f0}$<br>Where<br>$h_f$ = Frictional head loss in pipe from first to last inlet port, m (ft);<br>$h_{f1}$ = Frictional head loss across first inlet port, m (ft);<br>$h_{f0}$ = Frictional head loss across pipe when entire entrance flow is to flow through the pipe, m (ft);<br>$r$ = Ratio of flow through the last inlet port to that through the first inlet port (unitless); and<br>$n$ = 1/2 for orifices, 3/4 for straight edge weirs, and 5/2 for V-notch weirs. |
| Ten States Standards (GLUMRB, 1978) (NEIWPCC, 1998) | Channels designed for a velocity of 0.3 m/s (1 fps) at 50% design flow. |
| MOP 8 (WPCF, 1977) | Inlet port velocities 75 to 150 mm/s (15 to 30 ft/min). |
| U.S. EPA Process Design Manual, Suspended Solids Removal (U.S. EPA, 1975) | Maximum spacing between ports, 3 m (10 ft). |

report that this is not an overwhelming problem, and often a more stable flow pattern is obtained. Some designs get around this problem by providing an inlet channel with a sloping bottom or variable width or by providing variable inlet orifices along the channel length.

**INLET DESIGN.** It is poor practice to place the inlet too high in the tank so as to introduce mixed liquor in the clear water zone. This would have the effect of increasing the potential energy of the mixed liquor solids, thereby increasing the density current. For the same reasons, small deflectors to direct inflow upwards have not been successful (Crosby, 1983). Density effects that may lead to excessive bottom currents can be mitigated by decreasing the potential energy associated with the solids falling to the bottom of the tank. This can be done by positioning the inlet lower in the tank without placing it in the thickening zone, which is typically reserved for the bottom 1 m (3 ft) of the tank. Locating the inlet too low may scour the solids on the bottom and lead to resuspension. Inlet apertures should be positioned from approximately the 2 m (6.5 ft) depth to midtank depth. A method to calculate an inlet height is given by Krebs et al. (1995).

There is not much standardization of inlet design observed. An inlet that was fabricated with large cross-section tubes for low velocities and floc stability was introduced by Larsen (1977) (see Figure 9.6). Enlarging the size of the inlet zone and using the inlet energy for flocculation can improve suspended solids removals. Impinging flow streams against one another is an effective way of promoting flocculation. Inlet port velocities are typically limited to a range of 0.075 to 0.150 m/s (0.25 to 0.5 ft/sec). Das et al. (1993) demonstrated that velocities in excess of 0.6 m/s (2 ft/sec) may cause deflocculation of the activated sludge solids. Figure 9.7 shows a clarifier inlet diffuser design that works very well in both primary and secondary clarifiers.

To enable floatables to pass from the inlet distribution channel into the clarifier, the inlet ports can be designed to maintain unsubmerged conditions. Otherwise, special provisions must be made for easy removal of floating materials trapped in inlet structures not fully submerged but having submerged ports. As mentioned before, slide gates are commonly provided for flow adjustment to multiple inlets or for taking a clarifier out of service.

**INLET BAFFLES AND FLOCCULATION ZONES.** There is generally some type of baffle immediately downstream of the inlet openings to prevent jetting of flow into the tanks. The target baffles can be simple baffle walls, solid or perforated, spanning

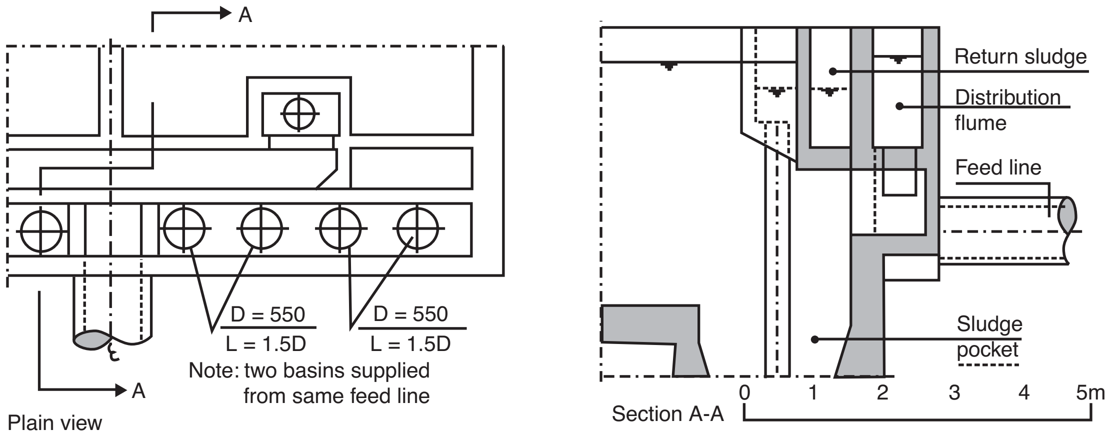


**FIGURE 9.6** Inlet design of Larsen (1977) to avoid floc breakup (note: D is in millimeters).

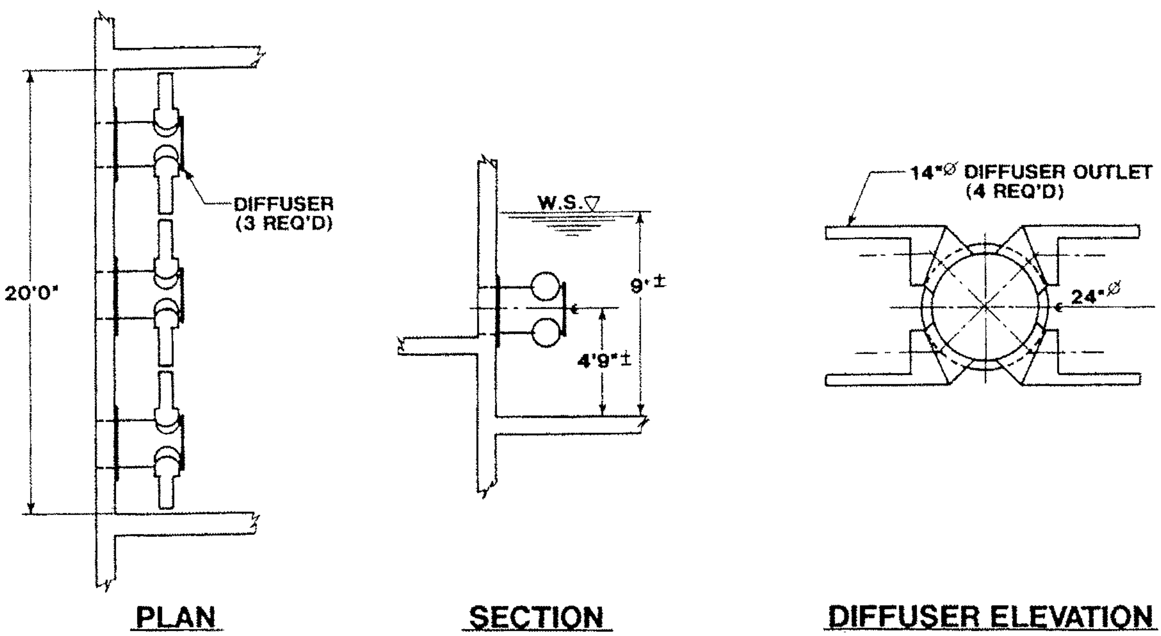


**FIGURE 9.7** Secondary clarifier inlet diffuser used by the Los Angeles County Sanitation Districts (1 in. = 2.54 cm; 1 ft = 0.304 8 m).

across the width of the clarifier, or they can be specialized inlet diffusers (see Figures 9.8, 9.9, and 9.10). Target baffles are recommended to extend from an elevation just below the water surface to 15 to 30 cm (6 to 12 in.) below the inlet points (WPCF, 1977).

Mau (1959) showed that a single vertical row of baffles that was slotted was effective in distributing flow. However, a second row of vertical slotted baffles, where the boards are opposed to the slots of the first baffle, improved energy dissipation and performance by causing flow impingement. Kawamura (1981) recommended the installation of three sets of perforated baffles spanning the full cross section. Okuno and Fukada (1982) observed the best removal efficiencies from baffles that had 5% open areas. Other investigators have tried more sophisticated designs (Collins and Crosby, 1980; Crosby, 1984b; Rohlich, 1951), with different degrees of success. Price et al. (1974) concluded that lack of symmetry is to be avoided, and complicated inlets do not necessarily give better results than simple ones.

Since early publications by Camp (1936, 1945, and 1953), a number of investigators have demonstrated that using the incoming energy to promote flocculation improves clarifier performance. This is used in the water treatment industry to polish drinking water after addition of chemicals for coagulation and flocculation, but, by no means, is it exclusive of wastewater clarification. Circular tanks lend themselves well to a central flocculation chamber, and, although there has been less research for rectangular tanks, an energy dissipation volume can also be provided with a defined

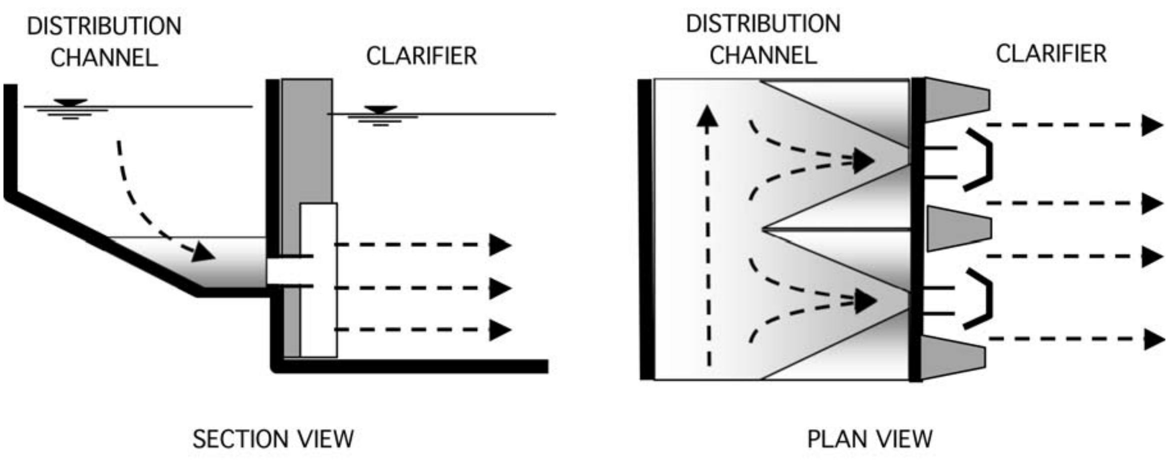


**FIGURE 9.8** Distribution channel with funnel-shaped floor (Krauth, 1993) with a Stuttgart inlet (Popel and Weidner, 1963).

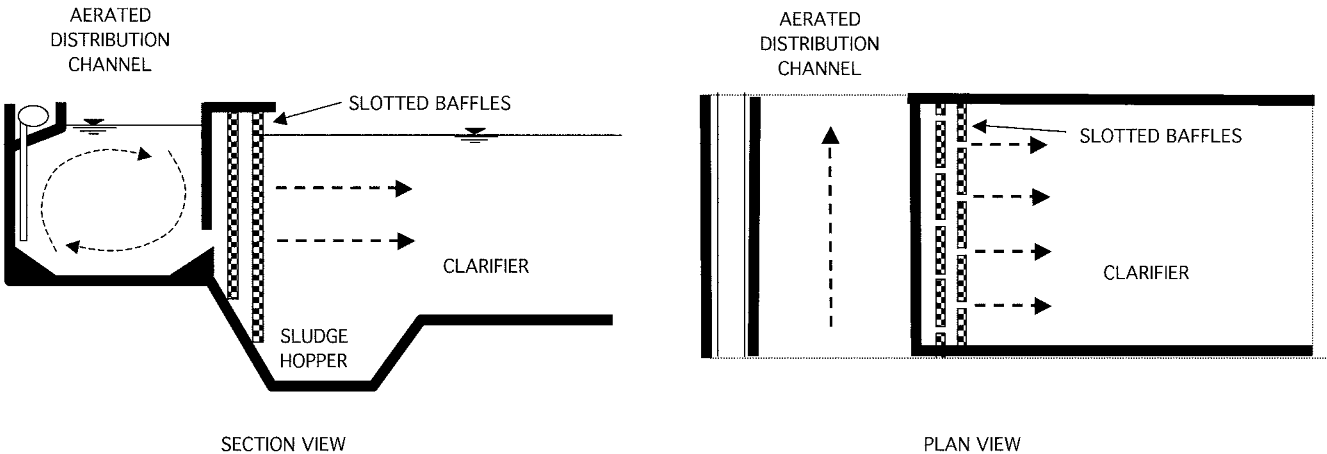


**FIGURE 9.9** Aerated distribution channel (Krauth, 1993) with two staggered slotted baffles to dissipate inlet energy.

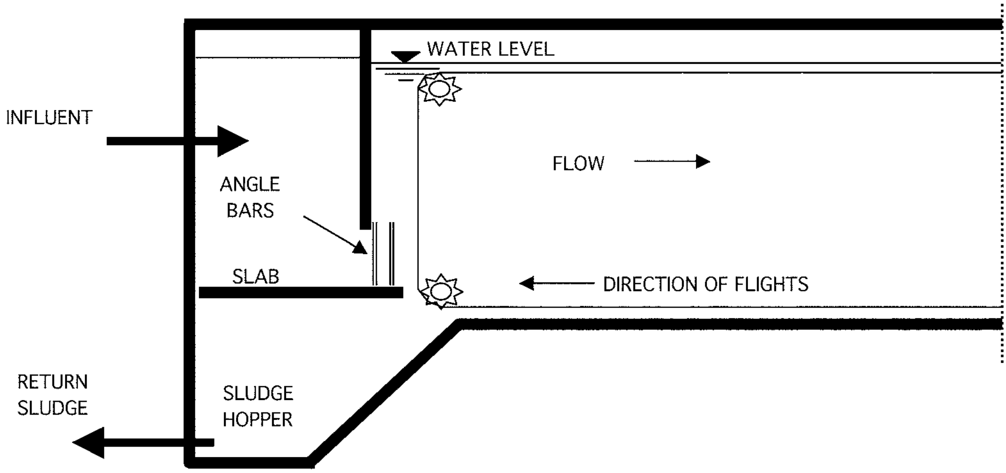


**FIGURE 9.10** Aerated distribution channel with horizontal slab deflecting inlet flow energy from sludge hopper at inlet end (reprinted from Krebs et al. [1995] Inlet Structure Design for Final Clarifiers, *J. Environ. Eng.*, Am. Soc. Civ. Eng., **121** (8), 558–564, permission from the publisher, ASCE).

turbulence level to act as a flocculation chamber. For instance, slotted baffles or rows of angle bars can be placed near the inlet opening to enhance flocculation and also reduce kinetic energy. The recommended G value of flocculation in the inlet is 30 to 70/s (Parker et al., 1971). The required volume of the inlet flocculator zone is calculated by residence time required for flocculation to be completed, which is approximately 8 to 20 minutes.

Experimentation by Kalbskopf and Herter (1984) indicated that a separate flocculation zone, operated at the inlet end of rectangular clarifiers, can be effective in producing a better effluent, although it is not a large improvement. Their data showed that a separate flocculation tank with two paddle mixers located ahead of the clarifiers improved clarification, and inlet zone flocculation improved it even further. The counter-rotating paddles used as stirrers had horizontal axes and rotated at 1.4 to 2.8 rpm (see Figures 9.11 and 9.12).

## LOCATION OF THE SLUDGE HOPPER

Sludge hopper location is critical to the design and operation of the clarifier. Hoppers can be placed on the influent end, effluent end, and at any location in the midtank

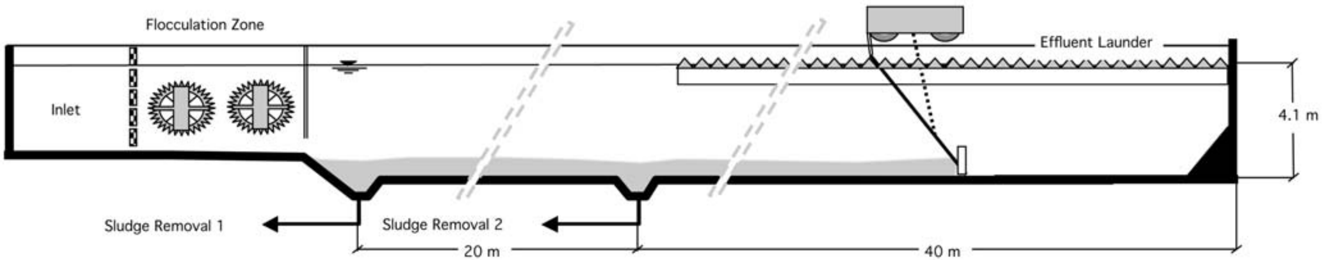


**FIGURE 9.11** Flocculator inlet zone with paddles. Sludge withdrawn near inlet and one-third down the length of the tank (Kalbskopf and Herter, 1984).

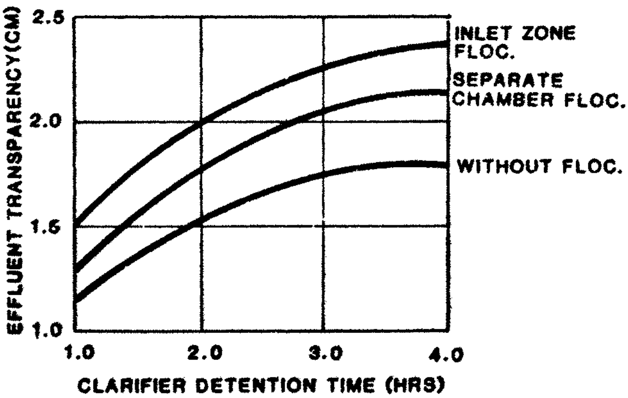


FIGURE 9.12 Improvement of effluent transparency with flocculation zone (Kalbskopf and Herter, 1984).

region. Sometimes, more than one hopper is provided for multiple scraper systems, so that a sludge removal header can be placed every 10 m (33 ft). In general, the clarifier floor is typically inclined toward the sludge hoppers with an average slope of 1%. Sometimes, more than one hopper location is provided. Hoppers at either end of the tank are most amenable to shorter runs of piping through galleries to the recycle pumps. Midtank hoppers are sometimes used when internal baffles are provided, because the gap provided by the hoppers and between the sludge collectors are a convenient location for a baffle. Midtank hoppers can also have transverse collection systems such that sludge removal is slightly different from tank to tank.

The optimum hopper location for longitudinal rectangular clarifiers has been debated for many years, especially for secondary clarifiers. In the early years of activated sludge treatment, hoppers were typically located at the front end of the clarifiers with a single flight system that directed the sludge to hopper. Gould (1950) developed the concept of moving the hopper to midlength or the effluent end. Over the last few decades, the trend for longitudinal rectangular secondary clarifier has been to follow the concept of Gould, where hoppers are placed at midtank or the effluent end. These tanks are commonly used in large plants and are designed to minimize density currents and to avoid other hydraulic problems. Gould tanks are characterized by high length-to-width ratios and effluent launders that are located away from the clarifier influent (see Figure 9.13). Primary clarifiers, on the other hand, are still generally provided with their hopper on the inlet side of the tank.

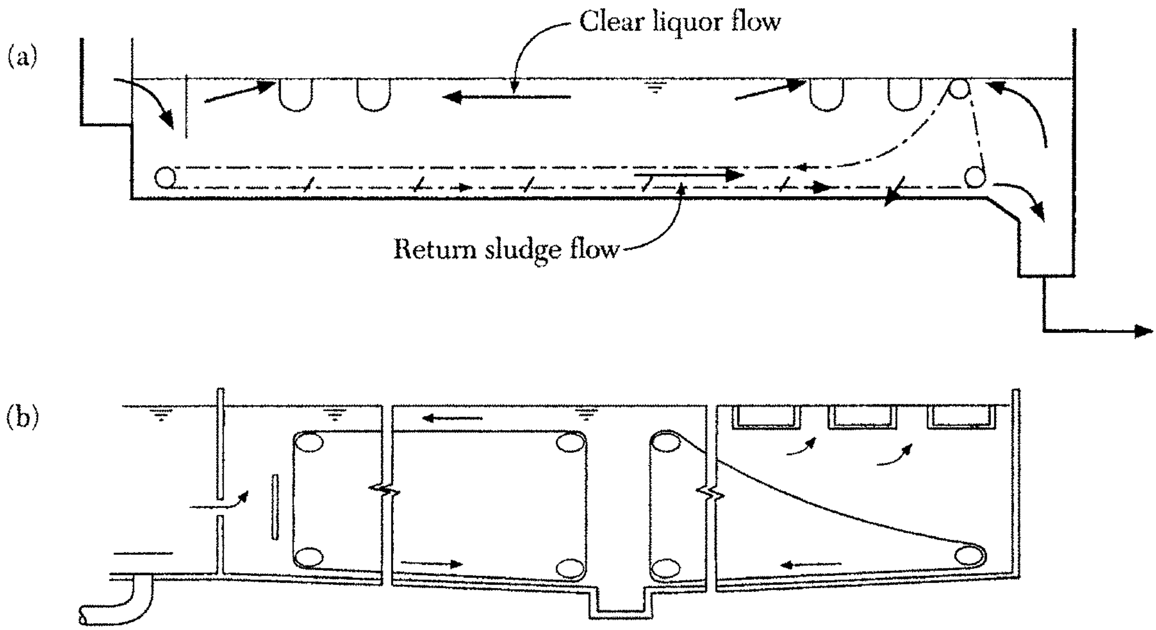


FIGURE 9.13 (a) Gould tank-type I with sludge hopper at outlet end, and (b) Gould tank-type II with sludge hopper at midpoint (reprinted from *Secondary Settling Tanks,* ISBN: 1900202035, with permission from the copyright holder, IWA).

The typical hopper shape for rectangular clarifiers is an inverted pyramid with a rectangular opening on top. The sides are recommended to have the slope of 52 degrees to prevent solids from accumulating on the upper walls. A single rectangular tank may have two or more withdrawal hoppers, each equipped with a withdrawal pipe. When hoppers are manifolded together, some investigators claim one hopper may start to remove diluted sludge while the other allows sludge to accumulate and thicken, exacerbating the flow difference. To prevent this from occurring, separate controls for each hopper are sometimes recommended. At the very least, there should be provision for backflushing each hopper individually (isolation valves and copious amounts of flushing water). Sometimes, a tank with a plugged hopper can be partially drained and the higher head in the other tanks can be used to backflush the blockage.

**INFLUENT END HOPPERS.** Sludge hoppers in longitudinal tanks are often placed at the inlet end, just below the inlet structure, and combined with scraper

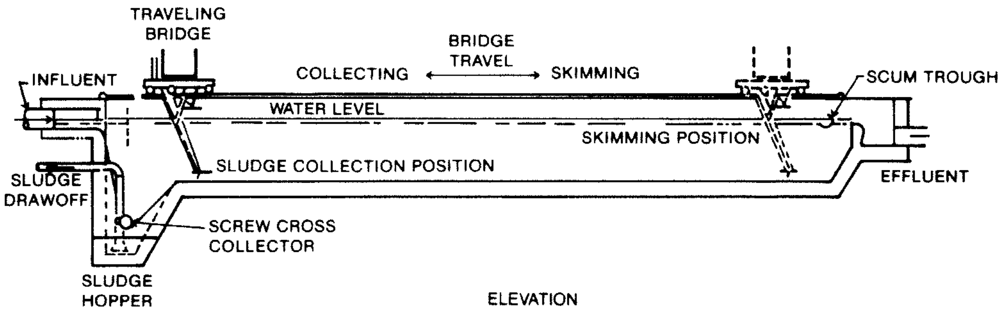


**FIGURE 9.14** Rectangular clarifier with traveling bridge sludge scraper (note that if suction mechanism was provided, hopper would not be needed).

removal systems (Figure 9.14). This is done more in primary tanks than secondary and tertiary tanks. Even though activated sludge solids are fairly light, if one looks at typical settling velocities, most flocculated solids will generally separate from the bulk fluid fairly quickly after the inlet. This change in velocity at the front end of a rectangular clarifier is in the range of 0.15 to 0.60 m/s (0.5 to 2 fps) through the inlet, to a range of 3 to 15 mm/s (0.5 to 3 fpm) in the tank proper. The idea behind placement of the sludge hopper at the influent end would be to allow early removal of the bulk of the sludge from the clarifier. This may be acceptable for primary clarifiers, which have heavier solids and low solids loading, but this may not be the case for secondary clarifiers. In the secondary clarifiers, the lighter activated sludge flocs that settle more slowly are removed more slowly, as scrapers are required to move that sludge from the effluent end to the front hopper.

Returning the sludge flow promptly also means that the horizontal flow velocity in the tank is not increased as it is by the return flowrate with a hopper at the outlet end. However, in situ experiments and numerical simulations show that these assumptions are not necessarily valid. The flight scraper induces a volumetric flowrate at the bottom of the tanks, which flows against the density current flowing forward above it. Also, flow instabilities are created when sludge is moved in the opposite direction to the density current. Because the scraping of the settled sludge to the influent end hopper is performed against the main flow direction, there is always concern over the possibility of breaking up the fragile activated sludge flocs, with a subsequent resuspension and carryover of fine flocs in the effluent. It can be deduced, therefore, that the influent end hopper design is not an ideal application for activated sludge clarifiers.

**EFFLUENT END HOPPERS.** Placement of the hopper at the end of the tanks (Gould Tank—Type I) reduces the adverse effects of countercurrent sludge removal that are encountered with the inlet-end hopper (see Figure 9.13a). The effluent end hopper design conceptually provides a more ideal solution for minimizing the breakup of the biological flocs, because the sludge transport now takes place in the same direction as the bottom density current. The sludge is also kept out of the relatively turbulent region of the inlet. Furthermore, the longer sludge detention time, resulting from the effluent end hopper arrangement, can enhance the flocculation and the dynamic filtration effects on the flocculent particles. Both of these effects help to separate the sludge particles from the fluid flow, which translate into clarifiers operating well at higher overflow rates. For instance, Wahlberg et al. (1993) showed that rectangular tanks with effluent end sludge collection can perform exceptionally well up to surface overflow rates (SORs) of 3.4 m/h.

However, one drawback of the effluent end hopper is that a large amount of solids are transported into the effluent region. This increases the danger of a washout of effluent suspended solids (ESS) if the sludge is bulking or does not have good settling properties. Furthermore, in the effluent end hopper design, the actual horizontal flowrate through the tank consists of the both the effluent flow plus the recycle flow under all operating conditions. Therefore, very high recycle rates can be self-defeating. Also, longer detention times in relatively long rectangular tanks can result in denitrification in secondary clarifiers and related degradation of the effluent.

**MIDLENGTH HOPPERS.** The midlength hopper design is a method of trying to get the best effects of both the influent and the effluent hopper designs. In rectangular tanks exceeding 40 m (130 ft) in length, the sludge hopper can be situated halfway or further (perhaps two-thirds of the way) towards the end wall. This is referred to as a Gould Tank—Type II. This means the sludge is scraped with the main direction of flow in the first half of the tank and against the main direction of flow in the second half of the tank (see Figure 9.13b). The figure shows that the midlength hopper design uses the density current to transfer most of the sludge to the midlength hopper for relatively quick withdrawal. Many operators prefer an activated sludge with low detention time in the clarifier, as this ensures the viability of the organisms in the sludge returning to the aeration tank. Furthermore, Figure 9.13b shows that the midlength hopper decreases the possibility of short-circuiting of the effluent to the weirs. This design uses the countercurrent flow pattern that develops on the surface (primarily developed by the density current on the bottom and relative to the main direction of flow) to cause the effluent to travel a long and circuitous path to the effluent weirs. In general, with the absence of a density current in the vicinity of the weir area, the midlength hopper design can provide good effluent quality, while also satisfying the goal of rapid sludge withdrawal.

**MULTIPLE HOPPER LOCATIONS.** As a slight variation, two or more hoppers can be placed in the intermediate region. This concept has been shown to perform very well. This is because sludge removal is not disturbed by turbulence in the inlet region, and sludge transport distances are decreased. Sludge is directed with the bottom current in the high-velocity region. After the bulk of sludge and effluent are removed, the remaining velocities are so small that the lighter sludge can easily be transported (Wilson and Ballotti, 1988). Past the sludge hopper at the midlength of the tank, there is only a small amount of sludge to transport. Because there are different sludge loading conditions in the first and second parts of the tank, a certain degree of flexibility is acquired so that each part of the tank can have different flight or scraper speeds, different blade heights, or different distances between the blades. Finally, another variation of hopper location has been investigated by Kalbskopf and Herter (1984). This clarifier system had blade scraper removal systems with two sludge withdrawal locations, one at the inlet and the second after one-third of the tank length (see Figure 9.11).

## SLUDGE REMOVAL SYSTEMS

Sludge removal systems for rectangular tanks are generally chain-and-flight or traveling bridge units. Typically, chain-and-flight sludge removal systems are used in rectangular clarifiers in the United States. Longer tanks generally have these systems. Traveling bridge collectors have been developed and used extensively in Europe for rectangular basins handling flows greater than approximately 1.55 $m^3/s$ (l mgd). They have also been used in the United States, but are not broadly accepted because of their higher construction and maintenance costs. Hydraulic suction systems using floating pontoons have been used on occasion, but have not become commonplace. Another fairly recent development is the reciprocating flight collector shown in Figure 9.15. The triangular shape of the flights allows the sludge to be alternately pushed on the forward stroke and then allowed to slide over the flight on the return stroke. Recall that, in longitudinal tanks, the sludge removal generally takes place with (cocurrent) or against (countercurrent) the main direction of flow. However, in

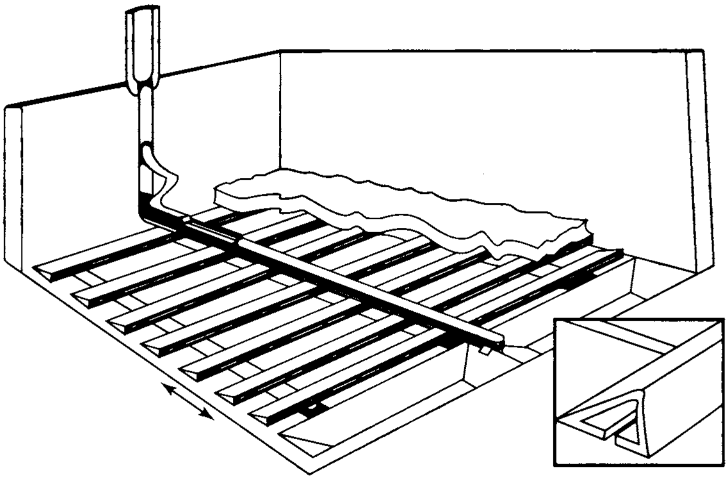

**FIGURE 9.15** Reciprocating flight sludge collector designed to allow sludge to slide over angled face on return stroke.

transverse tanks provided with flights or scrapers, the sludge is still directed along the longitudinal axis of the tank, or across (crosscurrent) the main direction flow. As an alternative, transverse tanks can also be provided with suction collectors.

**CHAIN-AND-FLIGHT COLLECTORS.** In the chain-and-flight design, the flights are attached to two parallel chains driven by sprockets and move along the clarifier floor, scraping the settled sludge to collection hoppers (Figure 9.16). The sprocket wheels are mounted on rotating shafts. The flights move slowly along the clarifier floor, scraping the settled sludge to the sludge hopper. At the same time, on their return path near the surface, the partially submerged flights serve as skimming devices to push any floating solids or foam to a skimmer pan or trough. This requires the use of four rotation points or sprockets. If the flights are not used to move the skimmings on the surface, only three rotation points are required. Sometimes, a five-sprocket system is used. It is analogous in operation to a four-sprocket design, with the extra sprocket helping to guide and hold down the flights in the bottom midtank area of long tanks.

Historically, redwood or metal flights and metal chains have been used for sludge collecting systems. However, it is more common to now use nonmetallic flights, chains, and sprockets in the clarifiers to minimize corrosion and wearing

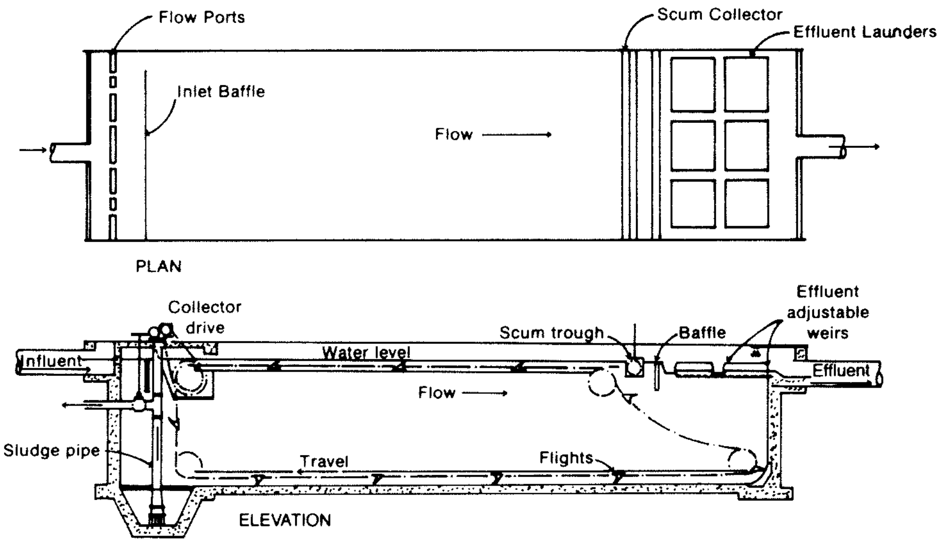


**FIGURE 9.16** Rectangular clarifier with chain-and-flight collector and hopper on influent end.

problems. Typical flights measure 5 to 6 m long (16 to 20 ft), depending on the width of the clarifier. Newer flights can be 10 m (33 ft) long. Flights generally are spaced at 3-m (10-ft) intervals and travel at speeds of 5 to 15 mm/s (1 to 3 fpm). The speed of the flights should be set to approach the rate of sludge movement by the density current to enhance flow stability and prevent turbulence in cocurrent sludge removal systems. Plastic-wear shoes fixed to the flights allow them to slide on rails near the surface and wear strips on the bottom of the clarifier so that the chain does not bear the full weight of the flights. In this way, less mechanical stress is developed, and less power is used to move the flights. An adjustable rubber scraper should be attached to the bottom edge, and the sides of at least some of the flights to provide complete sludge scraping and prevent unwanted stationary sludge deposits. A floor slope of 1% is common to facilitate the movement of sludge within these tanks. The slope on the bottom is useful for cleaning the tank after emptying.

The number of chains and direction of removal depend on the hopper location. To minimize the number of withdrawal pipes, some designs have provided for flights to move sludge to a single hopper for subsequent withdrawal. This design concept has limitations with activated sludge blankets. Moving all the sludge to a corner of a rectangular clarifier may produce solids flux limitation and potential for solids scour and resuspension.

Chain-and-flight systems can be applied to rectangular clarifiers up to 90 m (300 ft) long. The total length of the flight chain is limited by the stresses exerted on the chains. However, sludge characteristics and withdrawal rates may also determine the maximum unit size of the chain-and-flight's sludge collecting system. These same constraints for wooden flights and metal chains are similarly applied to the plastic or fiberglass flights and plastic chains. As plastic chain wears and stretches over time, it is normal maintenance to remove a link approximately every year on long clarifiers.

Because of their submergence in wastewater, the chains, cross flights, sprockets, shafts, and bearings have more serious maintenance problems than those of other sludge collecting systems, in which the equipment is not submerged. Many times, maintenance requires fully dewatering the clarifiers for repairs. This is all the more reason to fit the clarifier with materials that will resist corrosion and wear.

Flight intervals of 3 m (10 ft) provide a continuous series of scraping actions at a controlled slow moving speed. Such slow, but continuous, sludge collecting systems can help sludge thickening and can also prevent excessive sludge thickness on the clarifier floor. The continuous scraping actions can also be quite effective in concentrating the sludge of clarifiers with low solids loadings. Because sludge movements on the clarifier floors are caused, to a large degree, by the general hydraulic patterns or the density current, the movements of the sludge collector may actually have little effect on the movement of sludge. It is common for the height of the activated sludge blanket to be many times greater than the height of the flights. This means that the chain-and-flight sludge collectors may simply move at the bottom of the sludge blankets and have very little real effect on the gross movement of the majority of the sludge solids (Gould, 1950). It can be argued, however, that it may be operationally counterproductive to have stagnant sludge solids at the bottom of the clarifier without some movement being provided by the flights. McKinney (1977) suggested that rectangular clarifiers with chain-and-flight sludge collectors be used only for the activated sludge systems where low mixed liquor suspended solids (MLSS) levels would be maintained. However, there are many facilities whose operational experience suggest otherwise.

**TRAVELING BRIDGE COLLECTORS.** A traveling bridge collector can be equipped with either a scraper or a suction system. These systems were developed to solve the problem of having to dewater the chain-and-flight system when maintenance was required. While these systems are easier to maintain, they generally allow the settled sludge to accumulate to a greater extent before moving to the sludge hoppers for withdrawal. This accumulation may not create a serious problem for some types of chemical sludge from tertiary clarifiers. However, it may cause more serious problems for the final clarifiers in activated sludge plants, especially if a highly nitrified sludge is allowed enough time to denitrify. This combination can produce a rising sludge condition. Therefore, the size of rectangular clarifiers that can be effectively equipped with a traveling bridge sludge collecting system is affected by sludge characteristics, sludge loadings, and acceptable time for temporary sludge accumulation on the clarifier floor. Although a traveling bridge with scrapers is used for primary and some tertiary clarifiers, it seldom is used for activated sludge. A traveling bridge with suction mechanism is relatively common with secondary clarifiers, but not with primary clarifiers, because of clogging problems.

***Traveling Bridge Scraper Systems.*** A traveling bridge can be equipped with a single cross scraper that can be raised or lowered. Blade scraper systems are used mostly in Europe. The traveling bridge travels longitudinally back and forth on the rails located along both sides of clarifier walls. The weight of the scraper blade is often supported by wheels to minimize the frictional resistance and help preserve sealing strips. The blade is lowered to the bottom at the outlet end and is then pulled by a traveling bridge toward the hopper at the inlet, which is the usual hopper location for this type of system (Figure 9.14). The settled and thickened sludge on the tank floor accumulates in front of the moving blade. As the scraper moves back from the sludge hoppers toward the other end of the clarifier, the scraper is raised so that it can function as a skimming device at the clarifier surface. Otherwise, the blade can be lifted above the water surface so that the bridge can move back to the effluent launder end with a relatively high velocity. The height of the scraper blades has to be related to the recycle sludge flowrate, the sludge concentration at the bottom, and the removal velocity of the blades. Abwassertechnische Vereiningung (ATV) (ATV, 1988) gives guidance in determining the blade heights for scraper systems in rectangular and circular tanks. The sludge transport is assisted by a bottom slope of 1 to 2%, angled towards the sludge hopper. After being scraped into the hopper, the return sludge is withdrawn through a rising pipe or a siphon pipe.

***Traveling Bridge Suction Systems.*** Another sludge removal concept is the use of suction piping. Instead of a scraper blade pushing sludge to a hopper, the traveling bridge can be equipped with a suction system that removes sludge where it settles.

The sludge is discharged to a collection trough that runs along the length of the tank. These types of systems can travel back and forth in the tank. The suction systems are either equipped with a pump or make use of a siphon effect from the differential head between the water levels in the clarifier and the return sludge channel. The latter requires the use of a priming procedure to start the siphon flow. However, sometimes, air lift pumps have to be used to induce hydraulic suction. Air lift pumps, however, have pumping height limitations because the air addition point should be at least two times as far below the water surface as the height of the lift above the water. Another essential feature is that the top of the airlift pump discharge line must be cleared of air with a vacuum device to prevent airlocking. Figure 9.17 shows two traveling bridge suction devices using airlift and centrifugal devices to move the fluid.

One advantage of a hydraulic suction device is that it can be programmed to spend a higher percentage of time at the front end of the tanks, where presumably large quantities of sludge are deposited soon after entering the tank. The adverse effects of long sludge accumulation times for a highly nitrified activated sludge can be minimized by a suction system for sludge collection. In the case of the air lift suction system, this allows for the sludge to be aerated even before it reaches the aeration basin. Although this was thought to be advantageous in the past, with the advent of anoxic and anaerobic selectors, the aeration of return sludge could adversely affect the operation of those systems.

For a tank length higher than 40 m, ATV (1995) recommends increasing the removal frequency by installing two suction collectors. Although it recommends that longitudinal clarifiers should not exceed 60 m in length, transverse tanks can be much longer. For instance, at the wastewater treatment plant of the City of Zurich, Switzerland, transverse tanks with a length of 142 m (465 ft) are operated successfully with three suction removal systems operating in each tank (see Figure 9.18).

**DISCUSSION OF OTHER CONSIDERATIONS IN DESIGNING SLUDGE REMOVAL SYSTEMS.** In contrast to circular tanks, during low loading and low blanket conditions, the recycle flow concentration cannot be expected to be constant for traveling bridge scraper systems in rectangular secondary clarifiers, even under relatively even loading conditions. A large variation of recycle concentration can occur every time the system returns to start collection at the far end of the tank. Concentrated sludge is fed into the hopper only when the scraper blade is closer than 10 m (Ekama et al., 1997). With the chain-and-flight systems, however,

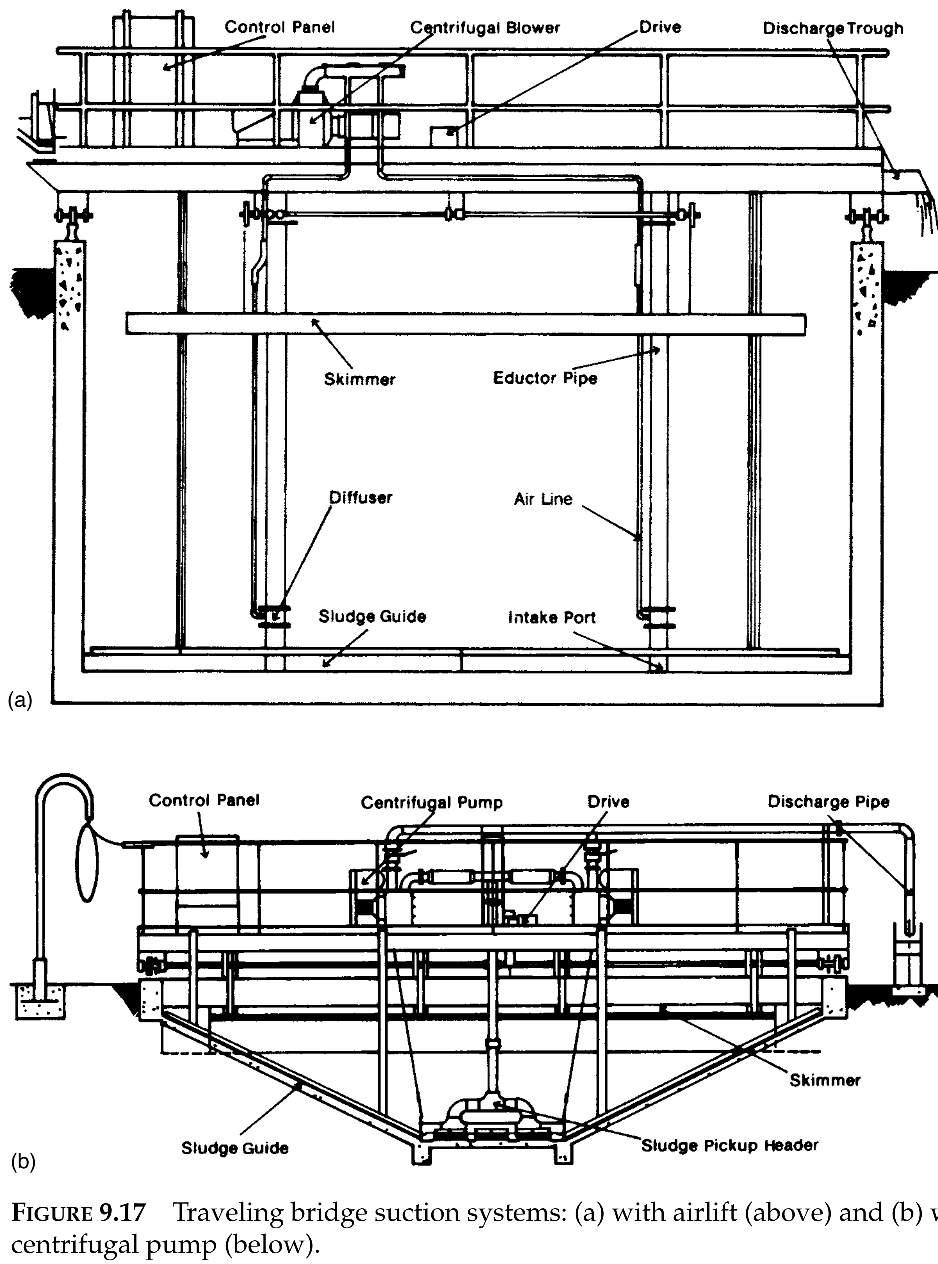


**FIGURE 9.17** Traveling bridge suction systems: (a) with airlift (above) and (b) with centrifugal pump (below).

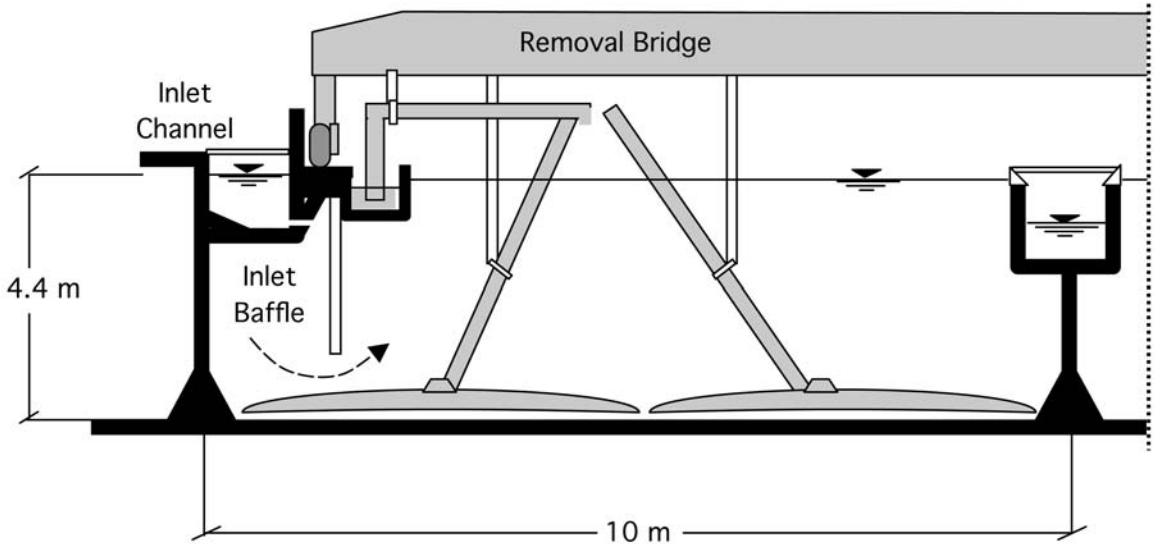


**Figure 9.18** Traveling bridge suction mechanism using differential head for sludge removal in transverse tank at wastewater treatment plant in Zurich, Switzerland.

the movement of sludge to the hopper is frequent enough to produce a relatively constant recycled sludge concentration.

The suction collector (also called an *organ pipe collector*) has a highly variable return activated sludge concentration. The variation depends on which part of the tank the suction collector is passing. A relatively high concentration is realized as a higher sludge blanket builds up just before the collector reaches the turning point at the end of a tank. However, hardly any sludge blanket is present on its return pass shortly afterwards. Also, the sludge suctioned near the end part of the tank is generally less concentrated than sludge removed near the influent end. Some investigators suggest that suction collection should be controlled by measuring the sludge concentration in front of the collector and adjusting the speed of collector or the recycle flowrate accordingly (Ekama et al., 1997). In a transverse tank, the concentration varies with the removal period, but is symmetric with regard to the turning points of the collector. One of the disadvantages of suction systems is that short-circuiting of the inflow to the sludge recycle can occur when the collector is near the inlet.

Because of unstable sludge flow conditions, caused either by the fluctuations in hydraulic head differentials or the variations in the sludge concentrations and characteristics, a suction sludge collecting system is not an ideal application for small clarifiers that are operated with a low sludge detention time. Furthermore, sludge collected by a suction system may have a lower recycle concentration than a conventional chain-and-flight system, which has the benefit of the thickening and conditioning effects because of the passage of multiple flights.

Theoretically, in an ideal flight system, the settled sludge layer is equal in height to the scraper's height, and the sludge is continuously moved toward the hopper with the correct velocity required for removal. This was the base assumption that Krebs (1991a) applied as a boundary condition in numerical simulations to result in a constant return sludge concentration. However, the actual conditions that occur near the scraper blade are much more complex. Baumer et al. (1996) observed flight passage through a glass side wall of a pilot-scale experiment and determined that it is the sludge in the region near the front side of the flight that is actually transported. A considerable part of the bottom layer of sludge is displaced vertically, where it can then be transported by the density current. A shear layer and a flow separation are produced at the upper edge of the flight, increasing the turbulence level locally, while also inducing a wave at the sludge surface. The energy of the scraper movement is, therefore, partly used for sludge lifting and partly dissipated via turbulence. It should be stated that the turbulence created by the removal mechanism does not seem to have adverse effects. The turbulence acts more like a gently stirring, which improves sludge settling and thickening. In contrast, the vertical wave at the surface of the sludge blanket exposes the sludge flocs to the density current and possible resuspension.

When sludge collectors move against the main direction of flow in longitudinal tanks, this often leads to hydraulic instability. A more stable flow pattern is established when the sludge collectors move concurrently with the main flow direction, for example, when the sludge hopper is placed near the end wall or in the intermediate region of the tank. The transport of the settled sludge with the bottom density current also is more efficient. It should be noted that this issue is not as pronounced in the suction system. Because sludge is removed where it settles by a suction system, the sludge is not transported to any other location in the tank.

## OUTLET CONDITIONS AND EFFLUENT REMOVAL

The two types of effluent collectors that are commonly used are surface launders (overflow weirs) and submerged launders (outlet tubes), which consist of collection pipes fitted with orifices. Use of either outlet design type for rectangular tanks must consider mitigation of the end wall effect caused by the deflection of the bottom density current. It should also address scum baffling and whether or not to provide outlet deflection baffles. If weirs are provided, the designer must decide on the location, orientation, and structural support of weirs and the weir loading rate. If submerged outlet piping is provided, the designer must decide on the location and orientation of the piping and the orifice size. There can also be a combination of submerged launders and overflow weirs, as provided by the Renton Treatment Plant in Seattle, Washington (WEF, 1998). These two types are provided for normal and very high peak flows, respectively.

**SURFACE LAUNDERS.** When weir loadings and surface loadings are low, the orientation and placement of weirs in secondary clarifiers are not critical, and they are perhaps even less critical for primary clarifiers. In longitudinal tanks, effluent surface launders can be oriented either longitudinally or laterally (transversely). Longitudinal launders placed directly on the clarifier sidewalls are by default single-sided, and those located some distance from the sidewalls can be single-sided, but are generally double-sided. Lateral launders located at the end of the tank are single-sided, and lateral launders located upstream from the end wall are typically double-sided. For typical rectangular clarifier weir configurations, see Figure 9.19. In transverse tanks, a single-sided launder is generally provided on the entire length of the outlet wall, which is the long side of the clarifier. This weir can be on the same side as the inlet (folded-flow pattern) or on the opposite side. Because the specific flowrate (flow per unit width) is much smaller in transverse tanks, a relatively low weir loading rate is obtained even with a single, one-sided weir.

**END WALL EFFECT**. Ideally, the clarified effluent should be removed after as much settling as possible has taken place, which corresponds to the area toward the end of the tank. However, unwanted hydraulic patterns, produced by the bottom density current in longitudinal secondary clarifiers, can be present in the region next to the end wall, especially with sludge hoppers placed at the end (see Figure 9.20). These density currents have a greater effect on activated sludge secondary clarifiers than on primary clarifiers and chemical sludges from tertiary clarifiers. The lightweight activated sludge floc may be lifted up, especially if it is a bulking sludge. This problem occurs when the sludge density current traveling down the clarifier is deflected upwards as it approaches the end wall. At the end wall, the solids tend to

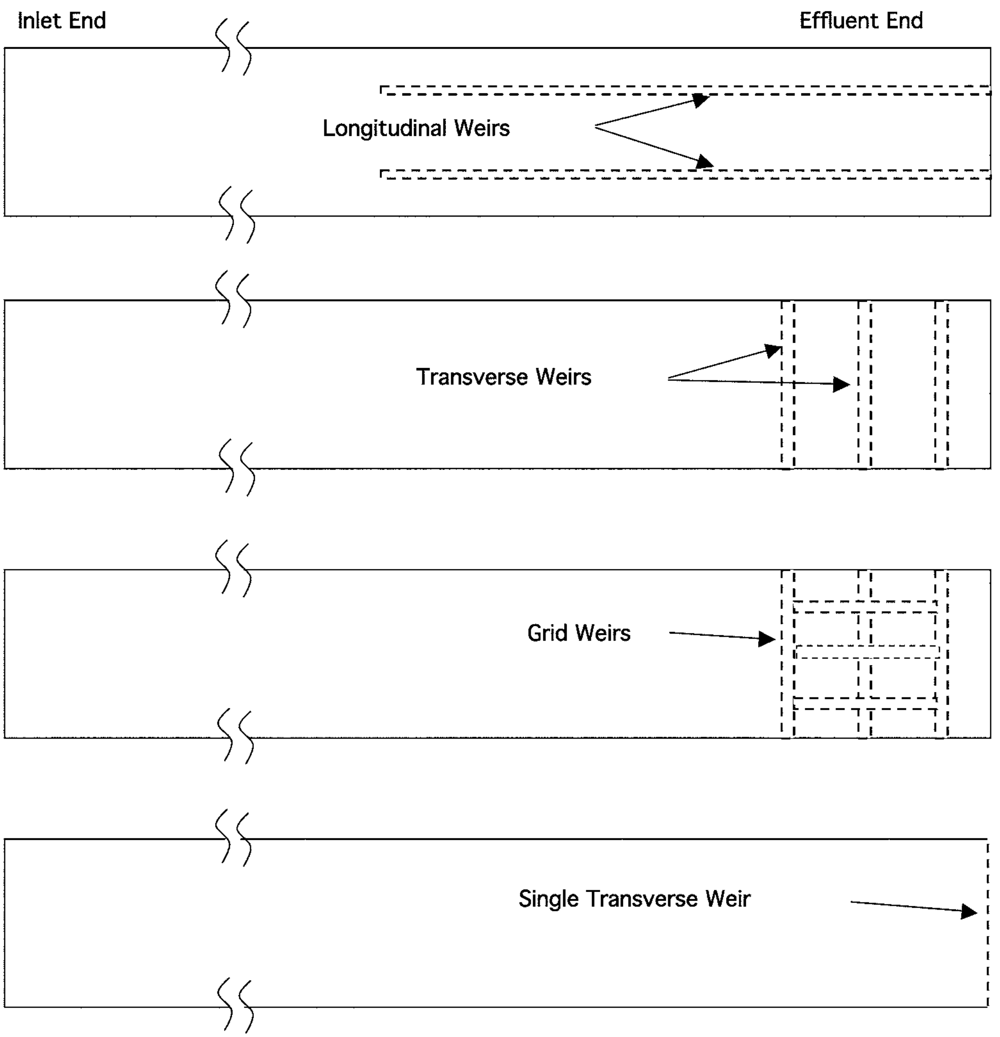


**FIGURE 9.19** Plan views of typical surface weir configurations.

mound up. Under severe loadings, the elevated sludge blanket can even be swept over the weir. Although the end wall effect and other eddy currents are the root causes for suspension clouds to rise in the clarifier, the suction effect of an effluent launder can start to remove suspended solids when flocs rise in close proximity or approximately a distance of 25 to 50 cm (10 to 20 in.) from the launder. The carryover effect also can be reduced effectively by limiting the horizontal velocities across the clarifiers to 20 mm/s (4 fpm), as indicated by Theroux and Betz (1959).

Because of the density current and the end wall effect, a longitudinal clarifier's effluent launder should not necessarily extend to the end wall. For the same reason, the last lateral launder should be an inboard (facing upstream) rather than an outboard weir (facing the end wall). Ideally, a well-designed outlet would remove effluent from a large portion of the tank and exclude the region close to the end wall. Anderson (1945) felt that the effluent weir should be moved upstream from the end wall of the clarifier to prevent the loss of floc from secondary clarifiers. Similarly, ATV (1991) suggests that, when a lateral launder is used, it should likewise be located a distance from the outlet wall of at least the water depth (see Figure 9.21). For existing facilities, the weir area near the end wall of a longitudinal clarifier can be blanked off with plates to avoid collecting effluent from that region of the tank. As a rule of thumb, the length of the weir blanked off at the end wall should be at least the depth of the tank. Alternatively, deflection baffles can also be installed below the weirs to deflect the upwelling caused by the density current.

In colder regions, short weirs may be preferable to long weirs, as these cause fluctuating water levels that will hamper ice formation. In locations where there is

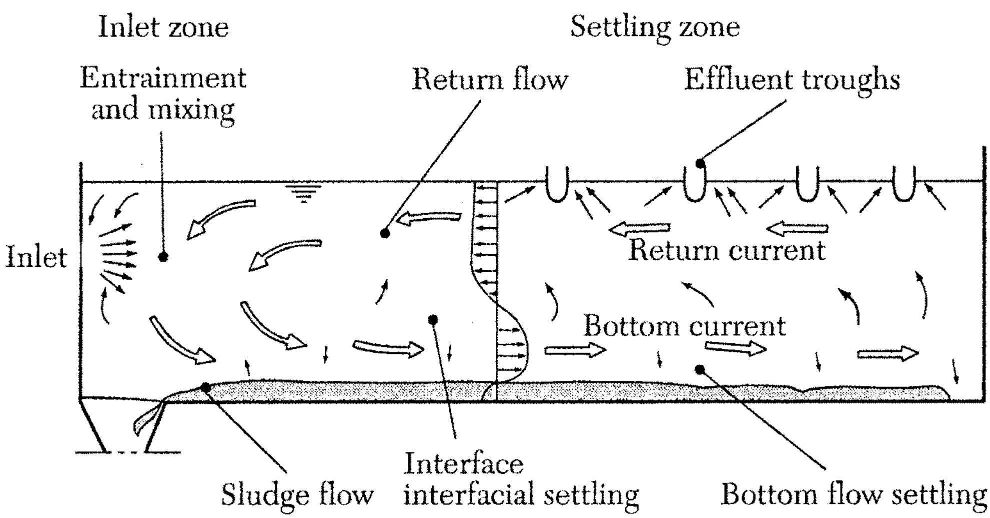


**FIGURE 9.20** Typical rectangular clarifier flow pattern showing the density current (Larsen, 1977).

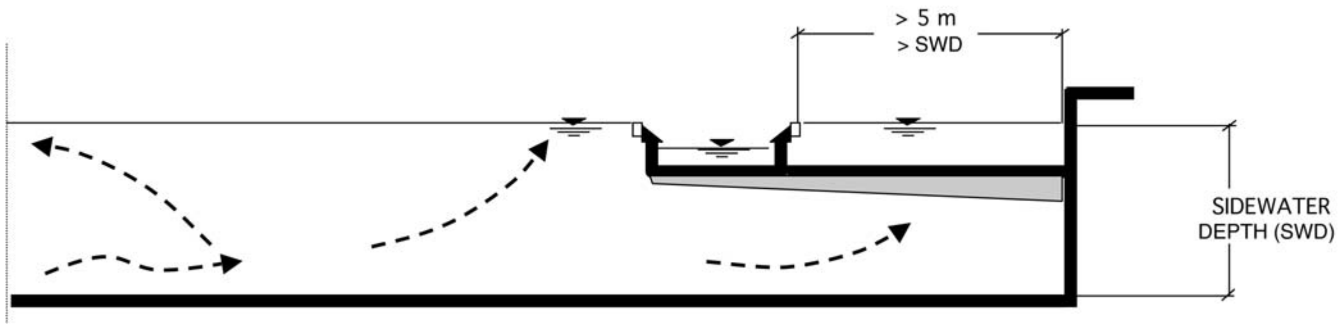


**Figure 9.21** Section view of recommended placement of transverse weir away from effluent end of longitudinal tank (reprinted from *Secondary Settling Tanks*, ISBN: 1900202035, with permission from the copyright holder, IWA).

susceptibility to wind, heavy cross winds on open tanks can easily set up a sloshing and surging of the water over the weirs. If the launders are fiberglass and not heavily braced, the long-term, back-and-forth movement of the weirs can lead to fatigue and premature failure. Even worse would be a wind-aided harmonic movement forming with fiberglass weirs that have no lateral support when the clarifier is dewatered and out of service. This can lead to catastrophic failure. To counteract the effects of wind, the launders can either be substantially braced, covered, or have the weir area provided with more freeboard to shelter it from the wind. Breakaway launders have also been designed so that structural damage to the tank will not occur, which is more costly and takes more time to repair.

Where bridge-type scraping mechanisms or floating suction-type units are used, these mechanisms are designed to pass between the outlet weirs or launders and, therefore, require longitudinal weirs to allow passage of the collector mechanism. Because flights are absent with suction mechanisms, longitudinal launders can then be supported from the clarifier floor. When weirs or launders are added to an existing basin, they may be cantilevered and spaced in the tank to obtain the necessary weir length. Many times, the launders are covered to prevent or reduce algae growth, to the chagrin of some operators (the ones that do not have to clean weirs) who want to see that part of the tank and assess the quality of the effluent.

Uniform withdrawal along the length of the weir is important to prevent localized regions of high velocity. Leveling of weirs within a tank and in relation to other tanks is a difficult, but necessary, task. Preferably, a slotted adjustable weir plate should be used so that it can be accurately set along its length and leveled with the outlet weirs of other clarifiers in parallel (IWPC, 1980). Because of leveling problems, v-notched weirs are preferred over straight-edged weirs (WEF, 1998). A v-notched weir is also used instead of a straight weir to prevent the unbalanced flow caused by wind action.

**WIER LOADING RATES.** Surface launders in longitudinal rectangular clarifiers are generally designed conservatively and to a maximum feasible length. For short to moderately long tanks, this may be up to 50% or more of the tank length. For longer tanks, parallel longitudinal launders that extend 20 to 30% of the tank length from the effluent end with a spacing of 3 m (10 ft) apart have worked well. For shorter and wider tanks, especially transverse tanks, a simple weir across the width of the tank may suffice. Extensive launder structures are sometimes not warranted, unless it is necessary to meet certain state design criteria or regulatory requirements. Excessive weir length requires more cost, effort, cleaning, and leveling. Weirs and weir toughs are designed so that they will not be submerged at maximum design flow. Boyle (1974) has described a method of the hydraulic design of weir troughs. Additionally, weir troughs are designed to maintain a velocity of at least 0.3 m/s (1 ft/s) at one-half the design flow to prevent solids deposition (Great Lakes-Upper Mississippi River Board of State Sanitary Engineers, 1978).

In relatively long and narrow clarifiers, evidence is lacking that weir loading rate (WLR) has any significant effect on primary clarifier efficiencies (Graber, 1974; Rankin, 1959). Because of this, many times, the simplest design solution for primary clarifiers is to provide a weir across the full width of the end of the tank, protected by a scum baffle. However, as the clarifier length is shortened, there is concern about the possible carryover effect of suspended solids resulting from the increased flow velocity at the weir. Wilson and Ballotti (1988) found that, with Gould—Type II tanks, the weir loading rate is of minor importance on the effluent quality. Furthermore, decreasing the effluent weir loading rate does not seem to prevent the loss of the floc in secondary clarifiers because of the end wall effect (McKinney, 1977). Deep tanks or tanks with baffles designed to deflect the density current also demonstrate that higher weir loading rates can be tolerated. By diverting the density currents from the effluent weirs, loading rates in excess of 520 $m^3/m \cdot d$ (40 000 gpd/ft) can be used for secondary clarifiers without sacrificing effluent quality.

In practical applications, the weir loading rates can vary within a range of 85 to 520 $m^3/m \cdot d$ (6500 to 40 000 gpd/ft) for secondary clarifiers. Weir loading rates for single-sided weirs are typically designed at approximately 250 $m^3/m \cdot d$ (6100 gpd/ft). To obtain more conservative weir loading rates, the engineer can easily double the weir length by providing weir plates on both sides of the effluent launders. For double-sided weirs, weir loading rates are typically lower, in the general range of 150 $m^3/m \cdot d$ (6100 gpd/ft) (ATV, 1991; Ekama and Marais, 1986; Kawamura and Lang, 1986).

Weir loadings are less critical than the criteria of overflow rate or the occurrence of flocculation, equal distribution, and the effect of the density current. In small tanks, the weir loading rates should be limited to 125 $m^3/m \cdot d$ (10 000 gpd/ft). In larger tanks within the upturn zone of an end wall effect, the weir loading rates should be limited to 250 $m^3/m \cdot d$ (20 000 gpd/ft). In larger tanks outside the upturn zone of an end wall effect, the weir loading rates can be limited to 375 $m^3/m \cdot d$ (30 000 gpd/ft). In any case, the upflow velocity in the immediate vicinity of the weir should be limited to 3.7 to 7.3 m/h (12 to 24 ft/h). Weir loading rates are stipulated by many state regulatory agencies; however, many times, these rates can successfully be exceeded (Tekkipe, 1986). Some outlet weir design data from various sources are shown in Table 9.2.

Because the weir loading rate, which is inversely proportional to the weir length, does not seem to have a very serious effect on the suspended solids removal, a balance between shorter to greater weir lengths should be reached. Too much weir length can create greater operational and maintenance problems, in terms of cleaning

**TABLE 9.2** Outlet weir design for rectangular clarifiers.

| Source | Design data |
|---|---|
| Process Design Manual, Suspended Solids Removal (U. S. EPA, 1975) | Daily average weir loading between 125 and 375 $m^3/m \cdot d$ (10 000 and 30 000 gpd/ft) (suggested for primary tanks as a comparison). |
| Ten States Standards (GLUMRB, 1978) | Weir hydraulic loadings of 180 $m^3/m \cdot d$ (15 000 gpd/ft) at design flows for large plants; 125 $m^3/m \cdot d$ (10 000 gpd/ft) for small plants. |
| (Hudson, 1981) | Discharge weirs should be used with maximum feasible length up to perhaps 50% or more of the tank length. Reasonable spacing between weirs up to a center-to-center distance of one to two times the tank depth. |
| (Sanks, 1978) | Size so average velocity at surface under the weir trough is 60 to 90 $m^3/m^2 \cdot d$ (1.0 to 1.5 gpm/sq ft) (empirical for water treatment sludge). |
| (Fair et al., 1968) | 375 $m^3/m \cdot d$ (30 000 gpd/ft) to avoid surges. |
| (Metcalf and Eddy, 2003) | WLR 125 to 500 $m^3/d/m$ (10 000 to 40 000 gpd/ft) with 250 $m^3/d/m$ (20000 gpd/ft) typical for primary clarifiers; WLR 125 $m^3/d/m$ (10 000 gpd/ft) at average flow and 250 $m^3/d/m$ (20 000 gpd/ft) at maximum secondary flow; WLR 250 $m^3/d/m$ away from upturn zone and 375 $m^3/d/m$ within upturn zone for secondary clarifiers. |
| Guides for the Design of Wastewater Treatment Work (NEIWPCC, 1998) | Adjustable overflow weirs with multiple weir troughs placed sufficiently far apart to avoid excessive upward velocity. |

and leveling. Flow characteristics are adversely affected if the weirs are not accurately leveled.

**SUBMERGED LAUNDERS.** As an alternative to surface overflow weirs, submerged launders or outlet tubes have been provided for effluent removal on many occasions. Sometimes, a better hydraulic design can be achieved with this method than with weirs. Submerged launders also eliminate the hydraulic problems that can arise because of differences in weir elevations. Gunthert and Deininger (1995) summarized the advantages of outlet tubes as follows:

- More uniform effluent withdrawal from the available surface area.
- Flow through the orifices under slight variations in static head is less sensitive than the surface effluent launders.
- Submerged outlet tubes thus maintain more uniform effluent distribution under windy conditions than do surface launders. This is less important in rectangular tanks than circular tanks because, although the water surface length exposed to wind in the longitudinal tanks may be greater when the wind blows in that direction, effluent withdrawal still remains symmetrical. In circular tanks, the wind effects are more likely to push flow to one end of the clarifier and cause asymmetrical conditions.
- Localized vertical velocities in the region of withdrawal are lower, and, therefore, the suction effect that could give rise to sludge clouds is decreased.
- Removal of floatables is reportedly easier because there are no obstacles like effluent launders on the surface. Also, a scum skirt is not required.
- Operational problems with algae growth are eliminated.

Submerged outlets can overcome some of the shortcomings of weir troughs, such as differential settling, short circuiting, and free-falling water, which can produce offsite odor problems (Lutge, 1969). However, sometimes these advantages are offset by the potential maintenance problems of rags or other items blocking the orifices and by fouling caused by biological growth in the tubes.

Submerged launders can be oriented longitudinally or transversely with respect to the tank, and submerged launders remove effluent from that part of the tank comparable to weir design. To counteract the end wall effect, the designer can decide not to have any outlet orifices in this area. Submerged launders are typically fitted with an automatic valve that controls the clarifier level or provided with a downstream weir that acts to control the water level (Tekippe, 1986). Typical arrangements of submerged outlet tubes are shown in Figure 9.22.

The outlet tubes should be designed for uniform withdrawal of the effluent. The velocity and head loss of the flow through the orifices into the outlet tubes should ideally be equal. The loss through the orifices should, therefore, be higher than the losses through the tube, resulting in a smaller orifice diameter and a larger tube diameter. By establishing a significant loss at the orifices, submerged collectors can also be used as the principal source of head loss required for uniform influent flow distribution among clarifiers operated in parallel. Submerged tubes are fitted with orifices for the specific flowrates that are expected. However, provision of a larger tube diameter allows for flexibility for increased capacity if orifices can be resized larger.

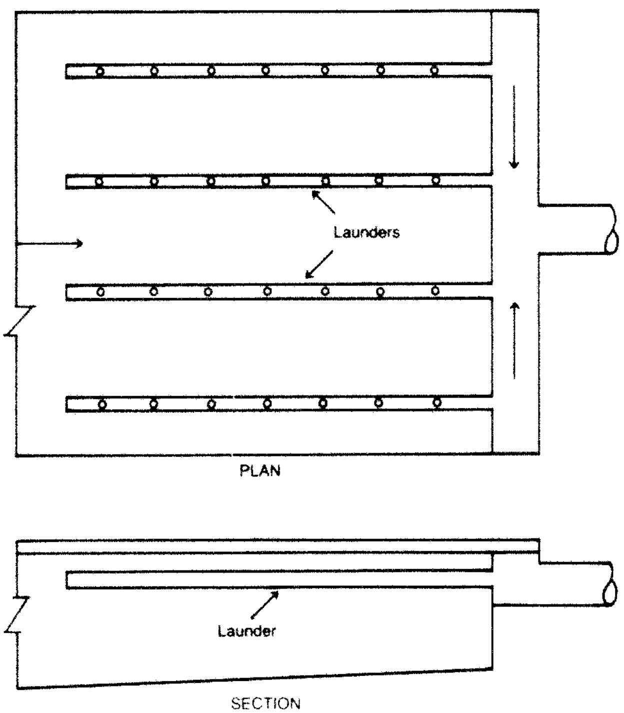


**FIGURE 9.22** Submerged launders consisting of pipes with equally spaced orifices.

Analysis of a number of head losses serves as the basis for design for the outlet tube system. These losses include the loss resulting from the confluence of flow to the orifices, the loss through the orifice, and friction losses through the tubes. The design should consider the following practical criteria developed by ATV (1995) and Gunthert and Deininger (1995). Although applicable to circular tanks, the following also apply to rectangular tanks:

- Orifice diameters should be in the range 25 to 45 mm (1.0 to 1.75 in.);
- Maximum velocity at the tube exit should be approximately 0.6 m/s (2 fps);
- Velocity range through the orifices should be 0.6 to 1.0 m/s (2.0 to 3.3 fps) under peak wet weather flow conditions; and
- Tubes should be supported at least every 10 m (33 ft).

Submerged launders allow scum to be concentrated conveniently at the far end of the tank instead of at an intermediate position upstream of weirs. With submerged outlets, the variation in water level under various flow conditions needs to be considered when designing the scum removal systems. This water level control can be accomplished by providing an overflow weir in a short channel section downstream of tubes. In larger plants, or plants concerned about large peak flows, submerged launders with automatic level control valves can be used. To prevent floatables from entering the submerged outlet tubes, the tubes are typically placed 30 to 35 cm (12 to 14 in.) below the water surface. Because the water layer above the tubes cannot be regarded as part of the clear water zone, the entire water depth of the clarifier should be increased to some degree over that of a conventional clarifier with effluent launders (Ekama et al., 1997). However, the required deeper depth is a significant cost adder in construction.

## REMOVAL OF FLOATABLES

In general, scum removal receives too little attention in clarifier design. This often translates into too much attention being required by the operator after construction. With increased implementation of BNR systems and the concomitant longer mean cell residence times (MCRTs), Nocardia-type foam is often unavoidable. In addition, sometimes denitrification can translate into unwanted scum production. Startup foams and occasional influent surfactant loading is also a consideration. Water sprays, hypochlorite sprays, or cationic polymer addition can help to keep the foam under control if operational changes are unable to do so. On occasion, removal of the foam upstream of the clarifier is preferable and performed with surface removal devices in the aeration tanks or distribution channels. These locations are areas where selective wasting can be practiced to control the population of foaming organisms. However, in most plants, skimmings generally end up as a sidestream flow and are recycled to an upstream part of the plant. In some larger plants, it is economical to concentrate the skimmings and remove them from the main influent flow stream. Some plants pump the concentrated skimmings to digesters for further treatment.

The collection and removal of floatables in the clarifier chiefly concerns both the inlet and outlet areas of the tank, because the foam generally passes through the middle reaches of the tank unimpeded. Generally, the floating material will pass onto a skimming device in another part of the tank. If influent baffles are used, they need to be slotted on the surface to allow scum to pass through to another part of the tank. Otherwise, foam will collect behind the baffle, and some sort of provision will have to be made to remove the scum separately in that location. A provision can be made for a small gate to be opened and pass scum when necessary.

Toward the outlet end of the tank, a scum baffle projecting from above the water surface and downward into the tank should also be provided to prevent scum from escaping into the weir region of the clarifier. The distance between the scum baffle and the outlet weirs should be 60 to 150 cm (2 to 5 ft). The depth of submergence should be 30 to 60 cm (1 to 2 ft) to prevent scum from escaping under the baffle and passing out with the effluent. Scum baffles generally are constructed of steel, aluminum, or fiberglass, and are fitted with stainless steel bolts.

In a rectangular clarifier equipped with flight-and-chain mechanisms, skimmer equipment is generally provided to remove accumulated scum and consists of a slotted roll pipe situated across the end of the tank. It is positioned at the point where scum is concentrated by the movement of the sludge flights that return on the surface of the clarifier. The slotted roll pipe is really a type of adjustable weir. The horizontally mounted pipe is partially submerged in each tank wall, with the slot extending the full length of the pipe. Bushings on the pipe allow rotation of the slot from an inactive position, where the slot is facing up above the water level, to a position where the slot is below the water level. In the latter position, scum actively flows through the slot and is generally removed from the tank by gravity flow. The pipe is rotated manually as needed for a few minutes, whenever it is required to remove the accumulation of scum from the tank. The scum passes through the wall into a scum box outside the tank, where it can be pumped elsewhere or perhaps flow to a plant sewer. Other designs use a separate short set of chains and flights or spiral flight arrangement to move the skimmings up a beach and into a trough. For rectangular clarifiers that are equipped with traveling bridge mechanisms, a scum remover can be mounted to the bridge itself. The scum remover can push the floatables in either direction and into a special scum trough.

Alternatively, pan-type skimmers are also used on rectangular clarifiers. This type of skimmer is generally assisted by water sprays. The sprays direct the scum to the smaller influence area of the pans (relative to the slotted pipe). Skimmings flow generally moves through the pan, down a flexible tube, in some cases, and then into a skimmings line positioned below and running transversely through a bank of clarifiers. Skimming pans can be fixed, adjustable, or the floating type. Fixed pans allow for skimming at one elevation and, therefore, are set for high-flow conditions. Adjustable pans can be manually adjusted a few times a day for different flow conditions. Floating pans are self-adjusting, but generally do not allow an operator to adjust for the quantity of skimmings to be removed, as conditions may change. Water sprays can be used to assist the skimmings towards the pan and prevent setup of the foam to keep it flowing.

Finally, it should be mentioned that, in activated sludge systems heavily laden with Nocardia, too much aeration of distribution channels can create a foam problem in the distribution channel and the clarifiers. Sometimes, this can be used as an advantage for selective wasting purposes. However, it should be realized that some sort of limited aeration is necessary in the distribution channel to help prevent the foam from setting up and sticking to channel walls. Water sprays in the distribution channel can assist to move foam when this limit is met. If foam is allowed to accumulate for a number of days, odor problems will occur. Sometimes, skimmer flights or a scraper blade are provided for distribution channels.

Water sprays deserve a certain amount of design consideration. For sprays in the clarifier, a nozzle that develops large droplets over a large area, such as a fan-type nozzle, is effective in moving scum to pans and beating a certain amount of foam back down into the water. Fine sprays that develop a lot of mist should be avoided for health reasons. Narrow jets that have limited area of influence should also be avoided. For moving thick foam through distribution channels, jet-type nozzles directed tangentially on the opposite side of the aeration downcomers are effective in freeing up the foam to travel downstream to a removal device. Finally, for each different type of nozzle used, the effective range, nozzle location, and direction of spray should be well-established empirically before final installation. If not, the capacity and required horsepower of the foam spray system can easily be wasted over a long period of time.

## INTERNAL TANK BAFFLES

To further improve clarifier conditions, internal baffles can be considered to enhance settling and provide a more clarified effluent. Internal baffles are modifications of the basic rectangular tank design, which seeks to enhance sedimentation in the region between the inlet and outlet zones. Two types of baffles have been investigated in rectangular tanks (solid and perforated), and the effects of each type of baffle can be entirely different. In addition, each of these baffle types can be sized differently or possibly configured in series. In general, because the density current is the dominating hydraulic factor in the longitudinal rectangular clarifiers, the effects of the baffles depend on their interaction with and the mitigation of the density current. It should not be overlooked, however, that much of what initially happens concerning the density current depends on the inlet structure. Finally, it should be stated that, for internally baffled tanks, the removal of sludge presents a major challenge.

**SOLID BAFFLES.** Solid internal baffles can be designed to have either a low or high profile in relation to the depth of the tank. As expected, both types of solid baffles will act as barriers to the density current and will store sludge to varying degrees. While a relatively low baffle acts as a flow barrier to stop or slow down the bottom current, it does not drastically alter the tank operation. A high solid baffle is one that extends higher than midwater depth and can even extend to a height just under the water surface. This essentially divides the clarifier into two basins in series and can dramatically change the clarifier operation (Ekama et al., 1997).

Barrier baffles have been tested by Bretscher et al. (1992) in rectangular tanks. When clarifier loading is typical (e.g., dry weather loading), a clarifier with a low solid baffle can still be treated as a single unit, because the compartments before and after the baffle will still interact. Under these conditions, a low solid baffle is relatively effective in stopping the bottom current. Although investigation is lacking on how this system performs under wet weather or dynamic loading, it is reasonable to assume that the beneficial effects of a solid baffle would vanish as the loading increases. This is because the barrier effect can cause a deflection of the flow stream or density current, and clouds of relatively high solid concentration could rise to the surface. Under increased loading and without proper solids removal, the sludge blanket upstream of the baffle could even increase to the point that it overflows into the second compartment. Depending on the height of the baffle, this could result in the formation of an even greater density current in the second part of the tank, resulting in deterioration of effluent quality (Ekama et al., 1997).

With a high barrier baffle, sometimes called a *dividing baffle*, the operation of the first part of the tank is totally separate from the operation of the second part of the tank. The first part of the tank acts like a highly loaded clarifier. The effluent of the first part becomes the influent for the second part, so that the second part of the tank performs more like a lightly loaded clarifier

Krebs (1991b) used numerical simulations to demonstrate the function of barrier (low baffles) and dividing (high) baffles and simulated their failure mechanisms (Figure 9.23). The simulated baffles were one-half and three-quarters of the water depth and were analyzed for dry and wet weather conditions. The analysis showed that, under dry weather conditions, the low baffle acts as a brake to the bottom current, and only a small amount of solids are transported to the second part of the tank. Under similar dry weather conditions, the flow stability and performance of clarifiers are enhanced by both the low and the high baffle designs. However, in wet weather conditions, the relatively high loading causes the sludge blanket to rise in the first part of the tank and overflow the low baffle to the second part of the tank, resulting in a pronounced density current. Because the latter part of the tank is designed for light loading, ultimately a sludge washout is predicted. With the higher baffle in wet weather loading, the sludge blanket surface does not reach the upper end of the dividing baffle. However, because of the decreased flow area above the baffle, the flow constriction above the high baffle can increase the velocity to the point where it causes short-circuiting with a direct surface current to the effluent launders. To prevent short-circuiting along the surface and to mitigate the density current in the second part of the tank, sometimes another baffle (similar to an inlet baffle) is provided downstream of the dividing baffle (Ekama et al., 1997).

Finally, it should be mentioned that a solid dividing baffle system should only be applied to an operation in which the variation of loading is small. Dividing baffles will only be successful if the first part of the tank has sufficient storage volume for all the sludge that is shifted from the aeration tank during peak loadings. For this reason, it is recommended that the first part of the clarifiers be designed with a greater depth than typical design practice. This, of course, drives up construction costs. Furthermore, this type of operation has the effect of propagating an elevated sludge blanket for long periods of time, something a great number of operators try to

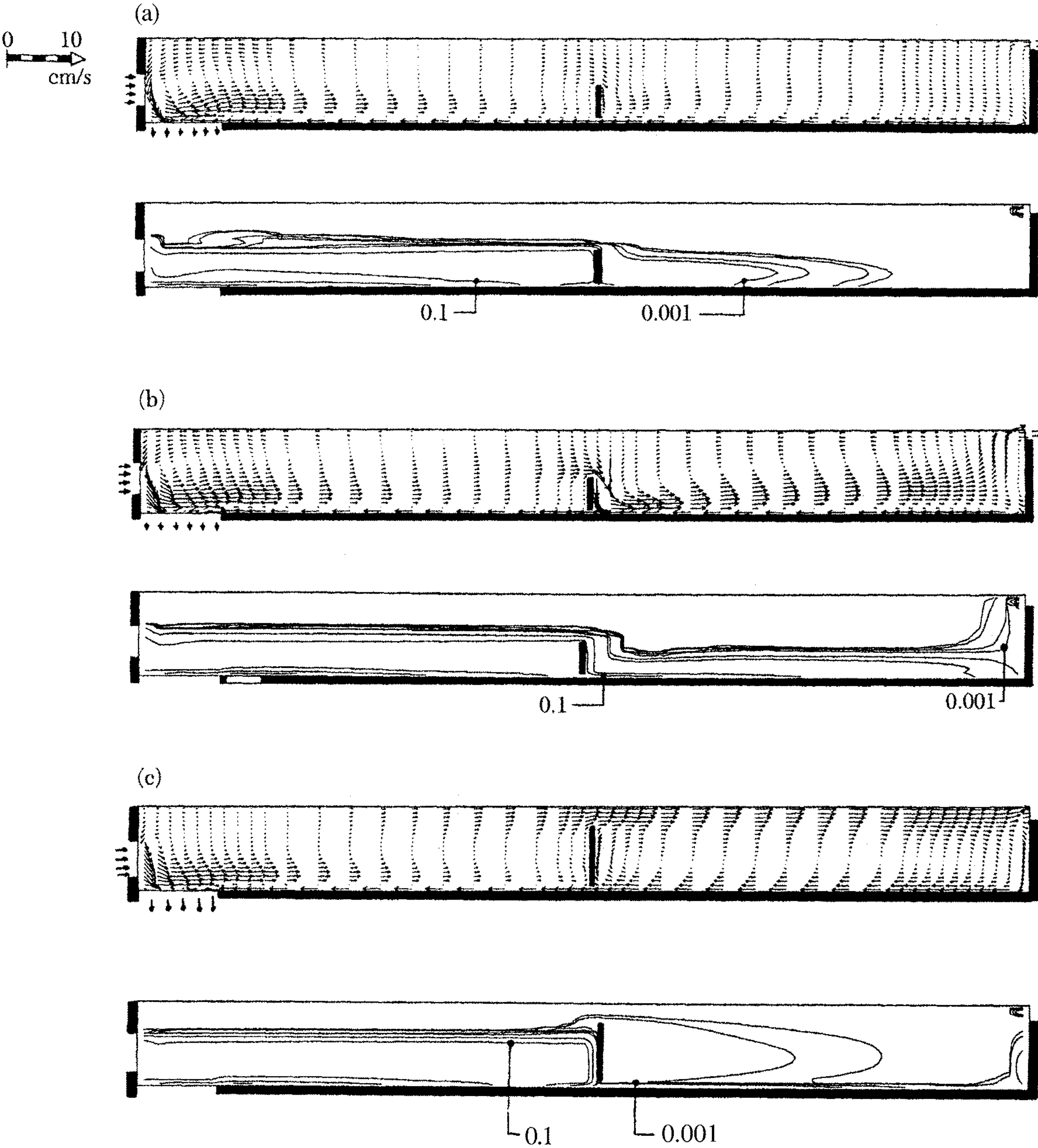


**FIGURE 9.23** Numerical simulation of three conditions with a dividing baffle (Krebs, 1991b). Upper drawing is the flow field (velocity); lower drawing is the volume fraction (solids): (a) dry weather load with baffle height at 50% of tank depth, (b) wet weather load with baffle height at 50% of tank depth, and (c) wet weather load with baffle height at 75% of tank depth.

avoid. Because of the long solids rentention time and possible depletion of oxygen, closer attention must be paid to the effects of denitrification. These can cause problems with rising sludge and inhibit thickening (Ekama et al., 1997)

**PERFORATED BAFFLES.** Perforated baffles were originally used for inlet distribution. However, they have since been adapted for use farther downstream in the tank. A single perforated baffle can be provided or a number of perforated baffles can be used in series. Well-designed perforated baffles have precise openings to constrict and distribute the flow across the entire flow cross-section of the tank. This allows the sludge solids and clarified liquid to have continuous and direct interaction between the tank partitions. Even though the sludge solids distribution is slightly affected, a tank with perforated baffles can still be regarded as a single unit, unlike a highly loaded clarifier with solid baffles (Ekama et al., 1997).

Kawamura (1981) performed hydraulic laboratory tests with perforated baffles to reduce the density current effect. The perforated baffles acted as a flow barrier, while also being pervious, and, overall, helped to increase flow distribution over a greater depth. Okuno and Fukuda (1982) used several perforated baffles to obtain an even flow velocity distribution after the inlet. Esler (1984) used slotted baffles located at the midlength of rectangular clarifiers. In later projects, he used two or more cross baffles. Esler and Miller (1985) used horizontal wooden bars of approximately 50% of free cross-section to improve the effluent quality by approximately 40%. Watanabe and Fukui (1990) showed that perforated baffles have the potential to enhance flocculation. In a laboratory experiment with an artificial flocculent suspension, they attempted to optimize a tank with several baffles serving as a flocculation reactor. Krebs et al. (1992) combined hydraulic laboratory experiments with a theoretical approach to optimize the design of perforated baffles with regard to flow distribution and flocculation enhancement in the sedimentation region. Baumer et al. (1996) tested the flocculation enhancement effects of perforated baffles in pilot-scale experiments with activated sludge (Ekama et al., 1997).

***Single Perforated Baffles.*** The use of single perforated baffles will have the following effects:

- If the porosity is chosen correctly, the head loss of the baffle will be slightly greater than other factors that may be causing uneven flow patterns. The flow pattern will then become more uniform and ideally proceed through the entire clarifier cross-section.

- Flow through the perforations in the baffle in the proper velocity range can induce shear layers. These will result in low-level turbulence that can enhance flocculation after the baffle.
- The depth of the sludge blanket will be increased in the chamber upstream of the baffle because of a slight barrier effect. Because the sludge blanket height after the baffle will be lower, a slight density current may be induced after the baffle.

As can be seen from the discussion above, the porosity fraction of the perforated baffle is extremely important. Krebs et al. (1992) investigated optimization of the porosity fraction and formulated the following guidelines:

- The jets of flow developed through the holes must be turbulent at minimum loading.
- The mean velocity gradient produced by the jets must be low, but still in a flocculation-enhancing range.
- A minimum hole diameter of 5 cm (2 in.) should be used to prevent clogging.
- The hole diameter should not exceed 10 cm (4 in.), because the individual jets persist horizontally and thereby effectively reduce the sedimentation volume.

A porosity fraction of 0.05 (or 5% open) is typically found by using these guidelines for longitudinal flow clarifiers. This coincides with the smaller porosity fractions successfully tested by Okuno and Fukuda (1982).

***Multiple Perforated Baffles.*** The problem of a fast-moving density current, which develops following a single baffle, can be mitigated by using several perforated baffles. This results in a stepped sludge-storage profile (see Figure 9.24). When the difference in sludge blanket height in each succeeding chamber is reduced, the density effect after each baffle is minimal. Additionally, the sludge concentration decreases from the first chamber to the last chamber, such that very little sludge mass ends up in the last chamber. In essence, it should be noted that a tank with several perforated baffles is partitioned into a sequence of flocculation chambers and sedimentation volumes. This is important to realize, as the fraction of volume used for flocculation should be limited to allow enough space for the more controlling factor of sedimentation. The absolute lengths of the flocculation zones are dependent purely on the flow velocity and the diameter of the perforations.

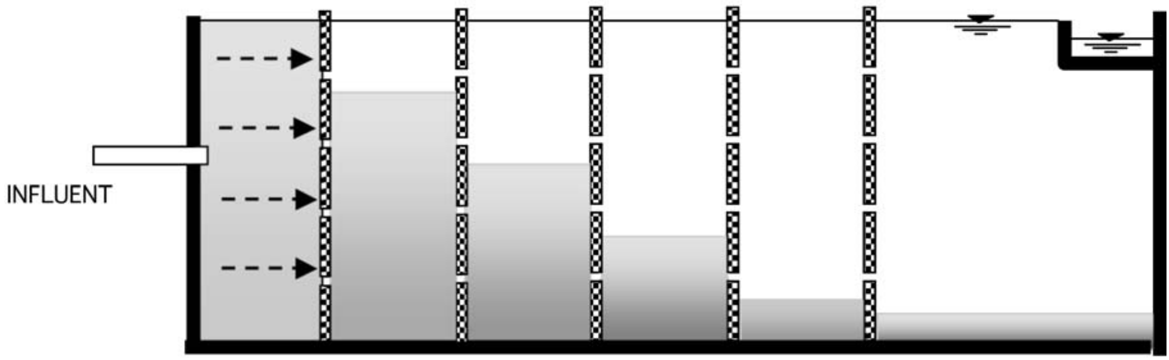


**FIGURE 9.24** Stepwise sludge distribution in a clarifier with several perforated baffles in series (reprinted from *Water Science & Technology,* **26** (5/6), 1147–1156, with permission from the copyright holder, IWA).

Baumer et al. (1996) compared a clarifier system with three perforated baffles to the original tank without baffles. The ESS concentration of the original tank was adversely affected by an overload condition after four hours of testing, while the tanks with the perforated baffles proved more stable with respect to dynamic loading. Okuno and Fukuda (1982) reported similar success with the testing of perforated baffles with porosity fractions of 0.40 and 0.05 in the inlet region. Even though the tank was overloaded according to the ATV guidelines, they reported that the perforated baffle system was resistant to a shock load and sustained ESS values lower than 20 mg/L. This was because short-circuiting from the inlet to the outlet was avoided. An estimate of the sludge mass showed that more sludge was stored in the tank with perforated baffles. The system allows sludge to be stored in the first chamber without causing a high sludge blanket in the vicinity of the effluent launder at the end of the tank (Ekama et al., 1997).

**SLUDGE REMOVAL WITH INTERNAL BAFFLES.** Internal baffles present major challenges with respect to sludge removal. Although providing a tank with baffles is theoretically advantageous for effluent quality, given the increased complexity of sludge removal systems, the installation of baffles becomes less practical. From the standpoint of hydraulics alone, longitudinal tanks are best suited with perforated baffles. However, in longitudinal tanks provided with perforated baffles, it is difficult to install conventional sludge removal systems. Because the direction of sludge removal is generally the same as the main direction of flow, this means these

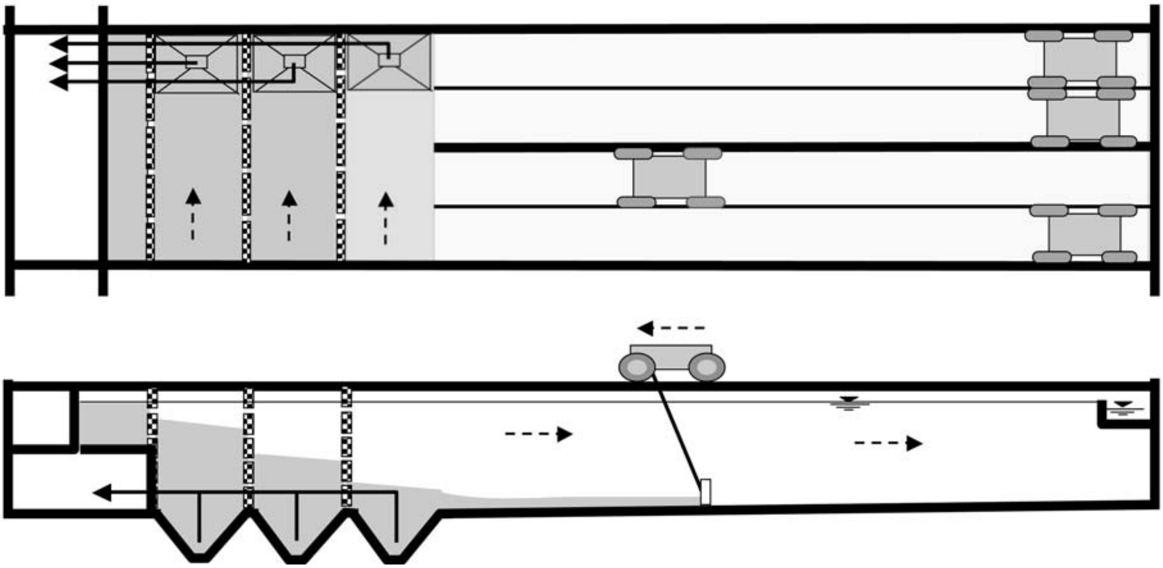

**FIGURE 9.25** Multiple perforated baffles with transverse sludge removal (Okuno and Fukuda, 1982).

collector systems somehow need to pass through the internal baffles or separate removal must be provided for each compartment. Figure 9.25 shows multiple perforated baffles with crosscollectors leading to individual sumps and a piped removal.

When a longitudinal tank is retrofitted with perforated baffles and the existing removal system is to remain, passage through the baffles must be made possible. Flexible rubber flaps that allow a suction removal collector to pass through perforated baffles have been constructed, in which the flaps close off after the collector has passed (Baumer et al., 1996). Rubber flaps were also tested successfully in a pilot-scale experiment of a longitudinal clarifier (Baumer, 1996) with a flight scraper system.

Because of the above problems in adapting typical sludge removal systems for perforated baffle systems and taking into account that sludge concentrations in the chambers are reduced in stepwise manner from the inlet end to the outlet end chambers, a better solution would be to remove the settled sludge separately from each chamber. This means that the suction collectors would have to be provided for each chamber. Alternatively, Okuno and Fukuda (1982) provided transverse scrapers in each chamber of a baffled longitudinal tank. They determined that the optimum baffle should be positioned 12 and 25% down the length of the tank from the inlet. By providing different sets of transverse and longitudinal sludge collectors, this eliminates the need to allow the passage of sludge removal mechanisms through openings or flaps. Large openings can negate the effects of the baffle and can even make a tank perform worse because of eddy currents. Additionally, moveable flaps essentially become another maintenance item, and their failure would not be readily apparent until after the tank is dewatered.

## STACKED CLARIFIERS

The concept of stacked clarifiers was originally proposed by Camp (1946). This design alternative is appropriate in locales where land for treatment facilities is not available or extremely expensive. Stacked clarifiers consist of hydraulically connected settling tanks located one above the other. The stacking effect essentially increases the clarifier surface area without increasing clarifier facility footprint. They are also called *tray clarifiers* and can be double-decked or even triple-decked. In this way, the footprint can be decreased by a factor of two or three, respectively. At the Deer Island Treatment Facility in Boston, Massachusetts, use of stacked primary and secondary clarifiers has reportedly reduced the required footprint of those facilities by 40% (Lager and Locke, 1990).

Stacked clarifiers were first constructed in Japan in the early 1960s. Because of space constraints, rectangular clarifiers have been stacked two or three deep at numerous facilities in Japan. Osaka City has operated stacked facilities with satisfactory performance for more than 20 years (Yuki, 1990). In the United States, stacked clarifiers were first constructed at the Mamaroneck, New York, treatment plant in 1993, and have been constructed in Salem, Massachusetts, at the South Essex Sewerage District. In the Mamaroneck facilities, stacked primary sedimentation tanks are designed for 350 000 $m^3/d$ (92 mgd), with design overflow rates of 22.3 $m^3/m^2{\bullet}d$ (550 gpd/sq ft) at average flow, and twice that overflow rate for peak flow (Kelly, 1988). Recent treatment plants incorporating stacked clarifiers include the Ulu Pandan and Changi East plants, both in Singapore and the Stonecutters Wastewater Treatment Plant in Hong Kong. In general, overflow and weir rates are similar to conventional rectangular and secondary clarifiers (Metcalf and Eddy, 2003).

There are two types flow regimes for stacked clarifiers: series flow and parallel flow. In the less common series flow unit, wastewater enters the lower tray, flows to

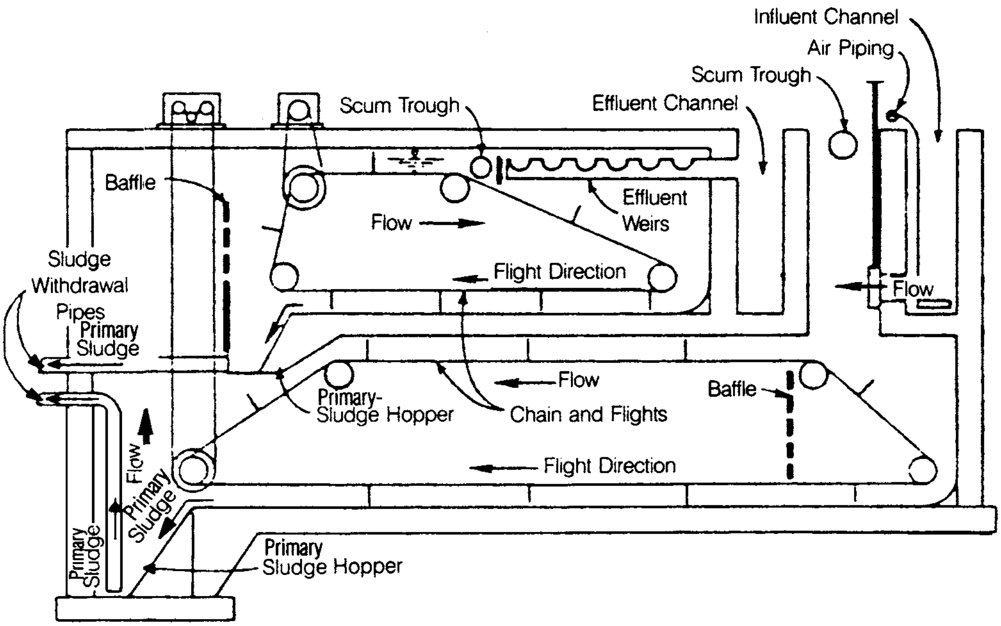


**FIGURE 9.26** Stacked rectangular clarifier—series flow type (from Kelly, K. [1988] New Clarifiers Help Save History, *Civ. Eng.*, **58,** 10, with permission from the American Society of Civil Engineers).

the opposite end, reverses direction in the upper tray, and exits the effluent channel (Figure 9.26). Baffles are used to straighten the flow path and minimize turbulence at the influent point in the lower tray and at the flow reversal point on the top tray. Effluent is generally removed by longitudinal launders located along the top tray (Kelly, 1988).

The parallel flow regime is the most common stacked clarifier configuration. In the parallel flow unit, influent piping conveys wastewater or mixed liquor from an influent channel to inlet pipes for each tray (Figure 9.27). To ensure equal distribution of solids to the both levels, the influent channel should be aerated, or otherwise mixed, and the inlets to each deck should be taken from the same elevation of the influent channels. Influent baffles are used in each tray to straighten the flow path and minimize turbulence.

Effluent can be removed via effluent weirs or perforated pipes. Weirs should be located at the same elevation for each deck. At the South Essex Sewerage District in

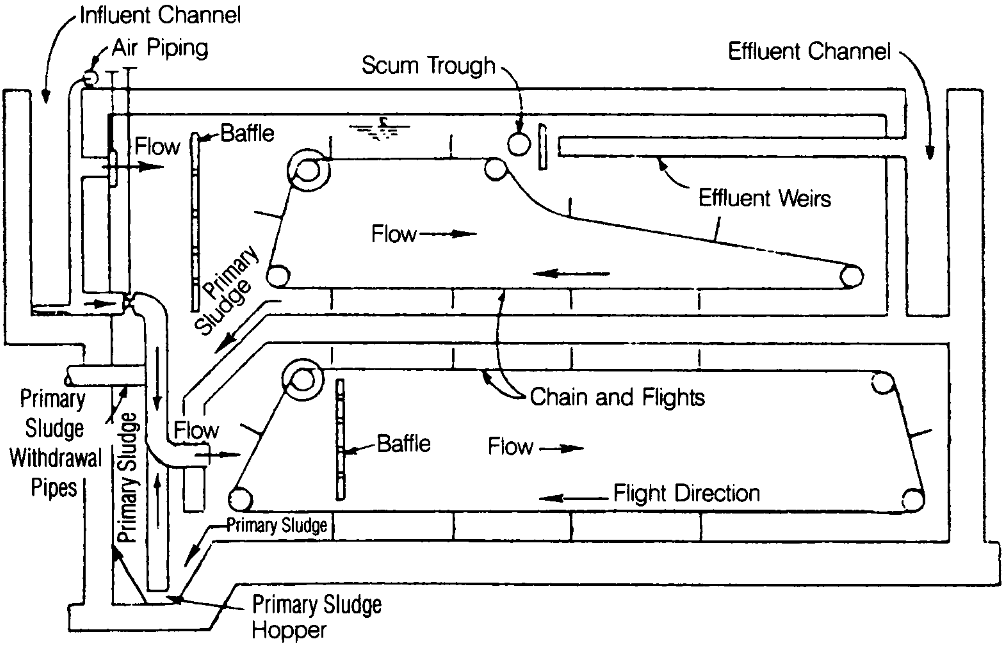


FIGURE 9.27 Stacked rectangular clarifier—parallel flow type (from Kelly, K. [1988] New Clarifiers Help Save History, *Civ. Eng.*, **58,** 10, with permission from the American Society of Civil Engineers).

Salem, Massachusetts, a single effluent weir is used at the end of each tray, rather than inboard effluent launders. Figure 9.28 shows an example of a single transverse effluent launder for each tank at the same elevation. In another available configuration, longitudinal launders remove effluent from the individual tanks. Similarly, if perforated pipes are used, they should discharge to a common water surface for flow distribution purposes. Yuki et al. (1991) described applications and arrangements for perforated pipes. In Japanese facilities with parallel-flow-type clarifiers, equal flow distribution between the upper and lower trays is often controlled at the outlet rather than the inlet.

Chain-and-flight mechanisms are used for sludge collection and removal from stacked tanks, because traveling suction devices for the lower tray would not be possible. Similarly, because the lower tray(s) is submerged, scum is only removed from the top tray. The inlet and outlet design in stacked tanks is challenging because influent flow patterns can possibly intersect the flow patterns of the primary sludge

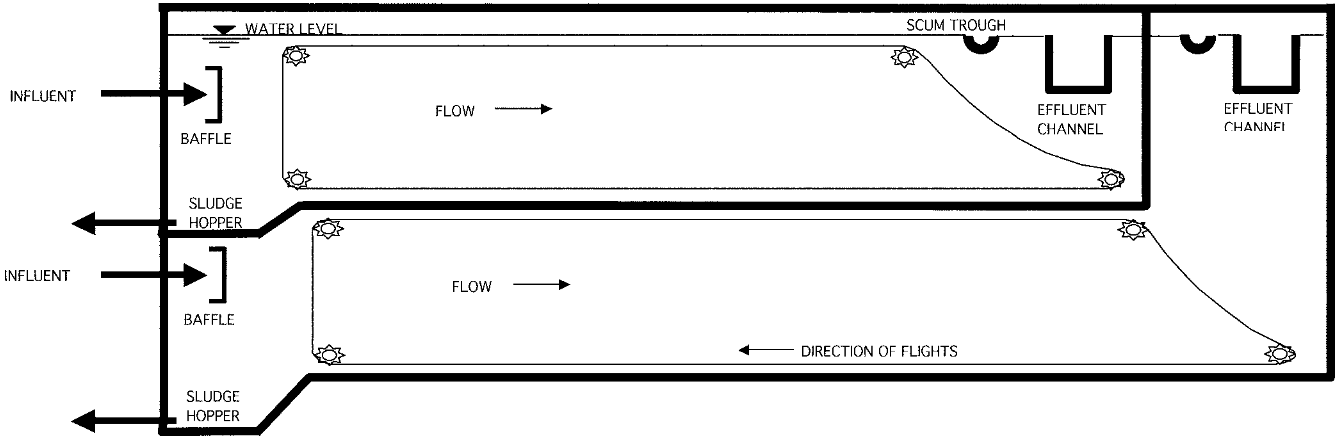


**FIGURE 9.28** Stacked rectangular clarifier—parallel flow type showing double-sided weirs at same water surface elevation.

(Lager and Locke, 1990). Figure 9.28 illustrates a common arrangement for removing sludge—a single hopper at the head of the lower tank, with sludge from upper tanks being dropped via chutes into the hopper. In this arrangement, influent flow must be piped past the sludge drop area. At the Ulu Pandan plant in Singapore, each tray has its own hopper, rather than having sludge drop from the top tank to the hopper in the lower tank.

All stacked clarifiers with hoppers have sludge drawn by sludge scrapers to the inlet end. This arrangement is not optimal for secondary clarifiers in activated sludge systems. As discussed elsewhere in this manual, better performance can be achieved if sludge is withdrawn from the effluent end of the tank or from a transverse hopper halfway or farther down the tank, as in a "Gould-type" clarifier. The arrangement of the stacked secondary clarifiers at the Changi East Plant in Singapore provides for sludge withdrawal halfway down the length of the clarifiers. There, the scrapers move solids to a transverse perforated pipe located in a midpoint hopper for each tank. The perforated-pipe arrangement also allows for a flat tank bottom.

Because of their more centralized design, stacked clarifiers generally require less overall piping, and, therefore, pumping requirements are also less. If covers are required for odor control, there is less exposed surface area to cover. However, they do suffer from more complex structural design and construction costs. The stacked configuration will generally result in a deeper structure, require more excavation, and require close attention to the buoyant effect caused by local groundwater conditions when stacked tanks are taken out of service. Another disadvantage is that any operational observation of the lower tray is precluded because of its underlying position. Also, maintenance of the lower tray is more difficult and subject to confined space entry requirements.

Ducoste et al. (1999) evaluated two different double-stacked clarifier designs using computational fluid dynamics (CFD). The first design used a single inlet with a passive hydraulic mechanism designed to split the influent flow between the upper and lower portions of the clarifier. Settled solids in the upper portion were conveyed by gravity to a hopper in the lower level. The second design incorporated individual flow control to actively split the influent flow between the upper and lower levels of the clarifier. Settled solids in the both levels were independently removed at the longitudinal midpoint of the clarifier by a slotted pipe. Both designs were served by individual effluent weirs provide at each level.

Results of the CFD showed that the first design suffered entrainment of flow by the inlet in the lower level, which could cause resuspension of settled solids. This problem could be solved by reducing the flow into the lower level of the clarifier or by extending the lower inlet pipe further into the tank. The second design showed that having a separate sludge-collection system made it a superior design, because the operation of the upper level did not affect the lower level. Finally, the positive flow control the second design provided resulted in better use of the tank geometry. Overall, the study showed that CFD evaluations of different stacked clarifier designs could be used to quantitatively determine whether the improved performance of one design was worth the relative increase in cost and complexity compared to another design.

## MATERIALS OF CONSTRUCTION AND EQUIPMENT SELECTION

**MATERIALS OF CONSTRUCTION.** For large municipal plants, rectangular clarifier walls, floors, and galleries are made of concrete. Generally, these do not require coating. For smaller plants and industrial applications, coated steel is used more extensively. A common coating specified for carbon steel that gives excellent corrosion protection has been cold tar epoxy, although, in some states, this coating has been restricted by volatile organic compound regulations. Nevertheless, some pinholes in application or scratches during construction occur. Further protection with this coating can be achieved by the installation of galvanic or impressed current cathodic protection systems. Cathodic protection systems are actually used in a very small percentage of clarifiers but have been used for industrial applications or waters with high salinity or low pH.

Galvanized steel can also be used and has been used with greater application in Europe than in the United States. A disadvantage of this coating is that it can be scratched or damaged in transport and in installation. However, if adequately protected and installed, it does offer good resistance to corrosion.

Internal mechanisms have most commonly been constructed using coated steel and sometimes aluminum; however, stainless steel and fiberglass have been gaining market share because of their abilities to avoid corrosion. The use of stainless steel has increased in recent years to provide additional corrosion protection for circular clarifier mechanisms. Either stainless steel type 304 or 316 can be used for this purpose. The latter is more expensive, but offers better corrosion protection in some installations. Hardware and fasteners are typically type 316 SS.

**Table 9.3** Common materials for chain-and-flight systems (adapted from information obtained from the NRG Products Web site (www.nrgco.com).

| Clarifier component | Materials/types | Comments |
|---|---|---|
| Flight assemblies | Redwood, glass reinforced polyester | Flights can be designed to have lips, be lipless, or can be fitted with wiper blades. Wooden flights are rectangular. Plastic flights are generally channel or diamond shapes. Special flight attachments compensate for twisting or misalignment. |
| Collector or carrier chain | Steel, stainless, glass reinforced polyester | Plastic chain will stretch and require frequent tightening or link removal. |
| Wear strips on carrying and return tracks | UHMW-PE, stainless steel | Plastic wear strips provided with slotted holes and beveled ends for thermal expansion. Stainless is smooth enough not to require wear strips. |
| Shafting | Steel, stainless steel | Stub shaft designs eliminate need for shafts. |
| Drive chain | Steel, stainless steel, cast iron, glass reinforced nylon | Two drive chains generally driven off one motor. |
| Return track assemblies | Steel, stainless steel, FRP track with UHMW- PE or PP brackets | |

*(continued on next page)*

Fiberglass and plastics have also increased in popularity as a construction material for rectangular tanks. Fiberglass or glass composite is commonly used for flights, weir plates, and scum baffles. It is also used widely now for construction of flocculation baffles. Walkways of clarifiers in small tanks are sometimes made of fiberglass. Aluminum tread plate and grating can also used in walkways and for baffles. Redwood flights are still common in many older facilities. However, with certain logging restrictions, the more rot-resistant grades of redwood are becoming more expensive. Fiberglass is nonbuoyant, will not retain water, and is one-third lighter than redwood. Although the wood alternative makes it easy to work with, wooden flights are

**Table 9.3** *(Continued)*

| Clarifier component | Materials/types | Comments |
|---|---|---|
| Driven sprocket | Cast iron, nylon, polyurethane, UHMW-PE, or UHMW-PE on cast iron or steel body | Standard or chain saver rims |
| Chain tighteners | UHMW-PE or nylon sprocket on nylon body | Can be used as tighteners or idlers |
| Stub shafts | Steel, stainless, cast iron, nylon | Alternative to long and heavy steel shafts and wall bearings. Adjustable horizontal and vertical alignment. |
| Collector sprocket | Cast iron, nylon, UHMW-PE, or UHMW-PE on cast iron or steel body | Solid or split designs and segmental designs Polyurethane or UHMW-PE on cast iron or polyurethane bodies. |
| Flight wear shoes | Metal, UHMW-PE, teflon, nylon, polyurethane | Replaceable pads that are positioned on the flight at the locations where it contacts the wear strip and the return rail. |
| Set collars | Cast iron, UHMW-PE, and nylon | Generally split and held in place by set screws or band clamps. |
| Take-up bearings and wall bearings | Metal with babitted or UHMW-PE lined inserts | Solid, split, and peak cap types. Self aligning with manual or automatic take-ups. |

Note: FRP—Fiberglass reinforced plastic; UHMW-PE—Ultra heavy molecular weight polyethylene; PP—polypropylene.

heavy and cumbersome and are largely replaced by manmade materials. Plastic components are more chemically resistant, are approximately 15% lighter than metal, have a lower frictional component, have superior wear and abrasion resistance, and reduce power and assembly costs. Some common materials for chain-and-flight systems are given in Table 9.3.

Shear pins are an important design feature of a rectangular clarifier, or any system that uses gear reduction to slowly move chain-driven mechanisms, such as flight collectors. Enormous torque can be developed, even by small horsepower motors, and one can easily find all the flights in a clarifier in a mangled condition if the wrong shear pin is inserted. Shear pins can be steel, aluminum, or plastic. Educating plant personnel about the proper shear pin material and size is the key to avoiding catastrophic failures. Shear pin breakage by itself does not signal a failure or overload condition, and it is necessary to have some sort of trip cam to alarm the failure or shut down the drive motor.

It is not uncommon for large banks of rectangular clarifiers to be provided for high-purity-oxygen activated sludge (HPOAS) systems. The covered reactor design, in conjunction with the pH depression resulting from an accumulation of carbon dioxide from respiration, makes corrosion of the concrete from carbonic acid likely, especially with an active wastewater with a high total dissolved solids or industrial waste contribution. For these applications, different concrete mixtures that are more acid resistant are recommended. In addition, the density of the concrete can be increased with little expense by lowering the water to cement ratio. This decreases the permeability of the concrete and helps to resist chemical and acid attack (Morton et al., 2000). If this is not done initially, a corrosion-resistant cementitious mixture can be shotcrete onto existing concrete surfaces needing repair. These construction and repair methods and a HPOAS reactor carbon dioxide stripping operation were used by the Los Angeles County Sanitation Districts's (LACDS's) Joint Water Pollution Control Plant (JWPCP) facility to combat corrosion in the HPOAS reactors and clarifiers and to prevent corrosion in the associated tunnel and outfall system (Pettit et al., 1997). In these situations, lined ductile iron piping is preferable to steel piping. Iron, brass, and aluminum should be avoided and substituted with stainless steel, fiberglass, and plastic.

**EQUIPMENT SELECTION.** The most costly equipment items for rectangular clarifiers are the sludge collection equipment, drives, and valves. Well-written specifications are the key to get a complete and workable system. The vendor's experience record and recent customer satisfaction is important in paring down the list of acceptable manufacturers for bidding purposes. Specifications for rectangular tank equipment are often categorized according to functional performance, structural loading of the equipment, mechanical design of the components, electric motors, controls and alarms, materials, and coatings for corrosion protection.

***Drives.*** The drive units for rectangular tanks generally contain speed reduction gearing, which transduces the high speed of the motor into the low speed and high torque necessary to allow the drive chain to rotate the collector sprocket. Helical bevel and worm gears, cycloidal speed reducers, and cogged gears have been used by the different manufacturers. Bearings are extremely important components of the drive mechanisms. The principal types include one in which steel balls run on hardened strip liners set in cast iron, and the second involve forged steel raceways. Jaw clutches are used for drive separation of two units driven by the same motor. Generally, there is some sort of mechanical protection provided for high torque conditions. These come in the form of shear pins, trip cams, and limit switch indication and electrical overload protection.

***Chain and Flights.*** Iron or steel chain can be used at most facilities. However, plastic chain is lighter and easier to work with. The plastic chain can sometimes cause increased preventive maintenance because of more wearing and stretching, which requires that links be taken out regularly. In some cases, the higher capital cost of stainless steel chain in primary tanks can save a lot of maintenance problems later.

***Valves.*** Because there are typically multiple inlet valves per rectangular tank, this can translate into a great many valves to operate and maintain. One of the most important considerations of operating and maintaining valves is exercising the valve. Frequent exercise will prevent deposits from interfering with valve operation. This is especially the case when there is a large amount of chemical addition upstream and phosphate removal. At large facilities, side mounted nuts will allow easier exercising of valves by use of a portable motorized stand. It is not uncommon for a valve stem to be bent by a well-meaning operator using a cheater bar on a stuck valve. Because a clarifier cannot be isolated with a faulty valve, this may eventually require a shutdown of an entire bank of clarifiers to replace one faulty valve stem. Ideally, some electromechanical means is provided for frequent exercising of valves. This may mean providing something other than the usual hand wheel to operate the valve.

# CASE HISTORIES

**INCREASE IN LENGTH OF RECTANGULAR CLARIFIERS.** The County Sanitation Districts of Orange County (CSDOC) operates a pure-oxygen activated sludge facility at the Wastewater Treatment Plant No. 2 in Huntington Beach, California. A process capacity evaluation was performed in 1990 that rerated the secondary treatment capacity at 3.5 $m^3/s$ (80 mgd), although it was originally designed for 4.0 $m^3/s$ (90 mgd) (Narayanan et al., 1997). The downrating was a result of area-based limitations of the clarifiers (related to surface overflow rate and solids loading rate) and hydraulic limitations resulting, in part, from the end wall effect and density currents, leading to high upflow velocities in the effluent withdrawal zone. The plant has twelve rectangular clarifiers, which are each provided with three side-by-side chain-and-flight collectors that transported sludge to an inlet hopper (see Figure 9.29). Because of available land constraints, additional clarifiers could not be built.

During the evaluation phase, the flow velocity in the weir area, or the effluent upflow velocity (EUV), was analyzed, and was thought to be more critical than weir loading rate. It was calculated that the theoretical EUV for particles in the 25 to 2000 μm range would be 1.1 m/h (3.5 ft/h) for an SOR of 1.1 m/h (635 gpd/sq ft) at a flow capacity of 3.5 $m^3/s$ (80 mgd). However, the existing EUV was really 7.6 m/h (24.8 ft/h) based on the same flowrate, indicating that a large fraction of resuspended flocs were being carried away in the effluent by this high velocity. Although blocking off the effluent launders at the end wall was considered, this would have further restricted the effluent withdrawal zone and increased the EUV. Also, the weirs only extended 7.6 m (25 ft) from the end wall. Blocking off the weirs a length equal to the water depth would have also substantially increased the weir loading rate. Instead, a decision was made to extend the 18.3 × 53.4 m (60 × 175 ft) clarifiers by 15.3 m (50 ft). By increasing the length of the clarifiers 68.6 m (225 ft), the withdrawal area was increased from 140 to 420 $m^2$ (1500 to 4500 sq ft) per clarifier, which decreased the EUV by a factor of three, to a value of 2.5 m/h (8.3 ft/h). (Parenthetically, it should be noted that this effectively also decreased the WLR, solids loading rate (SLR), SOR and made the tanks have more longitudinal characteristics (ATV, 2000). Modifications to extend the clarifiers were completed in 1995. Before the modifications, the ESS ranged from 9 to 21 mg/L. Following the modifications, the ESS ranged from 7 to 13 mg/L. The correlation between EUV and ESS is shown in Figure 9.30.

**RETROFIT OF MIDLENGTH HOPPER CROSSCOLLECTOR TO MANIFOLD SUCTION HEADER.** The Metropolitan Wastewater Treatment Plant (Metro) in St. Paul, Minnesota, is operated by the Metropolitan Council Environmental Services (MCES) and is the largest point discharge into the Mississippi River. With a design capacity of 11.0 $m^3/s$ (250 mgd), the plant is a conventional activated sludge plant that achieves seasonal nitrification and has 24 rectangular clarifiers, each with a midlength hopper. Each tank was equipped with three parallel chain-and-flight collectors oriented along the longitudinal axis of the tank (see Figure 9.31). The upstream and downstream collectors operate in parallel to the main direction of

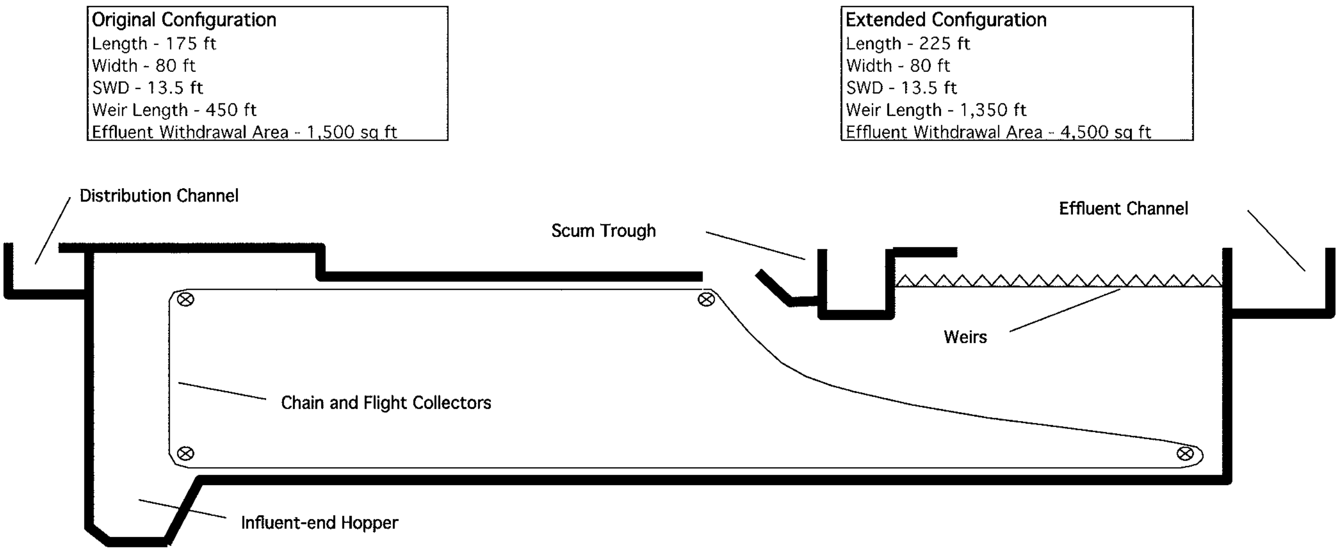


FIGURE 9.29 County Sanitation District of Orange County's plant no. 2 clarifiers with original and extended configuration dimensions shown (ft × 0.304 8 = m, sq ft × 0.092 9 = $m^2$) (Narayanan et al., 1997).

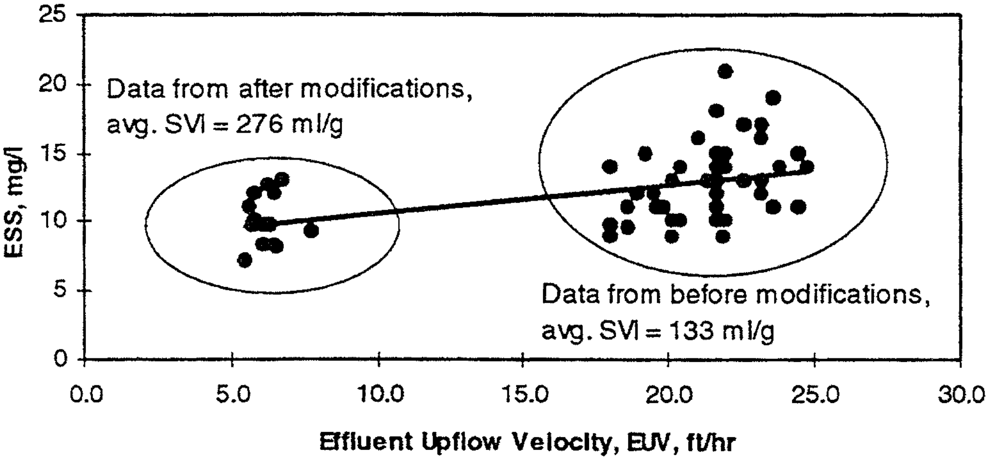


**Figure 9.30** Correlation between EUV and ESS at CSDOC plant no. 2 clarifiers (Narayanan et al., 1997).

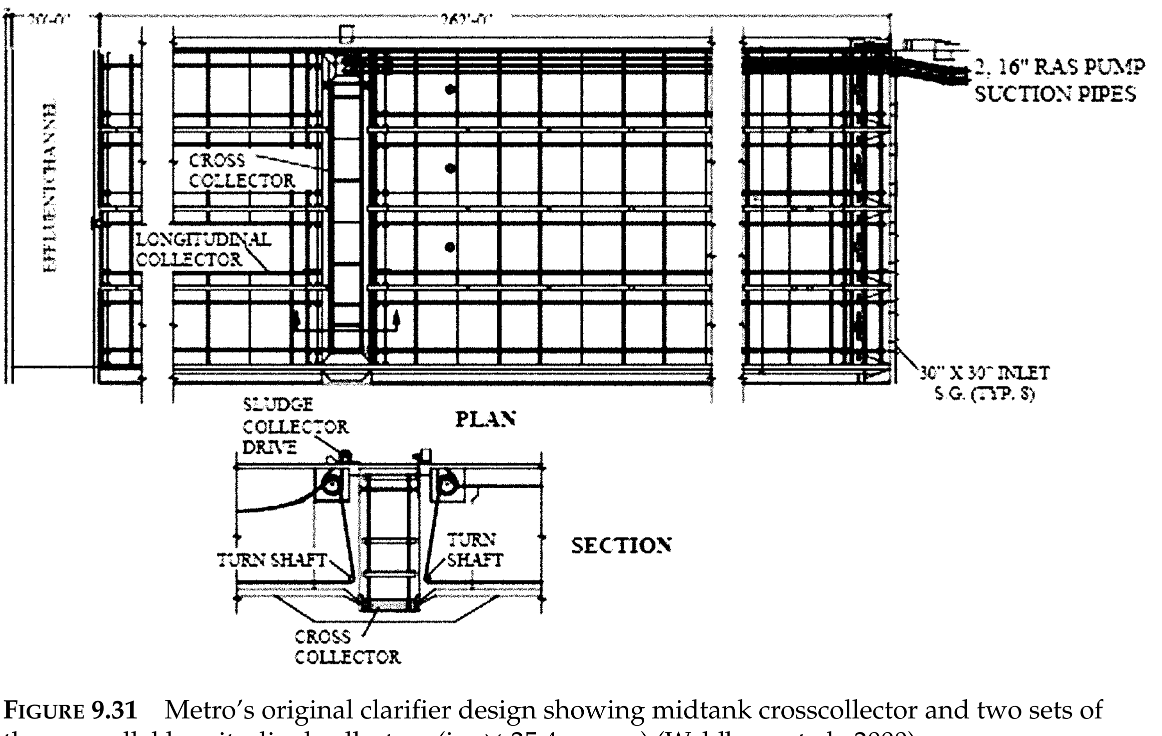


**Figure 9.31** Metro's original clarifier design showing midtank crosscollector and two sets of three parallel longitudinal collectors (in. × 25.4 = mm) (Wahlberg et al., 2000).

flow, while the hopper collector operates in transverse fashion, as a crosscollector to push sludge solids to a side hopper. As part of an evaluation to switch to a biological phosphorus (Bio-P) operation, MCES also decided to evaluate the clarifier operation (Wahlberg et al., 2000). This consisted of sludge settling, flocculation, and compaction testing to compare operation of the biophosphorus (Bio-P) sludge with fully and partially modified clarifiers. It was thought that the crosscollector area contributed to resuspension of flocs because of the drop in floor elevation at the midlength hopper and by transverse currents induced by the collector and asymmetrical sludge flow. Also, limitations in sludge transport capacity could lead to higher blankets. After resuspension, the end wall effect induced solids carryover at the effluent weir closest to the end wall.

Modifications included improving the distribution of the mixed liquor to the clarifiers, enhancing flocculation at the inlet of the clarifier, and promoting hydraulic stability by providing for a uniform flow profile. This was done by providing new mixed liquor energy dissipation baffles, new inlet baffles, and a new sludge withdrawal header, respectively. The mixed liquor energy dissipater baffles decrease the upward momentum of the flow entering the channel, which affected the proximal clarifiers. The new inlet baffle arrangement was actually two baffles. The first was a distribution wall that had four 7.5-cm vertical slits to produce 5 cm (2 in.) of head loss at peak flow. The other acted as a distribution wall to break up the vertical jets from the first baffle and distribute them horizontally (see Figure 9.32). Both baffles acted in concert to allow the inlet gates to be opened fully, which decreased shear of the floc across the gates and also promoted flocculation in their immediate area. The sludge withdrawal header actually consisted of two pipes with orifices that faced down into the tank, 10 cm (4 in.) off the floor. The evenly spaced orifices promoted an even flow profile in the tank.

The new suction header system with the Bio-P operation was able to decrease blanket average by 0.15 m (0.5 ft). The data showed that the existing nitrified activated sludge system using the existing crosscollector without the new baffles had an ESS of 11.6 mg/L, with a blanket average of 1.25 m (4.1 ft). The Bio-P sludge system provided with the new baffles and the existing crosscollector had an ESS of 9.5 mg/L, with a blanket average of 1.16 m (3.8 ft). Also, the Bio-P sludge system provided with the new baffles and the new sludge header existing had an ESS of 5.6 mg/L, with a blanket average of 1.07 m (3.5 ft). A state-point analysis was performed to show that operating in the Bio-P mode increased the clarifier capacity by 35% because of the 7% higher settling rate of the Bio-P sludge alone. However, flocculation tests that were done to remove the hydraulic effects showed that each of the sludge systems had the

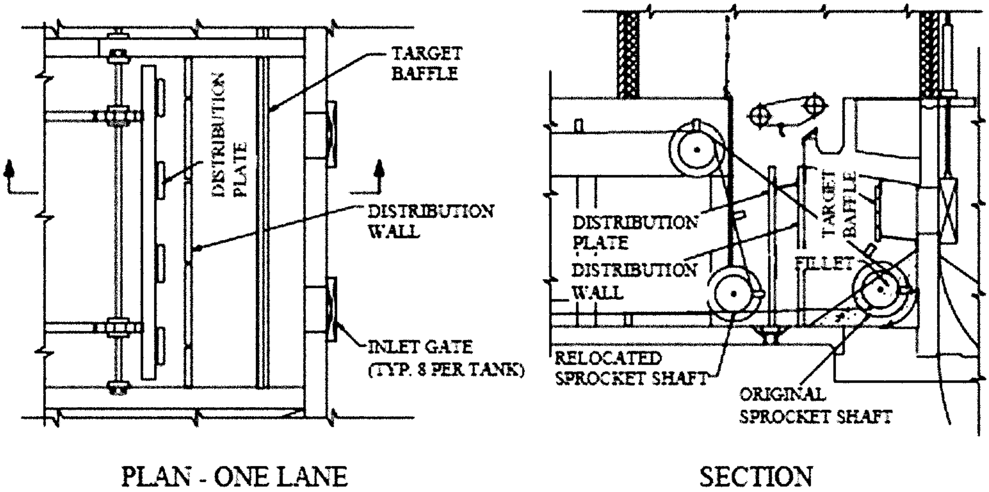


**FIGURE 9.32** Inlet modifications of Metro's existing clarifier—distribution plates for horizontal distribution of jets from vertical slots of distribution wall (Wahlberg et al., 2000).

potential to provide the same effluent. Therefore, the improved quality effluent was attributed to modified clarifiers with both the newly provided baffles and sludge withdrawal header.

**COMPARISON OF SHALLOW RECTANGULAR CLARIFIER WITH DEEP CIRCULAR TANKS.** A clarifier investigation was performed by LACSD at their San Jose Creek Water Reclamation Plant (SJCWRP). This is a 4.4-$m^3$/s (100-mgd) facility that, at the time of testing, was being run in a step-feed mode with reaeration of the sludge. Aeration is accomplished with fine bubble diffusers. Following secondary treatment, the effluent is disinfected and treated to tertiary standards with gravity filters. The clarifier testing followed the CRTC protocol established by the Clarifier Research Technical Committee of the American Society of Civil Engineers (Wahlberg et al., 1993). This included stress testing at target SORs from 0.7 to 3.4 m/h (400 to 2000 gpd/sq ft) over four hydraulic retention times (HRTs). The test clarifier was 46 m (150 ft) long by 6 m (20 ft) wide by 3 m (10 ft) deep. The rectangular clarifiers are sloped at 1% and equipped with chain-and-flight collectors performing cocurrent sludge removal at the effluent end into two hoppers (see Figure 9.33).

Excellent clarifier performance was exhibited during the stress tests under a wide range of sustained flows. The data showed a slight increasing linear relationship

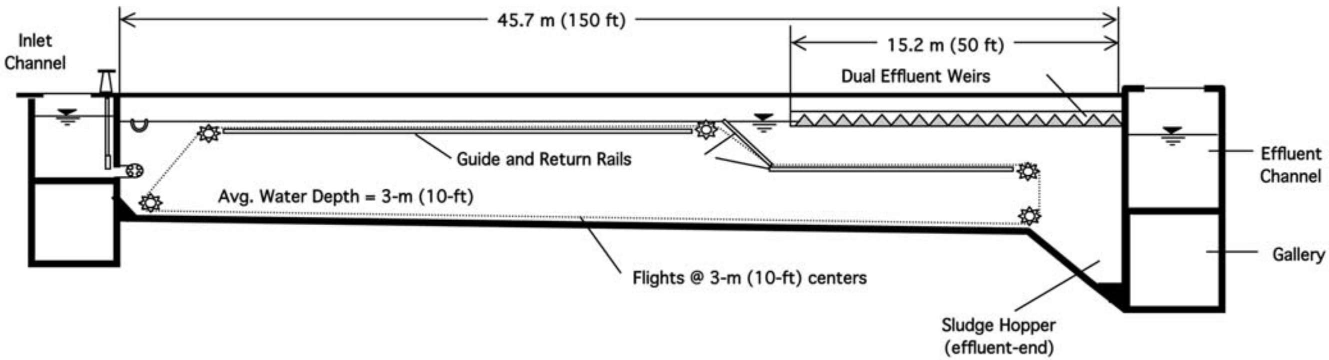


**FIGURE 9.33** Longitudinal section view of shallow clarifier with effluent-end sludge hopper at San Jose Creek WRP and operated by Los Angeles County Sanitation Districts (from Stahl and Chen, 1996).

between ESS and SOR (see Figure 9.34). Although flocculation and settling characteristics were determined not to be large influences on the ESS, the dispersed solids averaged a relatively low 5.6 mg/L (range of 2.2 to 10.2 mg/L) during testing. Flow pattern/distribution tests (dye and suspended solids concentrations) that were performed indicated favorable hydraulic characteristics and distribution after passing through the inlet diffusers (see Figure 9.35). Because of the limited range of attainable recycle flow through one tank (25% recycle at an SOR of 3.4 m/h or 2000 gpd/sq ft), it was not possible to test even higher SOR targets without a sludge washout. High blankets may have contributed to the higher ESS during the higher SOR conditions, although favorable results were still obtained, even though the blankets were relatively high compared to the overall depth of the tank (see Figure 9.36). Even in these conditions, the end wall effect was thought to be minimal, in part because of the drawdown effect of the sludge into the effluent end hopper.

In a related study, the results of the SJCWRP clarifiers were compared to the Metropolitan Water Reclamation District of Greater Chicago's (MWRD's ) Central Treatment Plant (CTP) and the New York City Department of Environmental Protection's (NYCDEP's) Rockaway Water Pollution Control Plant (RWPCP), each of which had been evaluated using the CRTC protocol (Wahlberg et al., 1994). The CTP had two test clarifiers, each with a circular design and 43 m (140 ft) in diameter with a sidewall depth of 4.3 m (14 ft) and a floor slope of 2%. They were both modified to be center feed with peripheral effluent removal. The CTP-7 unit was equipped with a

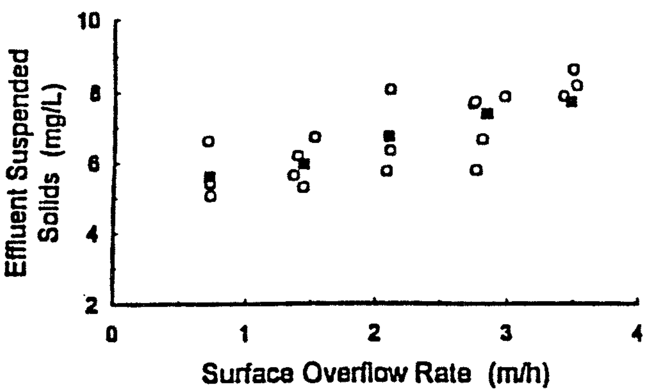


**FIGURE 9.34** Effluent suspended solids versus SOR at San Jose Creek WRP clarifier during stress tests. Raw data indicated by circles, average data indicated by squares (Wahlberg et al., 1993).

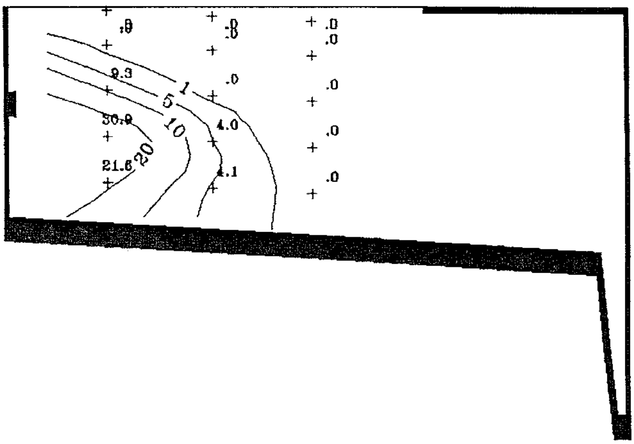

FIGURE 9.35 Dye flow pattern measured in ppb at 2 m/h (1200 gpd/sq ft) for San Jose Creek WRP shallow clarifier (Wahlberg et al., 1993).

scraper plow that moved sludge to the center for removal, and the CTP-9 unit was equipped with a draft tube system that removed sludge at 10 locations across the diameter of the tank with rotating rake arms. The RWPCP is a rectangular unit that is 69 m (225 ft) long by 17 m (57 ft) wide by 3.7 m (12 ft) deep. This clarifier is sloped at 1% and equipped with three chain-and-flight collectors performing cocurrent sludge

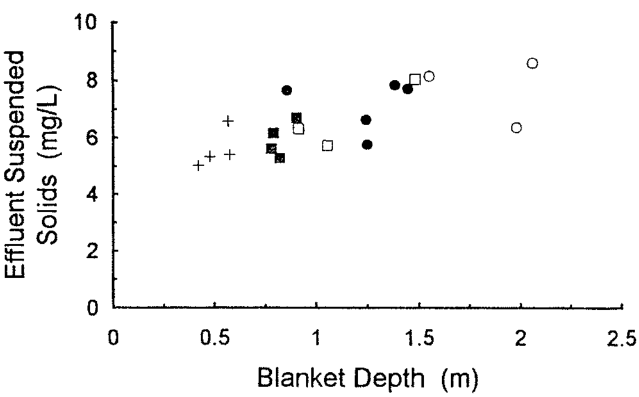


FIGURE 9.36 Effluent suspended solids versus blanket depth of San Jose Creek WRP 3-m deep clarifier during stress testing. Legend: solid squares = 1.36 m/h; open squares = 2.04 m/h; solid circles = 2.72 m/h; and open circles = 3.40 m/h (Wahlberg et al., 1993).

removal to the midtank hopper at the front of the tank and three chain-and-flight collectors performing countercurrent sludge removal to the midtank hopper at the back of the tank. A seventh chain-and-flight collector transports sludge to the hopper, and its travel is perpendicular to the flow. No relationship was found with SORs up to 2.5 m/h (1500 gpd/sq ft) using the combined data (although SJCWRP did so individually). The ESS also increased with increasing MLSS and SLR. It was also observed that clarifier performance deteriorates when non-ideal flow patterns develop.

Samstag (1988) compared large circular clarifiers at Renton, Washington, with a depth of 5.5 m (18 ft), to the SJCWRP rectangular clarifiers, with a depth of 3.0 m (10 ft), and indicated little difference in the performance (see Figure 9.37). Stahl and Chen

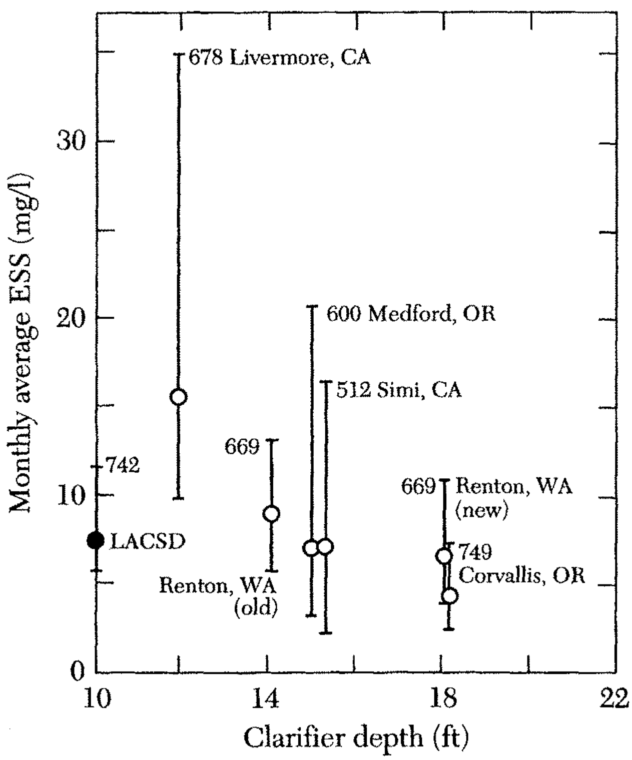


**FIGURE 9.37** The effect of clarifier depth on monthly ESS showing 10, 50, and 90 percentile values. Shallow rectangular clarifiers of LACSD (solid circle) shown with other circular clarifiers (open circles) having center flocculator wells (except Renton). Average SOR given numerically listed at top of percentile values (adapted by Stahl and Chen [1996] from Parker [1983]) (ft × 0.304 8 = m).

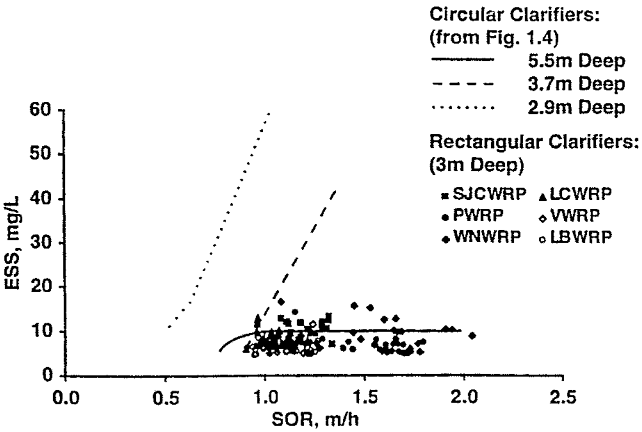


**FIGURE 9.38** Comparison of SOR versus ESS for circular clarifiers (from Parker and Stenquist, 1986) with LACSD shallow rectangular clarifier data superimposed (Stahl and Chen, 1996).

(1996) superimposed a plot of monthly average ESS concentrations from six of LACSDs water reclamation plants, which are 3 m (10 ft) deep and similar to SJCWRP, and compared them to other clarifiers. The superimposed data points showed the shallow rectangular clarifiers to be equivalent to a 5.5 m (18 ft) deep circular clarifier with a flocculator design (see Figure 9.38). The critical design features were pointed out to be the inlet diffusers that distributed and flocculated the influent flow and the cocurrent sludge removal. This allowed these shallow rectangular clarifiers to produce low ESS concentrations with little variability. This further lends credence to a compilation of data by Dittmar (1987) that led to the observation that there is no observable difference in clarification performance at average or peak hydraulic loading rates attributed to shape of a clarifier (rectangular or circular) alone.

Although the trend in recent years has been towards increasing depths of clarifiers for improved performance, the data presented from this testing show that shallow rectangular clarifiers with a well-flocculated mixed liquor can perform extremely well at elevated SORs. However, it still remains to be demonstrated how shallow rectangular circulars would compare to deep circular clarifiers at even higher SORs above 3.4 m/h (2000 gpd/sq ft). Some researchers have speculated that the success of the shallow rectangular clarifiers cited above could be because these systems were run as step feed plants with a front-end reaeration zone, mainly for carbonaceous removal only (Metcalf and Eddy, 1991). It is well-known that this mode of operation reduces the solids loading seen by the clarifiers. Also, in this mode, there are little, if any, sludge blankets in the clarifiers. However, it should be noted that, since the initial study of the SJCWRP clarifiers, the plant has converted to a step feed nitrification/denitrification mode of operation. It now operates with higher MLSS concentrations (up to 4000 mg/L), higher SLRs (up to 40 lb/d/sq ft), higher recycle rate (up to 80%), and with occasional 0.9-m (3-ft) blanket levels. The net result has been a more consistent settling sludge, and an average ESS of 6 mg/L with less variability, than with the previous fully aerated step feed operation.

**DEPTH REQUIREMENT STUDY FOR HIGH-PURITY-OXYGEN ACTIVATED SLUDGE CLARIFIERS.** In anticipation of upgrading treatment from partial to full secondary treatment at the JWPCP in Carson, California, the LACSD performed an assessment of their rectangular clarifiers (Hashman et al., 1997). At the time of the study, 8.8 $m^3/s$ (200 mgd) was treated with a HPOAS system using four reactor trains of 2.2 $m^3/s$ (50 mgd). Each train had 26 rectangular clarifiers, for a total of 104 clarifers, with cocurrent sludge removal at the effluent end using chain-and-flight collectors. With full secondary treatment, the plant would no longer be able to have a trimmed peak flow, but would have to treat the full diurnal flow condition. Previous testing in 1994 (Hashman and Wahlberg, 1995) found higher-than-expected ESS, and the clarifiers appeared as if they could proceed into an overloaded condition with a relatively small increase in flowrate. The objective of the investigation was to determine if deeper tanks were needed and if a different inlet design would be helpful.

The JWPCP clarifiers are very similar in design to the SJCWRP clarifiers, in most respects, except that they are 4.3 m (14 ft) deep to accommodate the generally increased solids loading of a HPOAS system. Additionally, they are 5 m (17 ft) longer, for a total length of 50 m (167 ft). The HPOAS systems generally have higher ESS than conventional plants, in part, because they are high-capacity systems with shorter aeration times and because the mechanical aeration leads to more dispersed suspended solids (DSS). The JWPCP also treats wastewater with more industrial waste, which is routed around the water reclamation plants. Therefore, the JWPCP influent can be expected to be somewhat more difficult to treat than the upstream plants. The testing followed the CRTC protocol established by the Clarifier Research Technical Committee of the American Society of Civil Engineers. This included stress tests, which were run at 100 to 250% of the design SOR (24.4 to 57.04 $m^3/m^2 \cdot d$ [600 to 1400 gpd/sq ft]) over four HRTs.

The second study has similar results as the first study, namely that at low SORs, the ESS rose moderately with increasing flow. However, above 48.89 $m^3/m^2 \cdot d$ (1200 gpd/sq ft), the ESS increased sharply with SOR (see Figure 9.39) (Note that following graphs have eliminated the over-range data from the previous study and the highest ESS data point, for clarity). This latter observation was ascribed to particle resuspension as the result of strong vertical currents at the end wall and launder region when a high sludge blanket was encountered and higher dispersed suspended solids loading (DSSLR) at higher SORs. Figure 9.40 shows stable but scattered ESS data below the 2-m (7-ft) blanket; however, above this, the ESS can

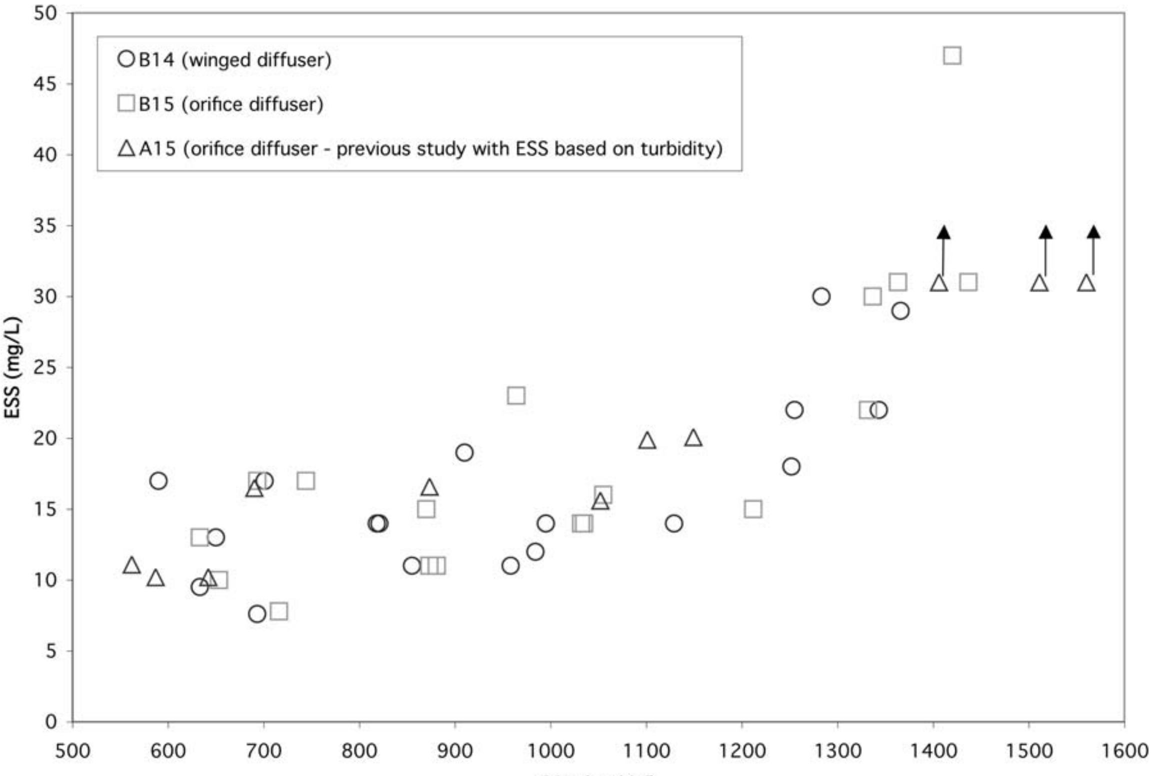


**FIGURE 9.39** Effluent suspended solids versus SOR for LACSD's JWPCP clarifiers (HPOAS system). The legend also applies to Figures 9.39 to 9.45. Arrows indicate turbidity meter is registering at upper end of range (gpd/sq ft × 0.001 7 = m/h; ft × 0.304 8 = m) (Hashman et al., 1997).

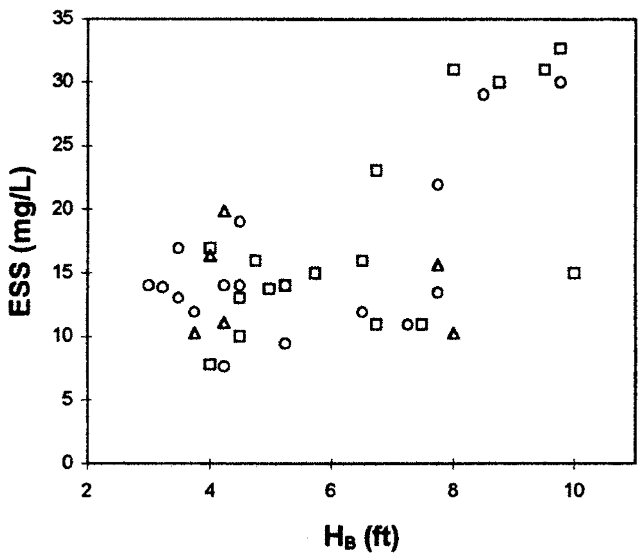


**FIGURE 9.40** Rating curve of sludge blanket height versus ESS for JWPCP 4.3-m-(14-ft-) deep clarifier (ft × 0.304 8 = m) (Hashman et al., 1997).

increase sharply. Predictably, when very high blankets were encountered, this resulted in carryover into the effluent launder. High sludge blankets are increased by high influent flowrates and short detention times. A plot of ESS versus SLR showed an upward trend, with some problems encountered over 127 kg/m$^2$·d (26 lb/d/sq ft) (see Figure 9.41). However, because some of the higher SLRs have lower ESS and because higher SLRs have historically been tolerated at the plant, this leads to the observation that something else was happening during the stress tests. It appears that the mixed liquor concentration was also decreased during the higher SOR tests. The thinning of the mixed liquor resulted in higher zone-settling velocities (ZSVs), which contributed to higher DSS. Because faster settling occurred with the thinner mixed liquor, there would be less of a sweep effect to catch the dispersed particles. The higher DSS combined with the higher SORs to eventually contribute to higher DSSLRs and ESS. These relationships are shown in Figures 9.42, 9.43, 9.44, and 9.45. Finally, it is interesting to note the disparity between the ranges of the DSS in the JWPCP's HPOAS system with mechanical mixers (10 to 16 mg/L) versus that of SJCWRP's fine-bubble diffuser system (2 to 10 mg/L) cited in the previous case history.

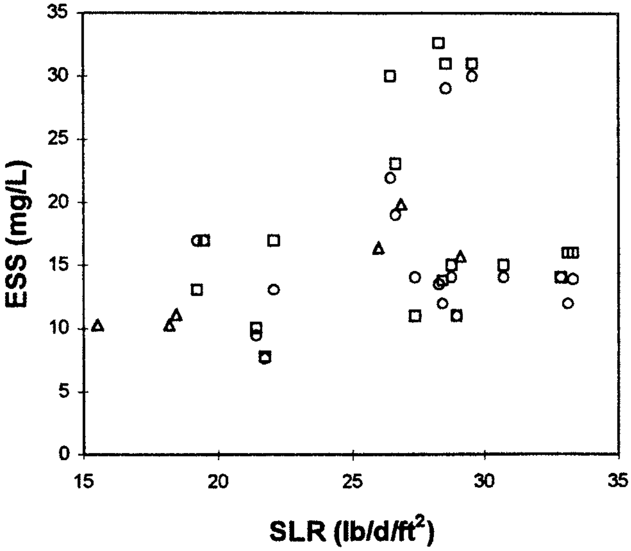


FIGURE 9.41 Joint Water Pollution Control Plant clarifier rating curve of SLR versus ESS. In conjunction with Figure 9.39, it shows how the MLSS had decreased during stress testing at higher SORs and affected the F/M ratio (lb/d/sq ft × 0.203 4 = kg/h/sq m) (Hashman et al., 1997).

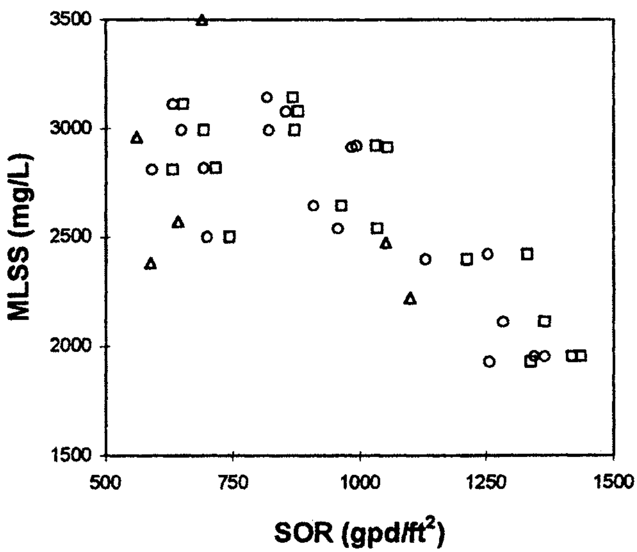


FIGURE 9.42 Mixed liquor suspended solids versus SOR for JWPCP clarifiers showing dilution of mixed liquor at high flowrates (gpd/sq ft x 0.001 7 = m/h) (Hashman et al., 1997).

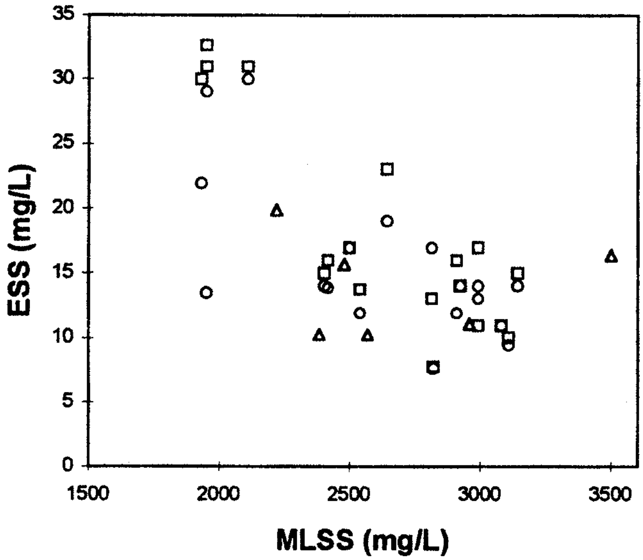


**FIGURE 9.43** Mixed liquor suspended solids versus ESS for JWPCP clarifiers showing trend of elevated ESS at lower MLSS concentrations (Hashman et al., 1997).

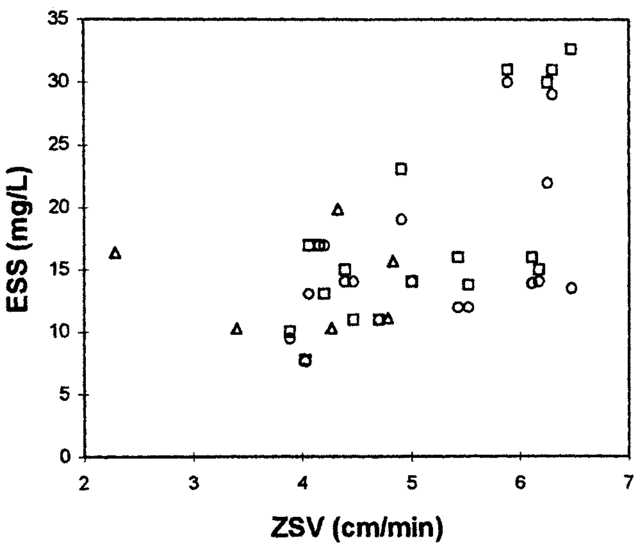


**FIGURE 9.44** Effluent suspended solids versus ZSV for JWPCP clarifiers showing a tendency of higher ESS values at high settling velocities (cm × 0.394 = in.) (Hashman et al., 1997).

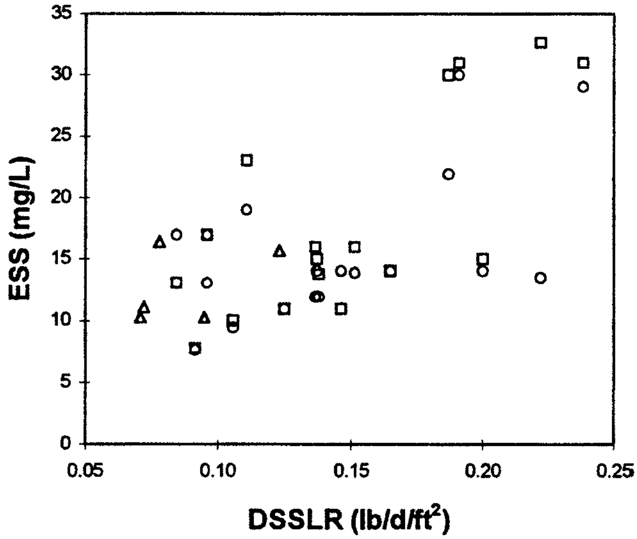


**FIGURE 9.45** Joint Water Pollution Control Plant clarifiers showing a trend to an increase in ESS with dispersed suspended solids loading (lb/d/sq ft × 0.203 4 = kg/m$^2$·h) (Hashman et al., 1997).

Modeling was performed to find the critical point in the SOR rating curve. The model showed that the current 4.3 m (14 ft) depth was adequate to accommodate the expected diurnal flow pattern, provided that the sludge blankets are not allowed to exceed a certain point. The critical blanket height was determined to be approximately 2.0 m (6.7 ft). If the depth was increased to 4.9 m (16 ft), modeling predicted the critical blanket height would be 2.4 m (7.8 ft). The diurnal flow test showed that the sludge blankets at peak flows were approximately 0.9 m (3 ft) higher than the blankets at low flow. Therefore, if the blankets at low flow were kept below 1.1 m (3.5 ft) at low flow, suspended solids should not be lost in the effluent at peak flow.

Two inlet designs were then compared by studying the ESS/SOR relationship in separate tanks. One tank had the existing tubular orifice design (see Figure 9.46), while the other had the LACSD's conventional spider or winged design that is installed on its upstream water reclamation plants (see Figure 9.7). The latter inlet design has been reported to give good results and promote flocculation. The existing inlet diffuser was shown to achieve lower ESS at low SORs, and the spider design was shown to achieve lower ESS at higher SORs. At typical SORs in the range 0.85 to 1.2 m/h (500 to 700 gpd/sq ft), no significant difference in ESS was found. However,

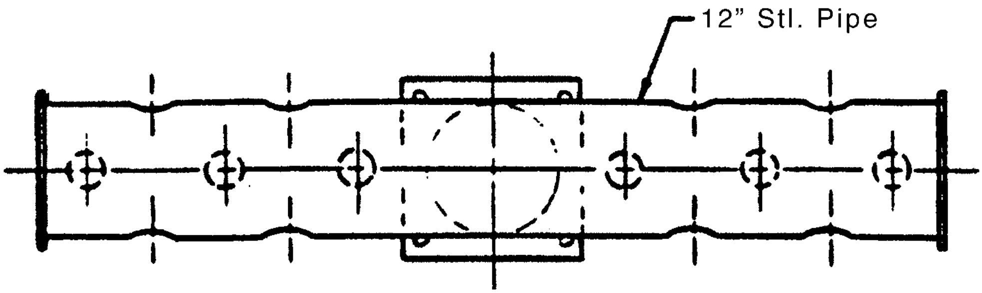


**FIGURE 9.46** Section view of existing inlet diffuser at JWPCP clarifier with 76-mm (3-in.) orifices jetting horizontally against wall, upward to surface and downward to fillet (in. × 2.54 = cm) (adapted from Hashman et al., 1997).

when ESS based on turbidity correlations was plotted versus SOR, it showed that ESS was significantly lower for the winged diffuser. This apparently different picture could be explained by the winged diffuser promoting flocculation and reducing turbidity, but having less effect of really removing suspended solids. This calls into question the practice of using turbidity measurements as an absolute measure of ESS instead of performing ESS composite sampling during stress tests.

Eventually, other investigations showed that, at peak flows, the existing inlet diffuser produced a boil at the inlet of the tank, such that mixed liquor continuously short-circuited to the skimming pan. Because the existing diffusers were corroded and needed replacement, a decision was made to retrofit the existing clarifiers with that design and the proposed clarifiers as well. This study also also pointed out that the flow distribution to the clarifier should be investigated (Hashman et al., 1997). With so many clarifiers (26) servicing one reactor, it had long been documented that individual clarifiers exhibited disparate blanket levels (high or low) during overall high blanket conditions. Therefore, a method was developed to fine tune the percentage of recycle from the problem tanks to keep their blankets (and hopefully thickening and clarification functions) in line with the other clarifiers, thereby helping to improve the overall system efficiency.

**CONVERSION OF LONGITUDINAL FLOW RECTANGULAR CLARIFIERS TO TRANSVERSE FLOW.** One-half of the Gwinnett County Yellow River/Sweetwater Creek Wastewater Treatment Plant in Gwinnett County, Georgia, was plagued with high maintenance demands and frequent ESS losses. This half of the plant had four rectangular clarifiers equipped with floating siphon removal mechanisms that frequently lost suction because of ragging of the collector pipes and had other mechanical breakdowns. In comparison, the other half of the plant had two peripheral feed, peripheral overflow (looped flow) circular clarifiers with rotating hydraulic removal headers. Even though the two halves of the plant had identically operated biological systems, the circular clarifiers produced a more consistent effluent. This is evidenced by the fact that the rectangular clarifiers had logged 26 ESS events over 25 mg/L, while the circular clarifiers only had three such events in the first half of 1997. The four rectangular tanks are 36.6 m long × 9.2 m wide × 3.9 m side water depth (SWD) (120 ft long × 30 ft wide × 12.75 ft SWD). The two circular clarifiers are 33.6 m (110 ft) diameter × 4.3 m (14 ft) SWD. From these measurements, it can be calculated that the circular tanks had 32% more surface area and were 10% deeper.

Gwinnett County decided to convert the rectangular tanks to transverse clarifiers with a looped flow pattern with fixed hydraulic removal headers that were tapered and fitted with inlet control orifices (Gross et al., 2003). These were embedded in the clarifier floors. Four shaft chain-and-flight collectors were installed to scrape sludge to the collector headers and skim the surface. Tapered steel influent/effluent channels were mounted on the same clarifier sidewall and constructed with a common divider wall. The influent flow was controlled by relatively small and closely spaced orifices, which were drilled through the bottom of the channel. Target baffles were positioned below each orifice to dissipate the vertical flow velocity. Additionally, an inlet skirt was provided beneath the channels to prevent turbulence and assure a low velocity in the clarification zone. Essentially, the rectangular clarifier conversion was designed to duplicate the pattern that Crosby (1984b) found in peripheral feed, peripheral overflow circular clarifiers, which had provided satisfactory results on the other half of the plant.

A limited stress test was done on a converted test clarifier because the exiting return sludge channels limited the overflow to 1.9 m/h (1100 gpd/sq ft). Although the sludge blanket depth stabilized at 2.1 to 2.4 m (7 to 8 ft), or approximately 25 cm (10 in.) below the inlet skirt, it was felt that the relatively high mixed liquor concentration (3100 mg/L) during the four-hour test may have forced the inlet flow into the blanket layer instead of short-circuiting on the top to the effluent channel. A uniform sludge blanket was found across the width of the clarifier. Before the conversion, the return sludge solids concentrations for the rectangular clarifiers averaged 3% higher than the circulars. After the conversion, this increased to 9% higher than the circulars.

An increase in the accumulation of rags at the end wall was compensated for, by withdrawing a greater volume of sludge at that location. The rag problem is still a nuisance but now the rags pass through the clarifier and collect on the telescoping sludge valve bails as they overflow the slip tube, which are accessible for cleaning. This prevented the necessity of monthly dewatering of a rectangular clarifier, which was the frequency before the conversion. More importantly, the ESS of the two sides of the plant became more balanced. In summary, the retrofit of longitudinal rectangular clarifiers demonstrated that transverse clarifiers were a viable option for design and showed satisfactory performance at high surface overflow and SLRs.

## SUMMARY, CONCLUSIONS, AND RECOMMENDED RESEARCH

Following from the text of the chapter and the case histories that have been presented, a number of observations are in order. There is still great debate about many design features of rectangular clarifiers. The design engineer can still expect to struggle with decisions about the proper width, depth, and length of rectangular clarifiers and hopper location. Many times, the design decisions come down to the traditional perspective of the consultant or the owner/agency. It is obvious that many designs can be fine-tuned. At times, retrofitting a rectangular clarifier with baffles, altering sludge collection, or changing the length of the clarifiers can have positive effects on removal efficiency. Other times, the results can be insignificant or not cost-effective, depending on discharge requirements and flexibility with upstream or downstream processes.

Some of the best performing rectangular clarifiers are those that are relatively shallow and the simplest to design, build, and operate. Following a simple design scenario, the engineer would probably opt for an inlet channel that had the ability to distribute the flows to multiple clarifiers equally and was able to flocculate the mixed liquor to some degree. There would be provisions for foam control and the capability of adding chemicals for flocculation and settling during upset conditions. Multiple longitudinal tanks could be provided for best hydraulic characteristics and to prevent downtime of large clarifier volumes. Each tank would be provided with a plastic chain-and-flight system to condition the settled sludge and reduce RAS pumping cost. The inlet of the rectangular clarifier would be positioned low in the tank, just above the blanket, and provide good distribution of the flow across the cross-section of the tank. Perhaps the flexibility of a future inlet flocculation zone or baffle should be built into the design, even though these feature may not be needed in the near future. Barring the rapid takeover of a new technology over conventional systems (perhaps possible with membrane technology), and because the design life of clarifiers could easily be in the order of 50 to 100 years, it seems reasonable to assume that loading conditions, reactor operations, and regulatory goals will change during that lifetime. Two hoppers for a nominal 6-m- (20-ft-) wide tank would be placed on the effluent end of the tank to take advantage of the density current and prevent short-circuiting to the hopper. The effluent end hopper would also take advantage of the gallery running transversely across the end of the tanks and result in short runs of return sludge collection piping. That piping would be slightly oversized to accommodate stress testing and higher return rates. Weirs would be double-sided longitudinal weirs spread over 30 to 50% of the tank and be covered to prevent algae problems. Some sort of baffling would be provided at the end wall or provisions of blocking the last section of weir by the end wall. Foam control would be accomplished by providing a slotted pipe or small pan skimmer with sprays. Although midtank hopper designs give excellent results, it also increases the complexity of the sludge removal systems by doubling the chain-and-flight mechanisms and lengthening the sludge piping. Other features, such as inlet hoppers, may be avoided because of potential short-circuiting. Cross flights would be avoided because of hydraulic instability. Internal baffles would probably be avoided because of their operational complexity and cost.

The above scenario does not seek to discredit other designs or the work of other investigators. Its main purpose is to guide the design engineer to a simple rectangular clarifier design solution first. After that, the increasing complexity and cost of various other design options need to be correctly linked to site constraints and regulatory requirements. For instance, would it make sense to have a complex secondary clarifier design, for a marginal increase in suspended solids removal of a few milligrams per liter, if tertiary treatment or gravity filtration is already required by effluent regulations? At some point, it may be more cost-effective to voluntarily put in a polishing step for tertiary treated effluent than to chase diminishing returns and capture the last of the elusive particles escaping the clarifiers.

It is important to note that rating curves that determine ESS versus SOR or any other parameter should be taken cautiously, as they are often site-specific. Site specificity is largely determined by influent wastewater characteristics, effluent regulations, and by the size, configuration, and operation of the aeration tanks (or biological reactors) and the changing settleability of the sludge. In part, when the engineer or operator is performing clarifier stress testing, they are really revisiting the basic design issue of determining the relative size of reactors versus clarifiers; larger reactor volumes can be coupled with smaller clarifier volumes or vice versa. A larger reactor may tend to favor lower ESS because the loading (food-to-microorganism ratio [F/M]) will be lower, and a higher MCRT can be tolerated. If a higher MLSS is used, it could have a tendency to slow down the zone-settling velocity, which would help to capture dispersed suspended solids. It would also lend itself to less vigorous and longer aeration times, which may promote more flocculation and keep the DSSLR lower, especially in the case of mechanical mixing. Alternatively, an increased reactor volume can be used to run with lower MLSS, which would decrease the SLR. Regardless of the way a larger reactor volume is run, the net effect would be a net decrease of the operating pressure or stress on the clarifiers.

One should also keep in mind that the shape of an ESS versus SOR rating plot is a horizontal or a nearly horizontal line, where the clarifiers are not particularly stressed and ESS is relatively steady or slowly increasing. As liquid flow and solids loading increases, sludge blankets increase with the velocity of the density current, and the end wall effect is exacerbated. As this and other hydraulic instabilities occur, floc resuspension or short-circuiting will increase the ESS, and a transitional area is reached that may produce more scattered data, but nevertheless can be interpreted as an upward trend. This phase of instability and diminished clarification, however, is transitional. With higher flowrates and solids loading, the clarifier ultimately fails as rising blankets are swept over in the weir zone. The ESS at this point essentially goes straight up and off the chart. When this happens during a stress test, these events are documented in the text, but the data are obviously never plotted. One could argue, therefore, that the most important thing to know about a clarifier is at which loading rate the clarifier (or clarifier system) becomes unstable for a given reactor operation. This not only helps to define the limits of the clarifiers, but also defines the operational limits of the reactor operation. From these limits, the most efficient operational point for an activated sludge system can be deduced, which is generally tied to the power cost for aeration requirements.

It is interesting to more closely examine how a reactor operation can be affected during a clarifier stress test. For a reactor that is operating close to its own critical loading point, by decreasing the number of clarifiers in service, less thickening is achieved. The MLSS of the reactor may decrease over next the 3 to 4 HRTs, even if the return rate is increased. Because treatment can be shown to be directly related to MCRT and reactor HRT and inversely related to loading, any increase in suspended solids may be more related to reactor kinetics or some variability in the chemical oxygen demand (COD) loading. Therefore, rating curves should be viewed as a snapshot of a particular time and operation. For a given system, there would conceivably be a family of curves, which shift depending on a number of variables, such as influent flow, COD loading, and MLSS. A hypothetical family of rating curves for a given system is shown in Figure 9.47. Different activated sludge processes and different types of aeration would generally fall into certain ranges in the family of curves. The HPOAS systems that are mechanically aerated can generally be expected to have higher ESS, whereas fine bubble and coarse bubble can be expected have

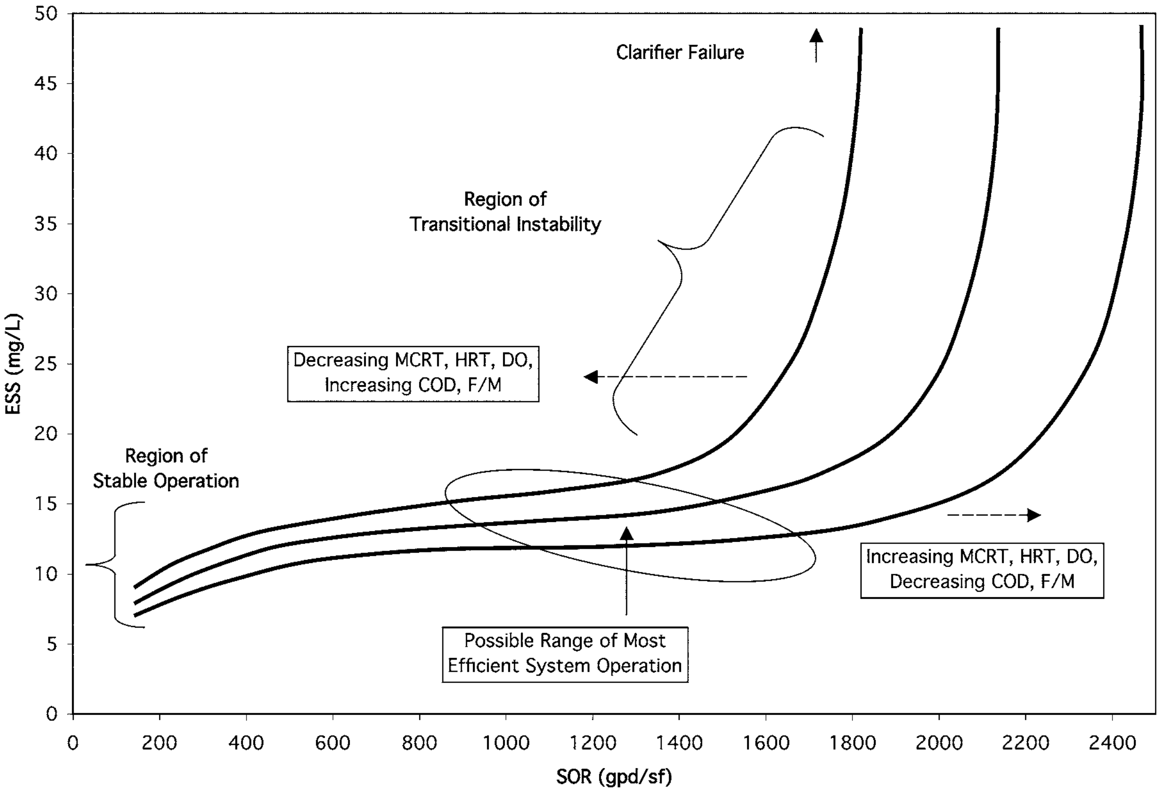


**FIGURE 9.47** Hypothetical series of SOR rating curves for the same conventional activated sludge system at peak flowrate with different operational conditions (gpd/sq ft × 0.001 7 = m/h; DO = dissolved oxygen).

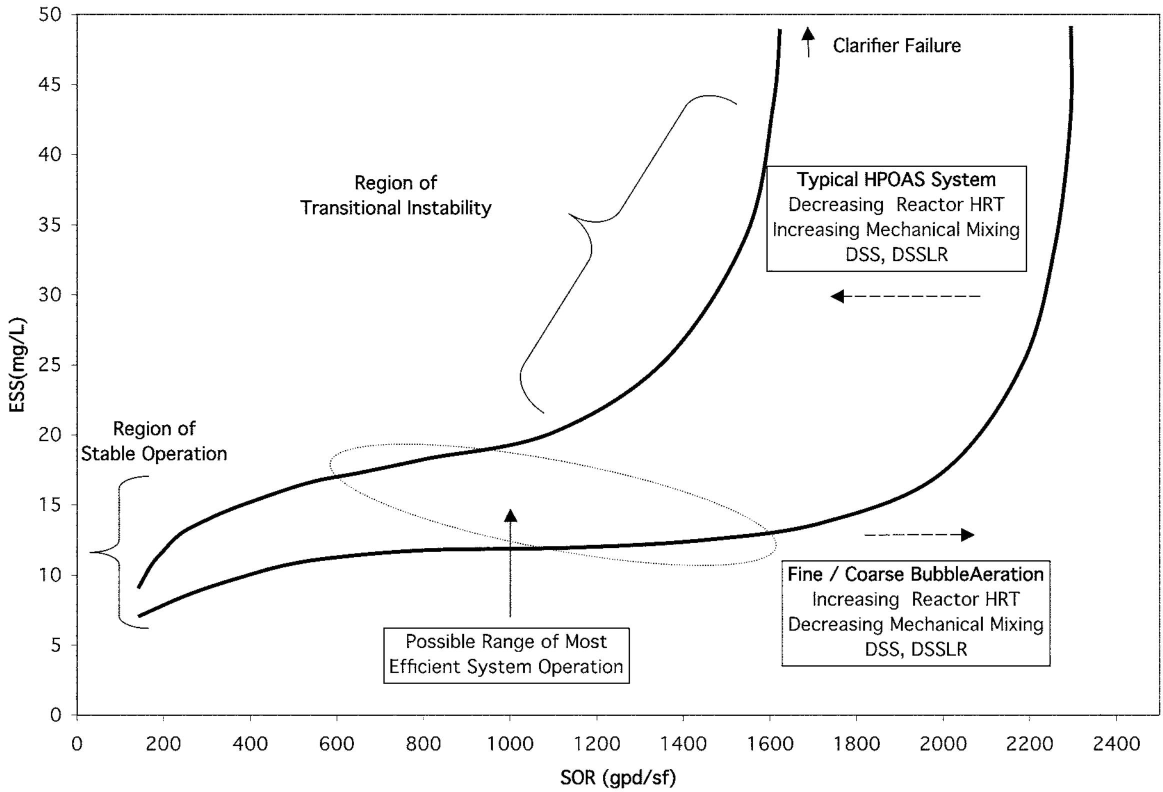


**FIGURE 9.48** Hypothetical ESS versus SOR rating curves for a conventional and HPOAS system (gpd/sq ft × 0.001 7 = m/h).

lower ESS. Because the HPOAS systems have shorter aeration times, they can also be expected to enter the transition phase of diminished clarification earlier. Increased clarifier depth would have the effect of delaying failure, but not necessarily decreasing the onset of transitional instability. A hypothetical set of rating curves for activated sludge systems, using similar clarifiers, is shown in Figure 9.48.

Perhaps the best way to verify the reactor/clarifier operation for any plant is by taking advantage of the occasions when a reactor and clarifier unit is taken out of service. By taking one or more of the clarifiers out of service every few days, in essence, a stress test is performed. By careful variance of the flows and loading to the reactor during different shutdowns, the family of clarifier rating curves can be established to definitively establish operating parameters for any given operating condition. It would be helpful if a data collection protocol could be established (similar to the CRTC protocol) for sharing this information.

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# Chapter 10

# Clarifier Performance Monitoring and Control


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## INTRODUCTION

Clarifier process performance monitoring and control takes many forms and can range from complete manual operation to complete automation of clarifier sludge collection and withdrawal and control of the clarifier's sludge blanket level. Properly designed, installed, and maintained instrumentation for clarifier monitoring allows generating a reliable online database for efficient process control and decisionmaking.

Removal of sludge in a manner that allows maximizing solids concentration, while minimizing the effect of sludge removal on effluent water quality, is a principal task and goal in clarifier design. Clarifier sludge concentration has a significant effect on the capacity and performance of the downstream solids handling facilities. Maximizing clarifier sludge concentration typically improves solids handling facility operations and increases their process capacity.

Clarifier underflow sludge concentration pumped out of the bottom of the tank can be maximized by appropriate design of the sludge collection and removal systems and adequate frequency of sludge withdrawal combined with provisions for close monitoring of sludge blanket depth. In optimal design and operation, sludge has to be removed from the clarifiers at the rate of its generation after providing ample time for sludge thickening to optimal concentration. If, during facility operation, the sludge concentration is reduced significantly, the clarifier sludge collection and removal rates could be reduced and, if needed, collection and removal could be discontinued until sludge concentration returns to an optimal range. On the other hand, an increase in sludge concentration above an optimal range and a steady increase in sludge blanket depth is an indication of the need to increase sludge collection and withdrawal rates to avoid the negative effect of excessively high sludge blanket zone on the clarification zone and ultimately on the clarifier effluent water quality.

Because the sludge generation and withdrawal rates can vary over time as a result of fluctuations of influent water quality and quantity, frequent or continuous monitoring of the concentration of sludge removed from the clarifiers and clarifier sludge blanket provide an indication of the overall clarifier performance.

# KEY PARAMETERS

**PRIMARY CLARIFIERS.** Primary clarifier performance is typically measured by their total suspended solids (TSS), biochemical oxygen demand (BOD), and phosphorus removal efficiency and by the condition of the primary sludge (sludge septicity, concentration, and volume). Primary clarifier performance in terms of effluent TSS, BOD, and phosphorus removal efficiency is discussed in detail in Chapter 2.

Proper primary clarifier sludge collection, removal, and withdrawal are of key importance for maintaining consistently high primary effluent quality and efficient and cost-effective solids handling. If primary clarifier sludge is retained for an excessively long time in the tanks, the sludge could turn septic because of its high organic content and anaerobic conditions in the sludge blanket. Sludge septicity is accompanied by the release of malodorous gases that may disturb the normal sedimentation process as they travel from the tank sludge blanket to the surface. Septic sludge is also more corrosive and more difficult to pump and dewater. Besides creating conditions for sludge septicity, maintaining a relatively deep sludge blanket in the primary clarifiers may also make sludge collection and withdrawal more difficult because of excessive compaction and, in extreme conditions, may cause damage of the sludge collection and withdrawal equipment (broken sludge collectors, plugged solids lines, and damaged pumps).

A widely accepted practice to prevent primary sludge septicity and its negative effect on clarifier performance is not to carry a sludge blanket by continuously or very frequently removing sludge from the clarifier. When not controlled appropriately, continuous sludge withdrawal or overpumping would result in conveyance of a very diluted sludge to the downstream solids handling facilities, which would have a negative effect on their performance.

In shallow primary clarifiers (i.e., clarifiers with a sidewater depth of less than 3.66 m [12 ft]) or clarifiers settling highly septic plant influent (i.e., influent which contains a hydrogen sulfide concentration higher than 50 mg/L), continuous sludge withdrawal, although not optimal from the downstream solids handling facilities, may be necessary to prevent the negative effect of the clarifier blanket on the primary effluent TSS concentration. Under this sludge removal approach, the sludge blanket is relatively thin, and the primary sludge concentration is typically between 0.5 and 1% solids.

Full-scale studies of shallow primary circular clarifiers at two large wastewater treatment plants (Albertson and Waltz, 1997) reveal a close adverse correlation between clarifier sludge blanket retention time and TSS removal. The sludge blanket retention time is estimated by dividing the solids mass in the clarifier by the mass rate of the solids removed in the underflow. For simplicity, the solids mass in the clarifiers in this study was determined assuming that the average sludge blanket concentration in the clarifiers is equal to 50% of the clarifier underflow concentration. The sidewater depth of the studied primary clarifiers at the two plants was 2.74 and 3.14 m (9 and 10.3 ft), respectively. The optimum primary clarifier sludge blanket retention time at the studied wastewater treatment plants was found to be 6 to 12 hours, and the optimum, maximum sludge blanket depths were 0.125 and 0.3 m (5 and 12 in.), respectively.

To avoid overpumping of diluted sludge and prevent the negative effects of an excessively deep sludge blanket and associated septicity, the primary clarifier sludge blanket and concentration must be maintained at optimum levels. The optimum primary sludge concentration is typically in a range of 3 to 6%, and the most viable sludge blanket depth is typically between 0.3 and 1.0 m (1 and 3 ft). The optimum sludge blanket depth would vary seasonally and would change during dry weather and wet weather conditions. This optimum sludge blanket depth would also depend on the clarifier overall sidewater depth and solids inventory and the influent wastewater septicity and strength, in terms of BOD and TSS.

Currently, at most plants, primary clarifier sludge concentration and blanket depth are monitored based on manual sample collection and analysis. However, reliable sludge concentration analyzers and blanket level detectors that can withstand the negative environment found in primary clarifiers are commercially available.

**SECONDARY CLARIFIERS.** Secondary clarifier performance has a significant effect on plant effluent water quality, aeration basin mixed liquor suspended solids (MLSS) concentration and performance, and efficiency of plant solids handling facilities. Key secondary clarifier effluent quality monitoring and control parameters such as TSS, BOD, ammonia, total nitrogen, and total phosphorus concentration and removal efficiency are discussed in detail in Chapter 4.

Besides effluent water quality, there are four key process variables that need to be monitored for efficient and cost-effective clarifier solids control: (1) the amount of solids retained in the secondary clarifiers, which is determined based on the concentration of the return activated sludge (RAS) and the waste activated sludge (WAS) removed from the clarifiers, and on the clarifier sludge blanket depth/volume; (2) the amount of solids in the aeration basins, which is determined by measuring the MLSS concentration and the plant influent flowrate; (3) the activated sludge settleability; and (4) the plant influent flow and waste load, significant fluctuations of which may result in the transfer of large amounts of solids from the clarifiers to the aeration basins and ultimately cause solids carryover with the secondary clarifier effluent.

***Monitoring of Activated Sludge Solids Inventory.*** The amount of solids retained in the clarifiers can be monitored by frequent manual or automated measurements of the sludge blanket depth and WAS/RAS solids concentration. Monitoring the concentration and volume of the sludge removed from the secondary clarifiers (WAS) and retained in the aeration basins (MLSS) is a critical part of any activated sludge system control strategy. The MLSS and WAS concentration and RAS recycle rate measurements are routinely used to adjust the WAS withdrawal rate and to maintain consistent steady-state activated sludge system performance.

Currently, manual sludge sample collection followed by gravimetric TSS analysis in the plant laboratory is the most frequently practiced method for monitoring activated sludge MLSS, RAS, and WAS concentration fluctuations. Typically, plant staff collects one to three sludge samples throughout the day and analyzes these samples for total suspended solids applying standard laboratory methods and procedures or using laboratory high-speed centrifuges. Standard laboratory TSS analysis is relatively time-consuming (typically 2 to 3 hours needed to perform), and, therefore, because of practical time constraints, it is generally completed only a few times per day for larger facilities and less frequently for smaller plants. Solids determination by centrifugation of activated sludge samples takes only 15 to 20 minutes and is widely practiced in many plants. This test is based on developing a graph correlating the volume of centrifuged solids (generally expressed as percent of the total sample volume) and the sludge suspended solids concentration based on parallel measurements of these parameters on the same set of samples. The graph is then used for direct reading of the solids concentration based on the volume of the separated solids in the centrifuged sample. Although less accurate, the centrifuge test is relatively quick to complete, inexpensive, and requires less skill to run than the standard gravimetric TSS analysis.

Currently, a number of sludge concentration analyzers are commercially available for online measurement and monitoring of MLSS and secondary clarifier WAS concentrations. Continuous solids concentration measurements allow for tracking solids inventory fluctuations in real time and getting more accurate representation of the activated sludge system performance. In addition, automated solids inventory monitoring avoids the effect of human errors and reduces the time required for sampling and sample processing. On the other hand, installation and operation of online instrumentation for solids inventory monitoring requires additional expense, more specialized operator skills for calibration and servicing of sensors, and frequent equipment field testing to avoid potential errors caused by inaccurate readings or instrument drift. Therefore, the use of such equipment is recommended to be established based on site-specific, life-cycle, cost-benefit analysis.

Successful use of automated activated sludge solids inventory control systems has been reported at a number of medium- and large-size wastewater treatment plants (Ekster, 1998 and 2000; Hinton-Lever, 2000; Samuels, 2000; Wheeler et al., 2001). Activated sludge system performance optimization study completed at the 93 000 $m^3/d$ (25 mgd) Burlington Skyway wastewater treatment plant in Halton, Ontario, Canada (Wheeler et al., 2001), has proven that automation of secondary clarifier sludge blanket level monitoring combined with close monitoring and control of activated sludge solids inventory can yield significant improvement of the plant effluent quality at minimal additional expense.

***Monitoring of Sludge Settleability.*** While monitoring sludge blanket, activated sludge solids inventory, and plant influent flow variations allows gaining a general understanding of the clarifier performance, it is also very advantageous to track changes in activated sludge settleability. As mentioned previously, primary and secondary sludge blanket depth and settleability do not fluctuate significantly under steady-state influent flow and load conditions. A sudden increase in sludge blanket depth in the secondary clarifiers at typical influent flows and loads and well-operating sludge withdrawal pumps generally indicates deterioration of sludge settleability. Currently, there are a number of widely used parameters and procedures for determining activated sludge settleability. These parameters and their measurement, advantages, and disadvantages are discussed in detail in Chapter 4.

***Plant Influent Flow and Load Monitoring.*** Accurate plant influent flow measurement and monitoring are essential for efficient control of the clarification process. In many plants, influent flowrate is used as a main activated sludge system control parameter, adjusting the transfer (RAS) rate of solids from the secondary clarifiers to the aeration basins proportionally to the influent flow changes. As a minimum, online flow measurement devices are recommended to be installed for continuous monitoring of plant influent flow and RAS and WAS flowrates.

# MONITORING AND CONTROL EQUIPMENT AND INSTRUMENTATION

**INTRODUCTION.** To date, automated clarifier monitoring and control has found application mostly in medium and large wastewater treatment plants. A recent survey of more than 110 wastewater plants at 45 utilities in the United States (Hill et al., 2001) indicates that only 10% of the surveyed plants use primary or secondary clarifier sludge blanket level monitoring instrumentation, and approximately 5 to 10% of the plants apply suspended solids concentration analyzers. This survey also indicates that, typically, primary sludge withdrawal control is timer-based, and timer frequency is generally adjusted based on sludge blanket depth readings. Facilities included in the survey reported using constant flowrate secondary sludge wasting more than any other WAS control strategy. The second most popular WAS control strategy is based on maintaining constant sludge age/solids retention time (SRT) in the activated sludge system.

Installation of automated sludge concentration measurement and blanket monitoring equipment is warranted for large treatment plants where sludge withdrawal from the clarifiers is generally continuous and where the sludge is treated in anaerobic or aerobic digesters. In small plants, where sludge is generally removed intermittently, or for facilities with minimal influent flow and waste load variations, installation of sludge concentration and blanket depth measurement instrumentation may have limited benefits. The sections below discuss the existing technologies and equipment available for measuring sludge concentration, density, and sludge blanket depth.

**MONITORING OF CLARIFIER DRIVE UNIT OPERATION.** Installing instrumentation for monitoring of clarifier sludge collection mechanism drive unit operation is a good engineering practice. The purpose of this instrumentation is to provide protection of clarifier drive gearbox and sludge collection flights/arms, and ultimately to prevent clarifiers from failure and the need for costly and lengthy repairs. Typically, monitored drive unit parameters are torque, power, and motion detection.

Wastewater plant operators sometimes use torque gauges or motor power monitors for an indirect monitoring of clarifier sludge concentration (Wilkinson, 1997). This approach, however, is relatively simplistic and inaccurate, because torque gauges and power monitors are designed to provide protection of the clarifier driver mechanisms against overloading rather than to indicate solids concentration.

***Clarifier Drive Torque Monitoring.*** Most suppliers of clarifier drives provide drive torque monitoring devices as a part of their drive mechanism package. High torque and high-high torque warning, alarm, and shutoff switches are typically installed at each clarifier drive mechanism. The plant clarifier programmable logic controller (PLC) is set to monitor clarifier drive on/off (motion) status, the clarifier high torque warning and high-high torque warning/shut-off switches, and to generate alarms. Torque indication can typically be read from a scale, which is expressed as a percentage of the maximum torque load.

If the high torque warning switch senses a high-torque condition (typically when the torque load reaches 40 to 50% of the maximum design drive torque), it sends a signal to the PLC, which generates a high-torque alarm. This alarm is generally displayed on the plant operator's main control panel. The PLC resets when the torque condition clears. If the high-high torque switch senses that the torque load reaches 80 to 85% of the maximum design drive torque, this switch sends a signal to the PLC to generate a high-high torque alarm at the plant operator's main control panel. If the plant operations staff does not turn the clarifier drive motor off after the actual torque reaches high-high level, the high-high torque switch will turn the drive motor off to protect it from overload and to protect the clarifier sludge collection mechanism from damage.

Some equipment manufacturers offer a positive torque overload protection of the clarifier drives. The positive torque overload protection means that the clarifier sludge collection mechanism drive is designed to produce a controlled preset maximum torque. The drive will run continuously at this torque, but, when needed, it will safely produce a higher, controlled, short-term running torque to keep the solids in the clarifier moving. When the drive with a positive torque overload protection experiences torque load demand above the high-high (cutoff) level, this drive will simply slip, without overheating or overstressing. This type of drive allows overcoming process upset without the risk of damaging the sludge collection equipment.

***Clarifier Drive Power Monitoring and Sludge Pump Withdrawal Rate Control.*** Similar to torque, the clarifier drive motor power (measured in watts) or current draw/amperage (measured in amperes) could be monitored to provide motor and drive overload protection. Power monitors for alternate current motors are readily available by a number of manufacturers, and they can be connected to the wastewater treatment plant PLC to initiate high or high-high power load alarms and drive shutdown, similar to that activated by the torque switches.

In addition, the power monitors could be used to control the clarifier sludge pump withdrawal rate. This is accomplished by the installation of a power monitor on the clarifier motor and by equipping the sludge withdrawal pump with a two-speed or a variable frequency speed motor. To establish the low and high power levels at which the power monitor triggers a change in the sludge pump withdrawal capacity, actual clarifier motor power readings are taken at target acceptable minimum and maximum clarifier sludge blanket levels. These target sludge blanket levels are established based on full-scale operational experience.

When the power monitor reading reaches the trigger level corresponding to maximum sludge blanket level, the power monitor sends an output signal to the plant clarifier PLC, which, in turn, adjusts the sludge pump motor speed upwards to increase the sludge withdrawal rate and vise versa. The power monitor generally also has low-low and high-high power level settings. At the low-low level setting (which is typically below the low power level corresponding to the minimum acceptable sludge blanket level), the sludge withdrawal pump is automatically shut down, because the withdrawn sludge would be of unacceptably low solids concentration. The high-high level of the power monitor is introduced to prevent the clarifier motor drive from overload if the sludge pump fails and/or the clarifier sludge blanket level exceeds a preset maximum.

***Clarifier Drive Motion Monitoring.*** The clarifier sludge collection mechanism motion is typically halted when the sludge blanket level in the clarifier is too high and excessive load is imposed upon the sludge collection flights/arms. Installation of loss-of-motion detection instrumentation is recommended as a minimum measure for clarifier drive and sludge collection mechanism protection. Typically, loss-of-motion switches are installed on the clarifier drives to detect when they stop moving. These switches typically generate an output signal to the plant clarifier PLC, which, in turn, triggers an alarm and may be programmed to automatically shut the drive motor off.

**SLUDGE CONCENTRATION AND DENSITY MEASUREMENT.** Sludge concentration and density measurements are used in wastewater treatment plants to monitor the solids concentration of various process streams to optimize primary and/or secondary treatment system performance. Generally, sludge removal from primary clarifiers is set on timers. This practice, however, often leads to large variations of sludge concentration resulting from fluctuations of sludge blanket level and settleability over time. More consistent sludge solids content can be achieved by frequently or continuously measuring sludge concentration and adjusting withdrawal rate based on the measured concentration. Continuous readings or signals from the sludge concentration or density analyzers could be set to start/stop or control the speed of the sludge withdrawal pumps to minimize pumping of diluted sludge to downstream solids handling facilities.

In the past, sludge concentration and density measurement instrumentation has found limited application in full-scale plants, mostly because of equipment measurement inconsistency and inaccuracy. Analyzer instrumentation problems were typically caused by the presence of air bubbles, sensor fouling, or a change in water color. The new generation of sludge concentration and density measurement equipment has built-in provisions to mitigate these problems and can provide consistent and accurate readings. Reliable sludge blanket and concentration analyzers are currently commercially available and have a proven track record.

Several different measurement methods or types of equipment are used for sludge concentration and density measurements, including light-emitting (optical), ultrasonic, and nuclear type solids analyzers. Table 10.1 summarizes key areas of implementation of the various sludge concentration and density measurement technologies.

Some of the commercially available analyzers are combined with sludge blanket level detectors, which generally amplify the benefits of automatic sludge monitoring and control.

***Light-Emitting (Optical) Analyzers.*** *PRINCIPLE OF OPERATION.* The operation of light-emitting (optical) analyzers is based on scattering of a beam of light by the suspended particles in the wastewater or sludge (Figure 10.1). The portion of the dissipated light is a function of the number and size of particles and ultimately of the solids concentration. Optical solids analyzers are instruments that measure the dissipated light, transmitted light, or both, and convert these measurements to a solids concentration measurement. Generally, optical analyzers consist of a light source, which emits light of a given intensity, and a photocell, which measures the transmitted light and the degradation of light scatter intensity along the path of the light beam (see Figure 10.1). The actual configuration of the light source and the photocell in the measurement instrument varies with the manufacturer.

Optical solids analyzers can be divided into two groups: (1) turbidimeters, which typically present solids concentrations in nephelometric turbidity units (NTU), and (2) suspended solids concentration analyzers, which indicate solids concentration in milligrams per liter, grams per liter, part per million, or percent. Some equipment manufacturers have combined solids analyzers that can measure both turbidity and solids concentration with the same sensor. To maximize accuracy, manufacturers typically offer different sensors for different solids concentration ranges and type of solids. Typically, optical suspended solids concentration sensors are divided into low, medium, and high solids analyzers.

The solids or turbidity analyzers can be installed in two configurations: (1) submersed, where the sensor is installed directly in the aeration basin or clarifier; and (2)

**Table 10.1** Areas of application of sludge concentration and density measurement equipment.

| Typical applications | Accuracy | Notes |
|---|---|---|
| **Light-emitting (optical) analyzers** | | |
| Sludge density meters: Sludge with solids concentrations from 0.1 to 6% solids.<br>Turbidimeters: Wastewater of turbidity between 0.01 and 10 000 NTU (TSS of 1 to 3000 mg/L). | +/– 0.5% of full instrument span | Avoid use for<br>• Primary sludge higher than 3% solids;<br>• Thickened sludge; and<br>• Wastewater with visibly apparent color. |
| **Ultrasonic analyzers** | | |
| Sludge with solids concentrations from 0.1 to 10% solids | +/– 5% of full instrument span | Avoid use for<br>• Wastewater; and<br>• An MLSS concentration lower than 2000 mg/L. |
| **Nuclear analyzers** | | |
| Sludge with solids concentrations greater than 4% and lower than 15% solids. | +/– 0.5 to 1% of full instrument span. | Avoid use for<br>• Wastewater;<br>• MLSS, RAS, and WAS; and<br>• Low-concentration primary sludge. |

inserted, where the sensor is connected to the sludge conveyance pipe (generally through a valve assembly flange enabling sensor insertion and removal without interrupting the process flow). Submersion-type sensors are supported by a handrail mounting hardware assembly or submersion extension pipe.

*TYPICAL AREAS OF APPLICATION.* Optical solids analyzers can be used for measuring a wide range of solids concentrations of both wastewater and sludge. Turbidimeter-type optical analyzers are most often used for monitoring plant secondary and tertiary effluent turbidity, especially when the plant effluent is applied for water reuse, where effluent turbidity is a regulated water quality parameter. The turbidimeters can operate well in a range of 0.01 to 10 000 NTU. Optical suspended solids analyzers are typically applied for measuring MLSS concentration and less frequently for low-solids primary sludge. Optical type analyzers can be used for measuring

MLSS, RAS, and WAS solids concentrations. They, however, are not recommended for measuring dissolved air flotation sludge, thickened WAS with more than 6% solids, and thickened primary sludge with solids content higher than 3%.

*KEY ADVANTAGES.* Optical analyzers are the most accurate instrumentation for measuring plant effluent turbidity and TSS concentration and low-solids sludge (such as activated sludge). Their accuracy is typically +/− 0.5% of the full measurement scale. Turbidimeters can measure very low turbidity concentrations up to levels of 0.01 NTU. Turbidimeter costs are relatively low, and these units are easy to install, calibrate, and maintain. Optical analyzers are currently the most widely used equipment for measuring activated sludge (MLSS and WAS) solids concentration.

*KEY TECHNOLOGY LIMITATIONS.* Solids and algae buildup and coating of the source of light or the photocell are generally the key problems with optical solids analyzers in wastewater plants. Therefore, their performance and accuracy are highly dependent on the frequent cleaning and calibration of the light source and sensors. Most of the newer generation commercially available optical analyzers are equipped with a self-cleaning assembly and have an optics arrangement that minimizes degrading factors, such as sensor fouling, from interfering with the solids concentration measurements. The system optics is generally protected by scratch-resistant

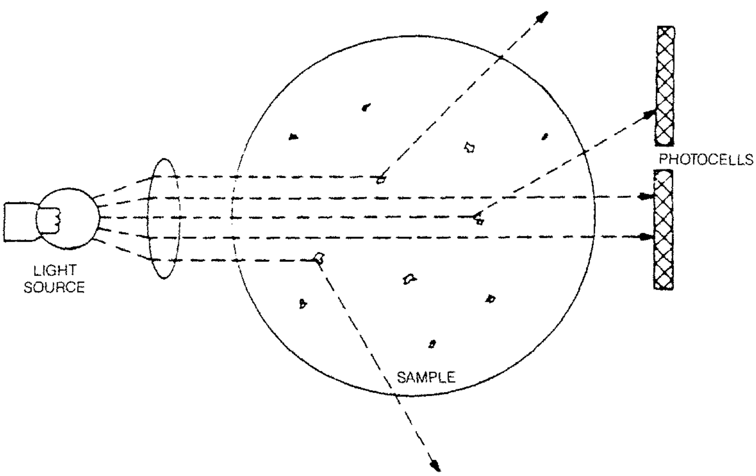


**FIGURE 10.1** General schematic of optical solids concentration analyzer (WEF, 1996).

materials. Some of the analyzers contain built-in provisions, which allow for compensation for measurement errors caused by air bubbles in the sludge or wastewater.

Optical analyzers have limited application for measuring solids in wastewater with visibly apparent color. Performance of this type of analyzer is sensitive to wastewater color. While color might not appear as particulate matter in a suspended solids monitoring application, the optical detector senses it as energy absorbent and reports it as suspended solids. Some of the state-of-the art optical solids analyzers contain provisions to compensate for the effect of wastewater color on instrument readings.

***Ultrasonic Analyzers.*** *PRINCIPLE OF OPERATION.* These types of analyzers include a source of ultrasonic signal and a receiver (transducer). The transmitted ultrasonic signal is dissipated by the particles in the sludge stream proportionally to their concentration, and the attenuated signal is received by the transducer (Figure 10.2). Solids concentration is determined based on the speed of sound movement through the sludge.

Most of the existing commercially available ultrasonic analyzers have simplified calibration procedures, which allow setting the instrument solids concentration

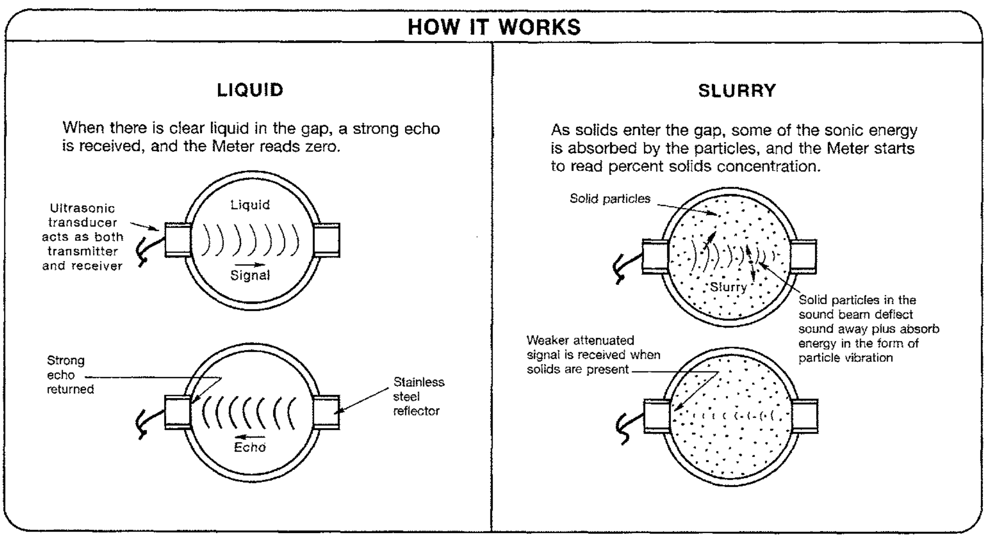


**FIGURE 10.2** General schematic of ultrasonic solids concentration analyzer (published with permission of Markland Specialty Engineering, Ltd., Toronto, Ontario, Canada).

reading to a 0.0 percent solids or milligrams per liter in clean water and then to calibrate in a known solids concentration. The calibrated ultrasonic meter gives linear readings between these two settings.

The ultrasonic analyzers are generally outfitted with self-diagnostic programs continuously monitoring the analyzer's operation for malfunctions, such as broken wires or improper voltage of the generated reading signal, which provide early warning if the measurement equipment malfunctions.

*TYPICAL AREAS OF APPLICATION.* Ultrasonic analyzers are typically used for measuring sludge concentration in a range 0.1 to 10% solids and most accurate for the medium to high end of this range. The exact range of their operation depends on the sonic attenuation of the particular sludge and the length of the sound path. The typical accuracy of ultrasonic analyzers is +/– 5% of the full scale.

Ultrasonic analyzers are most widely used for measuring primary sludge, WAS, and RAS concentrations. They are not appropriate for monitoring plant effluent, MLSS concentration, or plant sidestreams of TSS concentration lower than 1000 mg/L. In inline applications, the ultrasonic solids analyzers are suitable for pipe sizes larger than 100 mm (4 in.) and smaller than 300 mm (12 in.).

*KEY ADVANTAGES.* The ultrasonic solids analyzers are relatively inexpensive devices for measuring medium to high sludge solids content. Opposite to optical analyzers, they are very suitable for monitoring primary sludge concentration, but not for measuring aeration basin MLSS. The key advantage of ultrasonic analyzers, compared to optical sludge sensors, is that their performance is not sensitive to changes in wastewater color and to high concentration of gas bubbles. This makes the ultrasonic analyzers the preferred solids measurement device for dissolved air flotation systems.

*KEY TECHNOLOGY LIMITATIONS.* Similar to optical analyzers, ultrasonic solids analyzers are prone to solids buildup and coating of the ultrasound sensor. Their accuracy is also limited for measuring low density sludge, with solids concentration lower than 0.3%, and very high density sludge, with solids content higher than 10%.

***Nuclear Density Analyzers.*** *PRINCIPLE OF OPERATION.* Nuclear solids analyzers are noncontact density gauges. These devices measure the specific gravity of the sludge rather than its concentration. If the specific gravity of the solids and the water do not change significantly over time (plant influent wastewater is of relatively consistent quality), then the specific gravity measured by the nuclear analyzer can be correlated to the sludge solids concentration.

A key element of the nuclear density analyzers is a radioactive gamma ray source (Cs 137 or Co 60), which is typically contained in a lead-filled, steel-encased housing. A portion of the gamma ray source irradiation is absorbed by the solids in the sludge, and the remaining irradiation is measured by a scintillation detector. The radioactive source and detector are located on opposite sides of the pipe (Figure 10.3).

A continuous focused beam of radiation is transmitted from the radioactive source through the pipe and sludge to the scintillation detector. When the radiation reaches the scintillation crystal of the detector, the analyzer generates a signal that is proportional to the sludge density, which is transmitted to the equipment electronics for conversion to a 4 to 20 mA or other useable process signal. As the density of the sludge in the pipe changes, the amount of radiation reaching the detector varies. The greater the sludge density or concentration, the lower the radiation intensity that reaches the detector, and vice versa.

*TYPICAL AREAS OF APPLICATION.* Nuclear density analyzers are suitable for thickened sludge of high solids content (typically 4% or higher). They cannot be used for measuring plant effluent turbidity or TSS concentration or unthickened primary and secondary sludge. These analyzers can measure sludge density of up to 15% solids content. The nuclear density analyzers can typically be installed on pipes of sizes larger than 150 mm (5 in.) and smaller than 350 mm (14 in.). However, some equipment manufacturers recently introduced nuclear density analyzers that can be used for a wider range of pipe sizes (25 to 1000 mm [1 to 42 in.]).

*KEY ADVANTAGES.* The nuclear analyzer is the only sludge density or concentration measurement device that does not have direct contact with the measured sludge and therefore requires little maintenance. The noncontacting feature of these analyzers makes them very suitable for abrasive, corrosive, high-pressure, and high-temperature applications. This device has no moving parts and is reasonably sensitive to sludge concentration variations. The nuclear analyzers are generally strapped on the sludge conveyance line and can be located practically anywhere along this line (see Figure 10.3).

*KEY TECHNOLOGY LIMITATIONS.* Nuclear density analyzers are relatively expensive solids measurement devices compared to optical or ultrasonic analyzers and cannot measure low-concentration solids streams (below 4% solids) with accuracy

**FIGURE 10.3** Nuclear density analyzer (photo published with permission of ThermoMeasureTech, Inc., Waltham, Massachusetts).

comparable to that of the other two types of analyzers described above. They must be installed and maintained by operators trained and licensed in handling radioactive material (training is generally provided by the analyzer manufacturers).

The nuclear sludge density analyzers can operate properly only if the sludge conveyance line is full. These analyzers are applicable to measuring density of relatively homogenous sludge, which temperature and consistency do not change significantly over time. The nuclear analyzers are not suitable for solids streams with entrained air bubbles (such as dissolved air flotation thickened sludge) and applications at municipal wastewater plants with significant industrial wastewater contributors of frequently changing wastewater characteristics (i.e., cyclic discharges of large amounts of oil and grease, high temperature industrial waste, or great variations in density).

The nuclear density analyzers are very accurate measurement devices (+/– 0.5 to 1% of full instrument scale). However, because the solids concentration readings of these instruments are based on a correlation of the measured specific gravity and the solids content, their accuracy is affected by significant changes in the specific gravity of the measured sludge.

**INSTALLATION OF SOLIDS ANALYZERS.** The specific method of installation of the sludge concentration or density measurement devices varies depending on the type of instrument and recommended manufacturer installation details. The best location of inline sludge concentration measurement devices is on a vertical line with an upflow. The solids concentration measurement device must be installed at a location where the sludge is well-mixed and accurately represents the actual concentration. The operation range of the instrument must match the range of the measured solids concentration. Measurement devices must be located in such a manner that they are easy to access and maintain.

Generally, inline solids analyzers must be calibrated weekly. The analyzer sample lines (if used) must be flushed weekly and their flow checked daily to maintain consistent instrument performance and accuracy. The analyzer probe must be easily removable for service without shutting down process piping and disturbing the operation of the sludge pumping system (Figure 10.4). Depending on the skills of the plant staff, inline solids analyzer calibration and maintenance may require a total of 3 to 6 hours per unit per week.

Sludge sample lines must be large enough to prevent plugging. It is recommended to provide a flushing tap next to the instrument and a sample box, so that samples can be collected manually at the point of instrument installation for calibration purposes.

For large wastewater treatment plants, separate measurement devices are recommended to be installed on the sludge withdrawal lines from the individual clar-

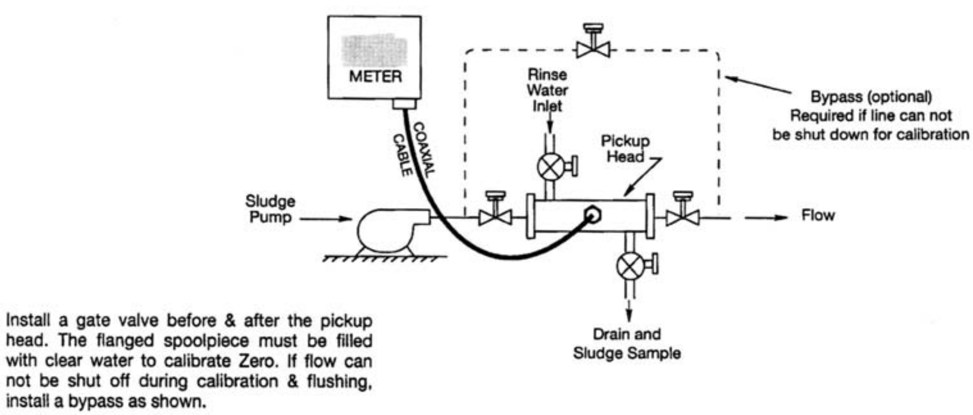


**FIGURE 10.4** Installation of sludge solids concentration analyzer (published with permission of Markland Specialty Engineering, Ltd., Toronto, Ontario, Canada).

ifiers to gain a better control over the operation and performance of these units. Sludge density measurement devices must be installed coupled with sludge flow measurement devices. The displays of the sludge concentration and flowrate measurement instrumentation must be located adjacent to each other for direct observation and comparison.

Inline solids analyzers are generally used for measuring primary sludge, WAS, and RAS concentration. When measuring MLSS concentration in aeration basins, analyzer sensors are directly immersed in the basins and mounted on holders off the basin walls. If a wall-mounted optical solids concentration analyzer is used, the sensor should be immersed at least 0.04 m (1.5 in.) below the activated sludge tank water surface and should be located as a minimum 0.15 m (6 in.) away from the aeration basin wall. If the wall is bright and reflective, the distance from the sensor to the aeration basin wall should be at least 0.3 m (12 in.). Installing the optical sensor too close to a wall can cause infrared light backscatter, resulting in a higher intensity signal. Optimum self-cleaning of immersed suspended solids analyzers is achieved by turning the sensor surface into the flow direction.

**SLUDGE BLANKET DEPTH MEASUREMENT.** Sludge blanket depth is a key indicator of primary and secondary clarifier performance. The depth of the sludge blanket is the distance from the clarifier surface to the blanket top. The blanket thickness is the distance from the top of the sludge blanket to the bottom of the clarifier. The sludge blanket typically varies daily within certain predictable limits, as a result of diurnal flow fluctuations. The blanket depth may also vary as a result of process changes induced by plant operators. Day-to-day fluctuations of clarifier sludge blanket in a plant operated under relatively stable conditions are relatively slow and are typically limited to within 0.3 to 0.6 m (1 to 2 ft). Significant and abrupt changes in sludge blanket depth in clarifiers are typically caused by either a large increase in influent flow (transient flow conditions) or by a stoppage or malfunction of the sludge collection and/or withdrawal systems.

In primary clarifiers, sludge blanket depth is one of the main parameters that triggers initiation of sludge withdrawal. In secondary clarifiers, this parameter can be influenced by a number of activated sludge system performance changes, and its fluctuations over time provide critical information for the overall health of the activated sludge system. Therefore, sludge blanket depth is one of the most frequently monitored parameters in wastewater treatment plants.

***Manual Sludge Blanket Measurement.*** Currently, sludge blanket level at full-scale treatment plants is most commonly determined by manual measurements using calibrated clear plastic tube (also named *core sampler* or *sludge judge*). To take a reading, the operator lowers the tube into the clarifier while holding a valve located at the bottom of the tube open. After reaching the tank bottom, the operator closes the valve and carefully removes the tube, which, at this time, is filled with solids to a particular level. Some samplers (i.e., the sludge judge) have a ball check valve at the bottom of the tube that is open as it is lowered and closes when it is raised. If the sample has been collected correctly, the depth of sludge in the tube is the same as the depth of the sludge blanket in the clarifier at the sampling location.

The key disadvantage of manual sludge blanket depth measurement is that it is a discrete sample measurement that gives only a snapshot representation of the sludge blanket level at a given time and location. Variables, such as the sampling location and time, location of the sludge collection mechanism at the time of the measurement, speed of tube descent, ambient light conditions, and subjectivity of operator readings and sampling skills, contribute to the sometimes limited benefits of the manual sludge blanket measurement.

One key advantage of the manual plastic tube sampler is that it also allows for collection of a sludge sample in which TSS concentration is representative of the average solids concentration of the sludge blanket, a parameter which could be used to calculate the sludge blanket SRT and, ultimately, to determine the optimum sludge withdrawal rate. Manual plastic tube samplers are reliable, inexpensive, practically do not require maintenance, and can be easily replaced if damaged. In addition, one manual plastic tube sampler can be used to monitor multiple clarifiers.

Another type of manual equipment for sludge blanket depth measurement is sight glass. This type of sludge blanket finder consists of a sight glass and light source attached at the lower end of a graduated peace of aluminum pipe approximately 38 mm (1.5 in.) in diameter. The sight glass is carefully lowered into the clarifier through the zones of clear liquor and individual particles, until the top of the homogenous sludge blanket is observed.

For small plants and plants with clarifiers where the sludge blanket does not vary significantly over time, manual sludge blanket depth measurement is a generally adequate, simple, and low-cost method for determining sludge blanket depth.

***Automated Sludge Blanket Level Measurement.*** In medium and large wastewater treatment plants with more complex activated sludge and solids handling systems, installation of instrumentation for continuous sludge blanket measurement interlocked with automated control of primary and secondary sludge withdrawal systems warrants consideration. The use of sludge blanket level monitoring and control systems minimizes solids handling costs by reducing the volume of water pumped and processed with the sludge. Dakers (1985) has found that sludge volume could be reduced by approximately 50% if automated sludge blanket control is used instead of manual clarifier desludging. A survey of automatic sludge removal systems in the United Kingdom (Burke et al., 1985) also points out that primary clarifier sludge concentration could be increased up to two times when clarifiers are desludged automatically by blanket level control. The benefits of automated sludge blanket depth measurements for WAS and RAS flowrate control have also been documented at a number of full-scale wastewater treatment plants (Bush, 1991; Dartez, 1996; Ekster, 1998 and 2000; Hinton-Lever, 2000; Hoffman and Wexler, 1996; Rudd et al., 2001; Samuels, 2000).

Automated sludge blanket depth measurement is recommended for consideration for wastewater treatment plants with significant variations of diurnal influent water quality and quantity and associated frequent shift of sludge blanket levels. In cases where sludge blanket level fluctuations are frequent (blanket level changes up and down several times per day with more than 0.32 m [1 ft]) per change) and clarifiers are relatively shallow (i.e., clarifiers with a sidewater depth of less than 3.66 m [12 ft]), the use of variable frequency drive (VFD) motors for the sludge withdrawal pumps is recommended for consideration. If VFD-controlled motors are used, sludge blanket monitoring instrumentation and pump control equipment operation can be interlocked to automatically adjust the clarifier sludge pump withdrawal rate to keep the sludge blanket at an optimum, near-constant level.

*SLUDGE BLANKET LEVEL DETECTORS.* Most of the commercially available sludge blanket level detectors are based on ultrasonic or optical measurement of sludge concentration. These devices generally have provisions for compensating sensor measurement for temperature, fouling, and aging.

*ULTRASONIC SLUDGE BLANKET LEVEL DETECTORS.* Typically, ultrasonic sludge blanket detectors have a sensor located just below the water surface of the clarifier, which continuously emits pulses of ultrasonic energy (Figure 10.5). These pulses are reflected in the form of echoes from suspended solids layers in the clarifier, detecting the interface between the light solids in the clarification zone and the sludge blanket. The blanket level analyzer then digitally converts the round-trip time of each pulse to compute the sludge blanket level and depth. The level is displayed

**FIGURE 10.5** Ultrasonic sludge blanket level detector (published with permission of Siemens-Milltronics, New York).

numerically and can be transmitted to the plant process monitoring and control system. Ultrasonic level analyzers can also be used to develop a profile of the sludge concentration throughout the clarifier. They can measure sludge blanket levels between 1 and 11 m (3.3 to 36 ft). The accuracy of the ultrasonic sludge blanket level detectors is typically +/− 1% of the reading, and signal resolution is 0.03 m (0.1 ft).

*OPTICAL SLUDGE BLANKET LEVEL DETECTORS.* Optical sludge blanket level measuring systems include a pulsed infrared light sensor immersed below the clarifier surface and attached to a cable driven up and down by a stepper motor equipped with worm gear (Figure 10.6).

Clarifier zones of different solids density are detected by measuring suspended solids concentration based on infrared light absorption. The optical sensor generates a signal proportional to the concentration of solids in suspension, which is converted to a frequency signal. The measured signal is compared with a preselected reference value for sludge concentration in the measuring transmitter. If there is a deviation, the sensor is moved up or down by the stepper motor until it reaches the sludge blanket zone of a particular concentration targeted for measurement. The optical sludge blanket measurement device determines the sludge blanket level from the number of steps carried out by the stepper motor and converts the result to an analog signal. Similar to ultrasonic analyzers, optical sludge blanket level detection systems can be used to measure blanket levels between 1 and 11 m (3.3 to 36 ft), with an accuracy of +/– 1% of the measured value.

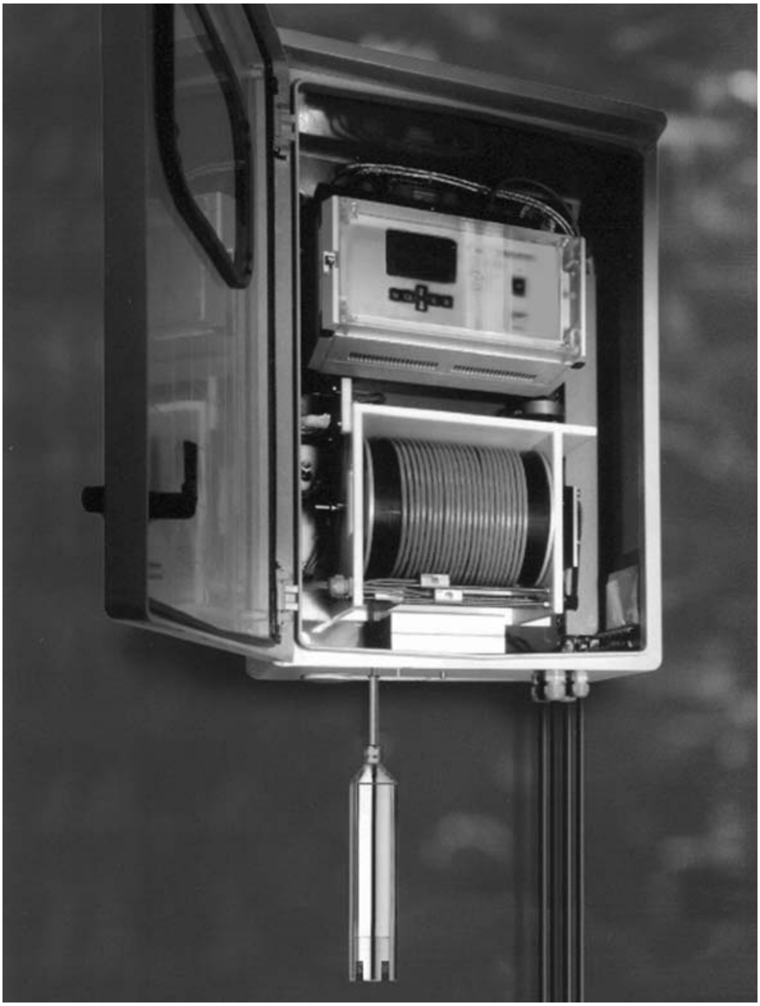

**FIGURE 10.6** Optical sludge blanket level detector (published with permission of Endress and Hauser, Greenwood, Indiana).

*TYPICAL AREAS OF APPLICATION.* Sludge blanket level analyzers are typically used to track blanket depth shifts to prevent clarifier performance failures and maximize sludge concentration. The same analyzers can be used in both primary and secondary clarifiers.

Ultrasonic sludge blanket level detectors have found a wider application to date, generally because of their extended capabilities to produce continuous clarifier solids density profiles and because of their lower cost. For a typical, well-settling sludge, ultrasonic analyzers produce very consistent readings. However, these analyzers may not be as accurate as infrared optical analyzers at plants experiencing slowly settling or frequently bulking or floating sludge, because these types of sludge do not have a well-defined ultrasonic echo. Under such conditions, optical sludge blanket level detectors would be a better choice.

The automated sludge blanket level measurement instrumentation is generally interlocked with the sludge withdrawal pumps, which are activated automatically when the sludge blanket reaches a certain level and automatically shuts down when the sludge solids decrease below a certain target concentration. Compared to manual sludge withdrawal, the use of sludge level control allows one to avoid pumping low-solids sludge during periods of low flows and light solids loading and to automatically stop sludge pumps when their suction begins creating funnels ("rat holes") in the clarifier sludge blanket.

The newest generation of sludge blanket monitoring systems are microprocessor-based and can be easily integrated in the plant centralized monitoring and control system. Some of the commercially available sludge blanket level analyzers provide a graphical representation of the suspended solids profile in the clarifier and alarm indication when sudden changes of the sludge blanket level occur (Figure 10.7). The level signal from the sludge blanket level analyzers can be transmitted to the plant control system and the main control room for a direct visual monitoring by the operators on duty.

*KEY TECHNOLOGY LIMITATIONS.* The benefits of continuous sludge blanket monitoring are less pronounced for small treatment plants with relatively small variations of plant influent flow and waste load and relatively deep clarifiers that have available large sludge retention volume and can carry significant blanket fluctuations.

Ultrasonic sludge blanket level analyzers are subject to "blinding" by gas bubbles. In primary clarifiers, gas bubbles are typically created by the septicity of the

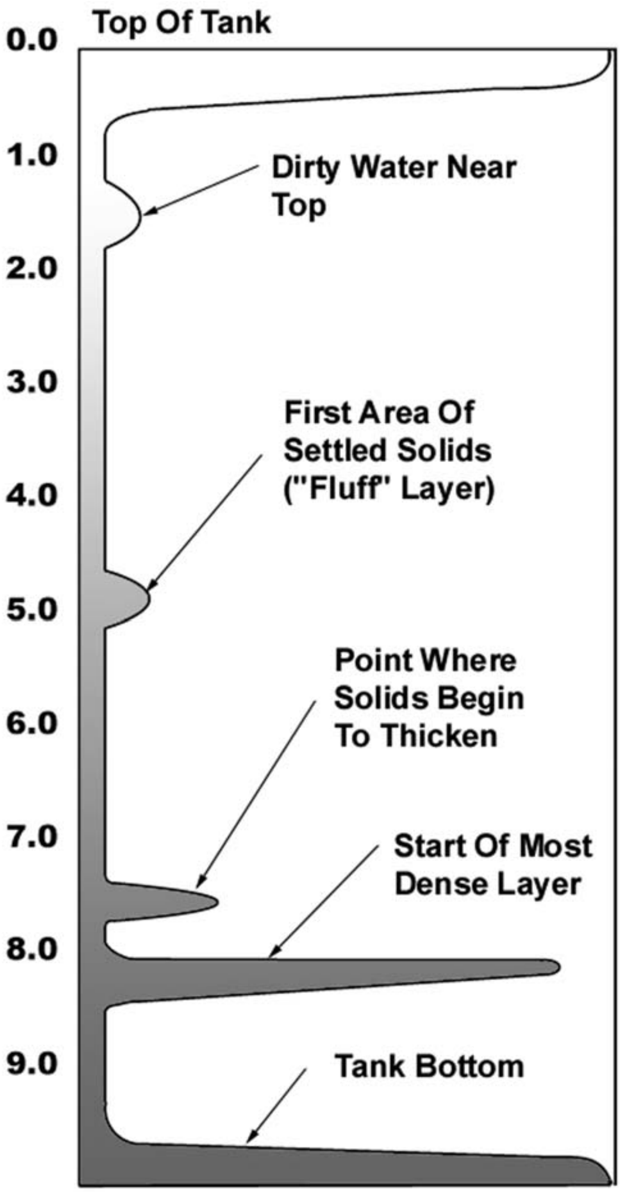


**FIGURE 10.7** Sludge blanket profile (published with permission of Royce Technologies, New Orleans, Louisiana).

primary sludge, while, in biological nutrient removal systems, nitrogen gas bubbles are generated as a result of uncontrolled denitrification in the secondary clarifiers. The gas bubbles, when trapped on the surface of the sonic sensor, alter the sensor readings. Therefore, the ultrasonic sensors must be designed with cleaning provisions. Generally, small utility water pumps are installed on the rail above the sensor or right on the sensor. These pumps typically run intermittently and wash the sensors to maintain accuracy of the instrument readings.

The use of optical sludge blanket level detectors is limited by their higher costs and relatively lower accuracy. Optical analyzers are subjected to interference by accumulation of solids on the analyzer sensor and by light reflection from nearby objects (smooth walls and sunlight reflecting tank and equipment surfaces).

*INSTALLATION OF SLUDGE BLANKET LEVEL DETECTORS.* Blanket level detectors must be installed in locations that do not cause interference with the normal operation of the sludge collection and removal system. Generally, the stationary sludge blanket meters are installed on the catwalk or on the side rail of the clarifiers (Figure 10.5). The stationary ultrasonic sludge blanket sensors are typically mounted 4 to 8 cm (1.5 to 3 in.) below the liquid surface. They are equipped with skimmer guards to protect the sensors from damage.

The best location for measuring sludge blanket depth is where the actual depth is equal to the average clarifier depth. In circular clarifiers, this point is typically one-third of the distance from the outside wall of the clarifier center to the middle. In rectangular clarifiers, the most appropriate location of routine sludge blanket measurement is typically at the midpoint of clarifier basin length. Because clarifier configuration, type, and size vary, the most representative location for measurement of the average sludge blanket depth is recommended to be established based on a series of manual sludge blanket measurements at three to five locations along the clarifier radius or length.

Typically, sludge collection arms of circular clarifiers rotate approximately once every 15 to 30 minutes, and the sludge collection mechanism (scraper or suction header) movement disturbs the sludge blanket. If the sludge blanket is measured manually, the sludge blanket depth readings are recommended to be taken when the sludge collection mechanism (bridge) is perpendicular to the measuring location. Taking the sludge blanket level measurement at this location minimizes the effect of sludge collection mechanism movement on the measurement. When sludge blanket level measurements are collected manually, at least two measurements have to be completed: one at the sludge collection arm at 90 degrees, and one at 270 degrees from the bridge. These two measurements have to be averaged to determine the average sludge blanket depth.

Automated sludge blanket level analyzers typically take continuous (several times per second) interface level readings. This enables the operating staff to observe the sludge blanket behavior in real time. Blanket depth measurement instrumentation can produce an "average" sludge blanket level or interface level by averaging the sludge profile at preset intervals of 15 to 60 seconds, which eliminates wide changes in the blanket level readings caused by sludge collection rake passage or temporary short-term sludge blanket upsets.

Individual automated sludge blanket level analyzers are recommended to be installed in all clarifier units of the wastewater plant rather than to install a blanket level detector in only one clarifier and use the detector reading to judge the sludge blanket levels in all the other plant clarifiers. Comparison of sludge blanket behavior of the individual units allows one to identify and promptly address potential problems related to uneven flow distribution among the clarifiers, malfunction of clarifier sludge collection and withdrawal systems, or other site-specific events that cause individual clarifier units to perform differently. For example, Figure 10.8 shows sludge blanket profiles of two identical clarifiers at the same plant performing differently at the same time. The vertical axis of this figure indicates the clarifier depth. The depicted tank is 4.3-m (14 ft) deep. The "zero" level depicts the top of the tanks (shown at the bottom of the figures). The horizontal axis represents the solids density along the depth of the clarifiers.

Clarifier units 1 and 2 receive the same sludge at the same rate. Both clarifiers have sludge blankets at almost the same depth. However, the sludge blanket profile indicates that clarifier unit 2 performance is inferior, and this clarifier is experiencing a solids washout. In this particular case, the washout was caused by malfunction of the sludge withdrawal pumps.

*SELECTION OF MONITORING EQUIPMENT.* Selection of the most appropriate instrumentation for the specific application is critical for the reliable monitoring and control of clarifier solids concentration and sludge blanket. Most sensors perform well under ideal conditions that manufacturers use to determine their specifications for accuracy, reproducibility, and other key operational parameters. How-

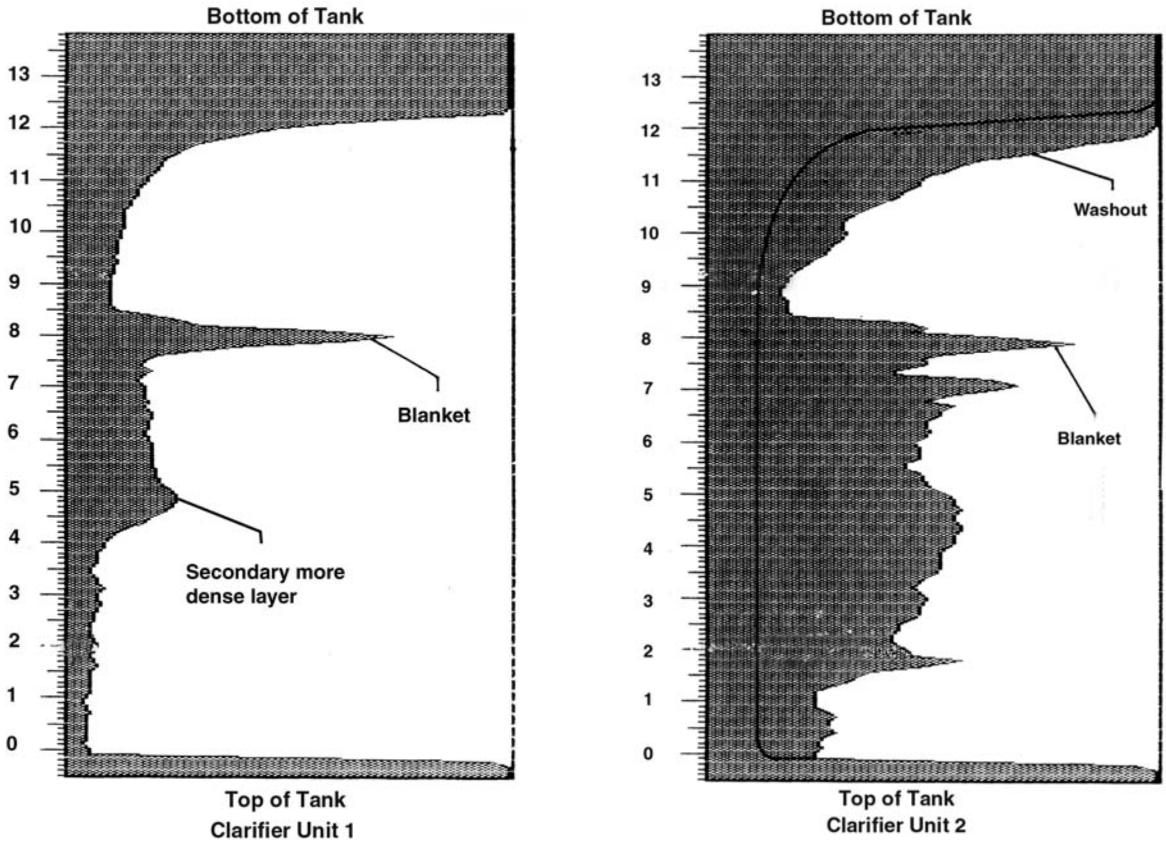


**FIGURE 10.8** Comparison of sludge blanket profiles in two identical clarifiers (published with permission of Royce Technologies, New Orleans, Louisiana).

ever, sensor performance in the field is sometimes unsatisfactory and requires a period of calibration and adjustment to the site-specific conditions of the application (Hill et al., 2001). Instrumentation field testing is invaluable in providing the information needed to select the most appropriate type and model of equipment for a given application. Although on-site testing by the end user is the most reliable approach to select the best monitoring system, such testing could be costly and time-consuming. Therefore, the extensive testing experience and equipment performance assessment information of organizations specialized in evaluation of water and wastewater treatment plant monitoring equipment, such as the Instrumentation Testing Association, Henderson, Nevada, are recommended to be used to aid and expedite the instrumentation selection process.

# CASE STUDIES

**CASE STUDY FOR ACTIVATED SLUDGE SOLIDS INVENTORY MONITORING—SAN JOSE/SANTA CLARA WATER POLLUTION CONTROL PLANT, CALIFORNIA.** The 632 000 $m^3$/day (167 mgd) San Jose/Santa Clara Water Pollution Control Plant in San Jose, California, uses online equipment for measurement of secondary clarifier WAS and MLSS for an automated waste activated sludge control (Ekster, 1998 and 2000). The automatic WAS removal rate/SRT control system provided efficient real-time control over solids inventory, which allows one to substantially reduce activated sludge bulking and Nocardia foaming, improve effluent quality, and decrease chemical usage for phosphorus removal and disinfection. In addition, the use of the automated WAS control system reduced routine manual clarifier sludge grab sampling and analysis by 80%. The estimated payback period for the automatic waste control system implementation was less than three years.

The automatic waste control system (see Figure 10.9) consists of analyzers for measuring MLSS concentration in the aeration basin and the WAS concentration, a controller (computer), flowmeter, and motorized control valve installed on the WAS line. The sludge concentration signal from the analyzers is transmitted to the controller. The controller compares the operational criteria (MLSS concentration or SRT) with their target values, calculates the necessary adjustment of the WAS flowrate, and sends a control signal to the motorized valve located on the WAS line.

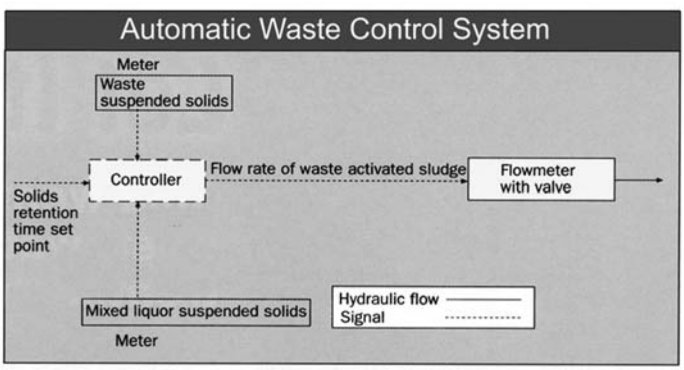


**FIGURE 10.9** Schematic of automatic waste control system (Ekster, 1998).

The installation of the automated SRT control system resulted in a significant reduction of WAS flowrate variation. Fluctuations of RAS and MLSS concentrations were reduced by almost three times, while WAS flowrate variation was reduced by a factor of seven. The significant reduction in WAS flowrate variation improved sludge thickener operation, and the earlier practice of bypassing caused by overload of thickeners become unnecessary. Polymer usage for sludge dewatering was also reduced by approximately 50% because of the improved activated sludge dewaterablity (elimination of the Nocardia bulking sludge problem).

Plant operators can select and adjust the target SRT setpoint, wasting strategies, and limits of system operation. The controller can handle both continuous and intermittent wasting strategies. Continuous wasting can be set to maintain either a stable flow or a stable load, depending on the capacity of the secondary clarifiers. Intermittent wasting can be set up with either fixed pump time schedules with flowrate control or fixed flowrate operation with variable pump time controls. Pump schedules for the entire week can be programmed, with up to four start and stop times each day. The operator can safeguard the control process by specifying the range of allowable MLSS, flowrate, or pump times and load values. The control system alerts the operator if any of these ranges need to be exceeded to maintain a preset optimal SRT. It also generates suggestions as to what ranges might be changed for improved control. If waste flow or wasting time has to be temporarily changed, for example, in the case of a dangerously high sludge blanket level, the operator can override the controller. The mass of wasted sludge in the manual mode is also included in the controller calculations when automatic waste mode is reinstated. When the SRT target value needs to be changed, the controller is programmed to alter the value gradually, eliminating the possibility for process upsets. A change in SRT target is generally considered in cases of temperature change, in anticipation of shock loading, a significant change in wastewater characteristics or sludge settling characteristics.

When an aeration tank is taken offline, it is relatively simple to alter the controller to compensate for the reduced process volume. The SRT controller induces stability and reliability and detects whether system elements are operating correctly. The operator is alerted if MLSS, RAS, or WAS flowrate values change in an unusual manner.

On several occasions, the reliability of the automated SRT control system was challenged when the RAS flow from the sampling sink was interrupted, and, as a result, the RAS sludge concentration analyzers produced invalid readings. The control system automatically detected problems with the RAS sludge concentration analyzer readings, activated warning alarms, and automatically changed the control algorithm to ensure that the faulty readings were disregarded. Once conditions returned to normal, the controller automatically changed the control algorithm back to the one used before the problems occurred.

**CASE STUDY FOR SLUDGE CONCENTRATION MONITORING—CLARK COUNTY SANITATION DISTRICT, LAS VEGAS, NEVADA.** The 333 080-$m^3$/d (88-mgd) Clark County Sanitation District biological nutrient removal plant in Las Vegas, Nevada, consists of eight 41 635 $m^3$/d (11-mgd) aeration basins, each with a dedicated secondary clarifier. The plant staff has successfully automated the secondary clarifier waste activated sludge withdrawal system, achieving improved nutrient removal coupled with cost savings or reduced operator attention and laboratory time, and ferric chloride use (Bain and Johnson, 1998). Taking under consideration all savings achieved by fully automating activated sludge system solids inventory control, the payback period of the improvements was less than three years.

Before implementation of the automated WAS withdrawal system, plant staff followed a conventional procedure of controlling activated sludge solids inventory by collecting grab samples of MLSS and RAS, analyzing MLSS and RAS concentrations, and calculating the new target wasting rate as follows:

$$\text{New WAS removal rate} = \text{Current WAS removal rate} \times \sqrt{(\text{current MLSS concentration}/\text{target MLSS concentration})}$$

The applied formula is semiempirical, and the square root factor is used to introduce a multistep gradual adjustment of the sludge inventory. The new WAS removal rate was set manually by adjusting the WAS pump discharge line valve position.

However, manually adjusting the MLSS concentration through WAS control has proven to be difficult because of the significant variation of the MLSS and RAS concentrations of the grab samples. The MLSS grab sample collected anytime between 6:00 and 9:00 a.m. could vary by +/− 15 %, and the RAS concentration could vary by +/− 40%. Because of the large variation of grab sample values, the manual control of the activated sludge system solids inventory was not producing consistent nitrogen and phosphorus removal. In addition, the manual WAS control procedure required substantial operator time and, for practical reasons, could not be performed more than two times per day. Therefore, the plant staff began seeking a way to obtain real-time MLSS data, with the ultimate goal of automatically controlling MLSS through online instrumentation.

The first step of the plant automation strategy was to install a solids concentration analyzer in one of the aeration basins. After being moved to several locations in the basin, the solids concentration analyzer was placed in the oxic zone of the aeration basin, 1.2 m (4 ft) from the surface, far enough from excessive air turbulence, but in a zone of mild agitation, where minimum velocity of 0.6 m/sec (2 ft/sec) could be maintained across the probe. This location was found to be optimal in terms of fouling and needed frequency of analyzer probe cleaning. For several months, plant staff monitored actual MLSS using the online solids concentration analyzer, compared concentrations with the target MLSS, and made manual WAS line valve adjustments.

After gaining comfort with the MLSS concentration analyzer, the staff discontinued daily grab sampling and laboratory analysis. The plant control system computer loop was modified to actuate the WAS flow valve, eliminating the need for manual adjustment. Based on a target MLSS set by the plant staff, the solids concentration analyzer signals the WAS valve to adjust its position, allowing automated continuous WAS wasting and MLSS control.

Automated online solids concentration analysis and MLSS control has proven to be superior to manual control in maintaining MLSS in the aeration basins near target levels. After several months of testing, automated solids analyzers were installed in all seven active aeration basins, and the entire plant was switched to automated MLSS control. Comparison of ammonia and phosphorus removal from total plant flow before and after implementing the automated MLSS control system shows improved plant performance, with ammonia and phosphorus levels well below permit limits.

As a further step towards improved plant performance, the plant staff replaced automated MLSS control with automated SRT control strategy, installing RAS concentration analyzers for all aeration basins. The SRT control strategy is better than targeting only constant MLSS control, particularly for situations in which extreme flow variations flush solids through the aeration basin. This control strategy aims to maintain target constant activated sludge SRT. The SRT is calculated as kilograms (pounds) of solids under aeration (based on automated MLSS concentration measurements), divided by kilograms (pounds) of solids wasted per day (based on automated RAS concentration measurements).

The experience at the Clark County wastewater treatment plant indicates that the use of online suspended solids analyzers can greatly enhance process control in activated sludge systems and improve the reliability of the biological nutrient removal processes. In addition, plant staff has determined that the overall costs of probe installation, operation, and maintenance are significantly lower than manual sample collection, laboratory analysis, and manual WAS control.

**CASE STUDY FOR SLUDGE BLANKET DEPTH MONITORING—LUMBERTON, TEXAS.** At Lumberton no. 2 wastewater treatment facility, Texas, located approximately 145 km (90 miles) east of Houston, a sludge blanket meter was installed on the plant's primary clarifiers (Duncan, 2000). The clarifiers are circular tanks with a unit volume of 1120 $m^3$ (295 000 gal). The sludge blanket meter is suspended on a rod and is located just below the clarifier's water level. The sludge blanket meter control unit is located near the meter sensor of the clarifier catwalk. The rod is hinged so that it does not interfere with the rotation of the clarifier rake mechanism.

The sludge blanket meter uses an ultrasonic technology. The ultrasonic transducer continuously emits sonic pulses that are reflected in the form of echoes from the sludge blanket, thereby detecting the interface between light fluids and sludge. A built-in microprocessor digitally converts the round-trip time of each pulse and uses it to calculate distance. The dual-point unit can track levels and control sludge pump functions in the two clarifiers. Clarifier level readings are displayed numerically and can be sent directly to the plant's control system. The sludge blanket depth monitoring unit also creates a composite graphic of the clarifier solids profile. Interlocking the sludge blanket depth monitoring unit with the primary sludge pump controls allowed for improving clarifier performance and increasing the average concentration of the sludge removed form the clarifiers from less than 1 to over 2% solids.

**CASE STUDY FOR SLUDGE BLANKET DEPTH MONITORING—ASHBRIDGES BAY WASTEWATER TREATMENT PLANT, TORONTO, CANADA.** The Toronto's 818 000-$m^3$/d (180-mgd) Ashbridges Bay wastewater treatment plant is the largest wastewater treatment facility in Canada. This plant has successfully applied sludge blanket level detection equipment and radio technology system for automatic operation of plant's primary clarifiers (Rudd et al., 2001). The Ashbridges Bay plant has moving bridge rectangular primary clarifiers. The clarifier control system uses ultrasonic sensors installed on the clarifiers' moving bridge.

The sludge blanket level system generates graphical profiles of the clarifiers' sludge blanket. The blanket level analyzer emits ultrasonic pulses at a frequency of four times per second. These pulses reflect from the various zones of solids concentration in the clarifier and are received by the instrument's detector. The sludge blanket instrumentation produces an echo profile of the clarifier sludge zones. The treatment plant operators use the clarifier sludge blanket profile to make process decisions.

Initially, the ultrasonic sludge blanket measurement system had occasional difficulties with accurately representing the sludge blanket profile because of offgassing problems. Gases generated in the clarifiers resulting from sludge septicity partially dissipated the ultrasonic sound of the instrumentation and thereby interfered with the normal operation of the sound detectors. Gas bubbles also accumulated on the sensors' surface, disabling the instrument readings from time to time. The problem with gas bubble accumulation of the sensor was resolved by installing the instruments on a two- to three-degree angle from the vertical direction. This allowed for the bubbles to roll over the sensor surface while the tank bridge is moving.

The signal from the ultrasonic sludge detectors is emitted via radio modems to the plant control system, which allows the plant operations staff to monitor the performance of the clarifier sludge collection and withdrawal equipment. This monitoring instrumentation also detects if the sludge collection mechanisms and sludge withdrawal pumps are functioning properly. In addition, the sludge blanket monitoring system allows optimizing the performance of the plant solids handling system by maximizing the concentration of the primary sludge pumped to the anaerobic digesters. The average primary sludge concentration was increased from 0.5 to 2% solids through the implemented sludge inventory monitoring and control improvements.

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# INTRODUCTION

The International Association on Water Quality's Scientific and Technical Report No. 6 (STR No. 6)—*Secondary Settling Tanks: Theory, Modelling, Design and Operation*—is an excellent summary of secondary settling tank design (Ekama et al., 1997). Therefore, the purpose of this chapter is to augment, not duplicate, the efforts of the STR No. 6 authors.

Apart from site-specific innovations, design practice worldwide is similar, with differences being in emphasis rather than philosophy. Most of the regional differences between settling tanks are either in the equipment fitted into the tanks, including sludge removal and inlet arrangements, or a preference for rectangular or circular tank designs.

**UNITED KINGDOM HISTORY AND DEVELOPMENT OF CLARIFIERS (STANBRIDGE, 1976).** The earliest forms of clarifiers in the United Kingdom were excavated flow through pits in Edinburgh, Scotland. These are dated circa 1829 and used to settle raw wastewater. The sludge was periodically dug out, dried on the banks of the pits, and then carted away for land application. Subsequently, flat-bottomed earth or clay-lined reservoirs were constructed for the same purpose before land treatment of the settled wastewater.

The use of lime as a precipitant for primary sedimentation was specified in a patent application as early as 1846. Lime addition was used before sedimentation in iron, earthwork, clay, and pitched stone tanks until the early 20th century. However, with chemical precipitation, tanks of a more robust construction were necessary, and rectangular tanks began to be constructed from brickwork or concrete. Both batch and continuous operations were used, although ultimately continuous flow settlement was recognized as more effective and less labor-intensive.

The early continuous flow tanks were rectangular horizontal, often with interior transverse baffles to induce zigzag flow patterns to increase flow path, and were often operated in series. These were manually desludged by squeegee, sometimes using a crane and bucket. A horizontal mechanical scraper was patented in 1864 to be followed in 1929 by a power-driven traveling bridge scraper for radial upward flow tanks. Desludging techniques developed to include sloping tank floors to screw pumps running along the bottom of the tank and bucket elevators to lift the sludge.

The upward flow tank originated in Dortmund, Germany. The tank was deep with a small footprint, sludge removal was easier, and the sludge was of higher dry solids content and reduced volume. With the introduction of mechanical scrapers, it was largely replaced by horizontal or radial flow tanks

**WATER INDUSTRY TRENDS AND THEIR EFFECT ON DESIGN PRACTICE.** ***Changes in Design Practice.*** Over the past few years, the side water depth of secondary clarifiers in the United Kingdom has become deeper and the floor slope shallower. For example, a typical secondary clarifier floor slope was between 12 to 20 degrees with a 2-m side water depth. The typical tank design in the United Kingdom is currently for a 3-m sidewall depth with a 7.5-degree floor slope. This said, design and build contractors have installed tanks with 3.5- to 4.5-m sidewall depths with flat to 4-degree floor slopes.

This change brings United Kingdom design practice in line with continental design practice.

***Current European Focus.*** Because of high energy costs, Europeans are particularly aware of the operating and maintenance cost of systems. A major consideration in equipment selection addresses these lifetime costs and has driven many of the utilities to purchase their systems via design–build–operate (DBO) routes. These DBO projects promote innovative and efficient solutions, where conservatism had previously hindered innovation (Voigt, 2002).

***Staffing Levels.*** Typically, privately operated wastewater treatment plants have less staff and fewer instruments. For example, in the United Kingdom, staffing levels on wastewater treatment works were traditionally high, often with staff housing on-site, and multiple generations used to operate and maintain the works. The operators took great pride and ownership of the works and frequently produced excellent effluent quality from what would theoretically be considered an overloaded works. The old role of a works chemist and on-site laboratory facilities allowed for early identification and remediation of potential process problems.

Since privatization of the water companies, staffing levels have been drastically reduced and massive capital spent to upgrade the works, and wide-scale remote monitoring was implemented to replace the labor force.

***Planning Issues.*** Wastewater treatment plants were traditionally located at a reasonable distance from residential areas. With the high population density in Europe, there has been increasing housing development around previously isolated wastewater treatment works. This has led to the requirement to cover unit processes and extract and treat foul air. A limit of five odor units at boundary or no nuisance at the nearest receptor is common. The settling tank design has to be adapted to accommodate these constraints, and requirement for covering or enclosing tanks with buildings for odor control can lead to preference for small footprint solutions.

Additional planning constraints have limited traffic movements for both construction and sludge disposal and require intelligent and diverse solutions from design engineers.

***Legislation.*** National and European legislation is continually demanding improved effluent quality. The current drivers in the United Kingdom include the Directive on Urban Waste Water Treatment, Groundwater Regulations, Directive on Freshwater Fish, River Quality Objectives, Directive on Bathing Waters, Directive on Shellfish Waters, and Habitats and Birds Directives. Some of these directives are statutory; some are best practice guidelines which potentially could become statutory (Warn, 2002). Investment in tertiary treatment for effluent polishing is often seen as a more robust solution than optimum settling tank design.

***Consents.*** Many European discharge permits are based on random grab or spot samples. The permit typically states a maximum value and a 95th percentile value. These are applied to suspended solids, biochemical oxygen demand (BOD), chemical oxygen demand (COD), and ammonia.

In Germany, one violation incurs an immediate fine at significant cost (Voigt, 2002). In the United Kingdom, the head of the company is held responsible and could receive a prison sentence.

The United Kingdom Royal Commission's "general standard" (1912) was for effluent quality of 30 mg/L suspended solids, 20 mg/L BOD, and is still used as a consent for many wastewater treatment works discharges. However, the above legislation has led to more stringent consents for many urban wastewater treatment works and nutrient removal (ammoniacal nitrogen, nitrate, and phosphorus), and COD and metal limits are frequently imposed. For example, 5 mg/L BOD and 2 mg/L ammonia-nitrogen are not uncommon standards for discharge into freshwaters.

These constraints have led to technological development of upstream processes and a necessary appreciation of the effect on sludge characteristics.

The disinfection requirements for the Directive on Bathing Waters demands UV transmissivity limits, and, as a consequence, tertiary treatment following secondary clarification is often implemented, as are alternative processes such as membrane bioreactors negating the requirement for both primary and secondary clarification.

The regulatory requirements for monitoring and audit of UV disinfection are so onerous that frequently membranes offer a lower whole life cost solution.

***Conditions.*** The sewerage systems in Europe are typically combined systems, receiving rainfall and groundwater runoff in addition to the foul wastewater loads from domestic and industrial sources. The hydraulic resilience of both the sewerage systems and wastewater treatment plants is required to accommodate not only diurnal flow variations, but also sustained peak flows incurred by storm conditions. The first flush of strong wastewater as the sewers are scoured under storm conditions imposes considerable load on the receiving wastewater treatment plant, and the process must be designed to accommodate this.

The flow that must be treated at the treatment plant is fixed by the discharge consent. In the United Kingdom, this flow is calculated on the basis of domestic, visitor, and industrial discharges. The maximum flow to treatment is typically three times the dry weather flow. This means that, in wet climates, a final clarifier may sustain a high load for several days.

The infrastructure is also subject to fresh or saltwater inflow or infiltration. For example, during winter wastewater treatment works can receive high flows for several months at a time.

Many coastal sites suffer from saltwater inflow or infiltration during high tides. The choice of materials and the mixing effect of the influent structure are critical design considerations. It is not uncommon for some coastal works to experience peaks up to 5000 mg/L of chloride. This is also a concern in areas where salt water is used to flush toilets.

***Site Conditions.*** Within Europe and Asia, there is often very little land area to build new works or to extend existing works. This has led to adoption of alternative technologies to conventional circular clarifiers, with a small tank footprint being a necessity. Solutions include stacking process units and retrofit of, for example, Lamella plate separators.

***Effect of Collection Systems on Settling Tank Design.*** Hydraulic management of the collection system can either improve or degrade the performance of the works. For example, the use of large storage tanks to reduce combined sewer overflows may sustain the period at which the works is under peak hydraulic loading. On the other hand, regular flushing of inverted siphons and flat sewers can be used to prevent the grit lanes and primary tanks from being overloaded when the sewer is washed out after a period of low flows.

Sewerage models have been extended to include crude models that predict the movement of solids and soluble pollutants in sewers. However, these models do not predict the effect of hydrolysis and the growth of biofilms and biomass. Biological management of the sewer is a new area of research (Hvitved-Jacobsen, 2001). For example, an aerobic sewer may increase the fraction of solids that can be settled in the primary tanks, while an anaerobic sewer may reduce the settleable solids.

## PROCESS DESIGN

**TYPES OF SETTLING TANKS.** The four main types of settling tanks are

- Primary,
- Secondary (follows suspended-growth processes),
- Humus (follows fixed-film processes), and
- Storm (retains first flush and treats storm flows).

**FUNCTIONS OF THE SETTLING TANK.** The settling tank performs the following three functions: (1) clarification of the effluent, (2) thickening of the underflow, and (3) storage of sludge during dynamic flow events. Membrane bioreactors and intermittent plants (e.g., sequencing batch reactors [SBRs]) must perform the same functions. In a few situations, primary tanks may be used for flow balancing at small works or works serving commuter populations.

**SETTLING TANK CONFIGURATIONS.** There are three horizontal flow configurations.

- Circular (horizontal flow),
- Rectangular (longitudinal flow), and
- Rectangular (transverse flow).

Of the three, rectangular transverse flow is the less common. The preferred configuration for secondary and humus tanks is circular. Both longitudinal flow rectangular and circular tanks are used for primary treatment. In the United Kingdom, storm tanks are generally rectangular, while, in some parts of Europe, storm tanks are often offline primary tanks.

Vertical flow (e.g., Dortmund) tanks are used as primary, secondary, and humus tanks at small treatment plants (serving populations less than 2000 population equivalent).

Lamella settlers are used primarily as primary settlers. In some biological aerated filter works, Lamella tanks are used to treat the backwash. Lamella settlers have been used as secondary settlers with mixed success.

# PRIMARY TANK DESIGN

## TYPICAL UNITED KINGDOM DESIGN PARAMETERS

### *Retention Period*

$$\text{Retention period (h)} = \frac{\text{Total capacity of unit (m}^3) \times 24}{\text{Rate of flow of wastewater (m}^3/\text{d)}}$$

A nominal retention period of 2 hours at maximum flow is typically used.

### *Surface Loading*

$$\text{Surface loading (m}^3/\text{m}^2\cdot\text{d)} = \frac{\text{Maximum rate of flow (m}^3/\text{d)}}{\text{Effective surface area (m}^2)}$$

### *Upward Velocity*

$$\text{Upward velocity (m/h)} = \frac{\text{Maximum rate of flow (m}^3/\text{d)}}{\text{Effective surface area (m}^2) \times 24}$$

In Britain, surface loading rates of between 30 and 45 m/h are commonly used. These correspond to 1.2 to 1.9 m/h maximum upward velocity. The Institute of Water Pollution Control (IWPC) guidelines for primary sedimentation tanks are as follows (IWPC, 1980):

- Horizontal flow tanks: 30 to 45 $m^3/m^2\cdot d$.
- Radial flow tanks: maximum 45 $m^3/m^2\cdot d$.
- Upward flow tanks: upward velocity 1.2 to 1.8 m/h at maximum flow.

### *Weir Overflow Rate*

$$\text{Weir overflow rate (m}^3/\text{m}\cdot\text{d)} = \frac{\text{Maximum flow (m}^3/\text{d)}}{\text{Total length of weirs (m)}}$$

Values of 100 to 450 are generally used.

### *Horizontal Velocity*

A range of horizontal velocities between 14 and 54 m/h have been reported for rectangular tanks. However, different aspect ratios and maximum upflow velocity can limit the horizontal velocity (Nicoll, 1988).

**HORIZONTAL FLOW TANKS.** Figure 11.1 shows a typical horizontal flow settlement tank.

**RADIAL FLOW TANKS.** Figure 11.2 shows possible standardization of radial flow primary tanks.

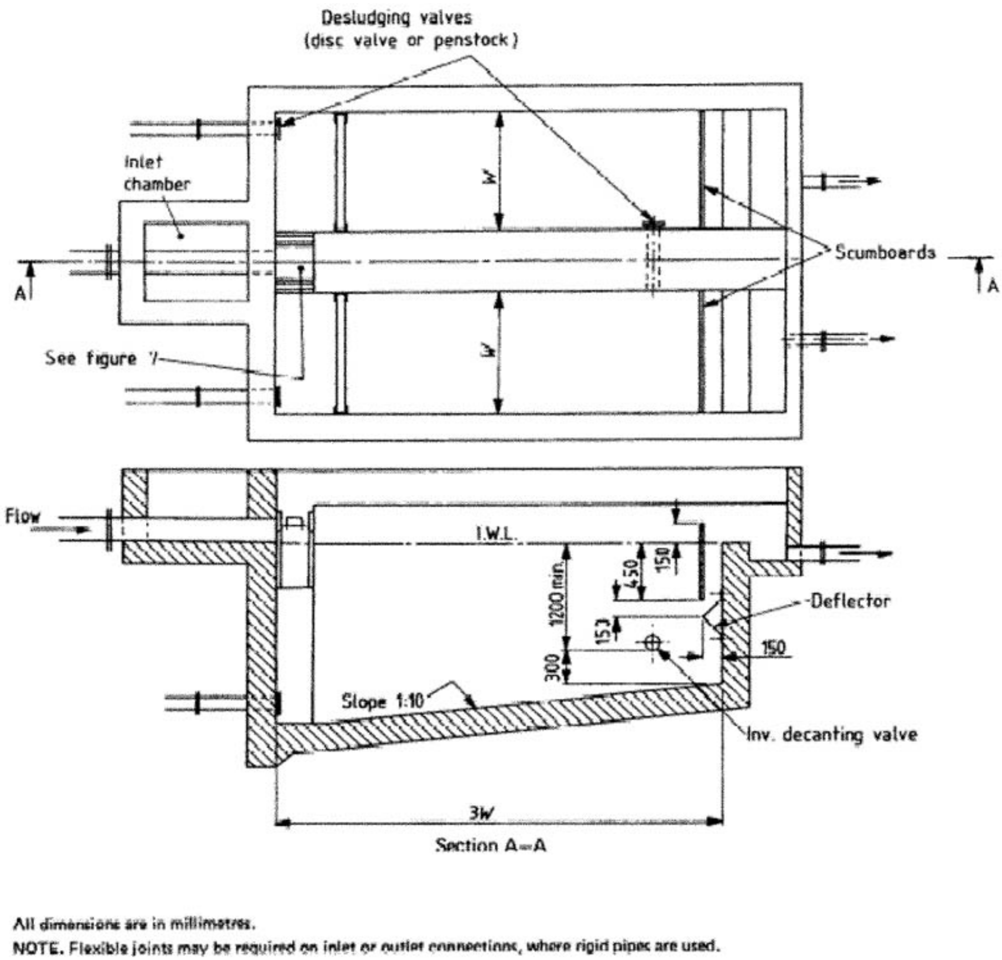


**FIGURE 11.1** Typical horizontal flow settlement tank (courtesy of British Standards Institution, London; all dimensions are in millimeters; flexible joints may be required on inlet or outlet connections, where rigid pipes are used).

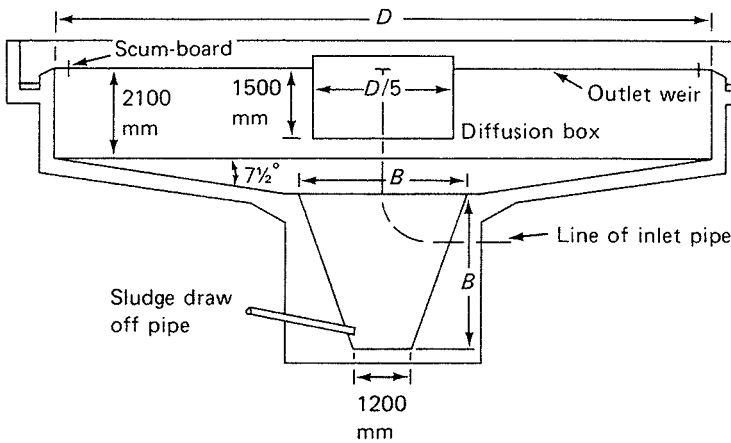


**Figure 11.2** Possible standardization of radial flow primary tanks (courtesy of the Chartered Institution of Water and Environmental Management, London).

**DESLUDGING.** Traditionally, horizontal flow rectangular tanks were often manually desludged by draining and squeegeeing the sludge. Primary tanks were used as sludge thickeners to minimize sludge volumes produced. Currently, the common practice is replacement with automatically desludged radial flow tanks. Desludging is generally by progressing cavity pumps on a timer. Typical desludging regime for a large works is six times a day for 15 minutes at a time. This frequent desludging regime leads to thinner primary sludges, which are frequently mechanically or gravity thickened before digestion.

Small works desludging regime is often constrained by minimum pipe diameters and pumping velocities and can challenge the design engineer to meet these constraints with clients' standard design requirements.

**COSETTLEMENT.** There are conflicting schools of thought on cosettlement of waste activated and trickling filter sludge in primary sedimentation tanks. Before automatic pumped desludging, primary tanks were operated as thickeners, and it was believed that the addition of secondary sludges provided a lubricating effect, making the primary tank desludging pipework less likely to block. With nitrifying plants, the general opinion is that denitrification of the secondary sludge in the primary tanks can lead to rising sludge and poor performance. Secondary sludges are generally separately thickened, mechanically or by gravity, and mixed with the primary sludge before digestion.

**ODOR CONTROL.** Many of the wastewater treatment works in the United Kingdom are close to residential areas, and odor control is a huge issue. Odor control is generally a planning permission requirement for new works, and odor abatement notices can be served on existing works. It is common practice to cover inlet works, preliminary treatment processes, primary tanks, and sludge handling areas and treat the air space for odor.

## HUMUS TANK DESIGN

Trickling filters are preferred over activated sludge systems in many parts of the world because they consume less energy, require less maintenance, and produce less sludge. However, the suspended solids quality achieved from these works is not as good as that from an activated sludge plant. For example, one small village works in Wales met a 60:40 consent with no electrical supply to the site. The humus tanks were desludged three times a week using a portable pump, primary treatment was provided by septic tanks, and the tricking filter distribution was hydraulically driven.

Sludge from a low-loaded trickling filter tends to follow a type I and II settling curve. Therefore, improved performance from an existing tank can only be achieved by inducing flocculation (e.g., polymer or chemical dosing) or by tertiary treatment. However, given the mineralized nature of the sludge, co-thickening humus sludge with primary sludge is common practice.

In situations where trickling filters cannot be installed, utilities have used submerged aerated filters (SAF), rotating biological contactors (RBC), or biological aerated filters. The settling tanks following both RBCs and SAFs are similar in design to those following trickling filters.

**TYPICAL UNITED KINGDOM DESIGN CRITERIA.** ***Surface Loading.***

$$\text{Surface loading } (m^3/m^2{\cdot}d) = \frac{\text{Maximum rate of flow } (m^3/d)}{\text{Effective surface area } (m^2)}$$

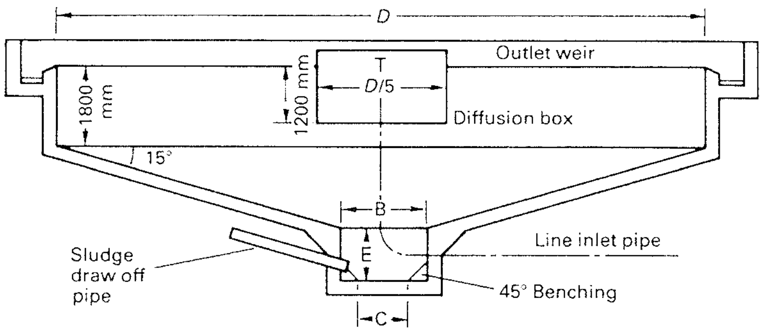


**FIGURE 11.3** Possible standardization of circular humus tanks (courtesy of the Chartered Institution of Water and Environmental Management, London).

Typical maximum surface loading rate is in the range of 1.2 to 1.5 m/h (29 to 36 $m^3/d$).

***Horizontal Flow Tanks.***
Capacity of the tank = 135 × population$^{0.85}$
Length to width ratio of at least 3:1.
Minimum depth 1 m at shallow end, 1.2 to 1.5 m preferable.

***Upward Flow Tanks.***

$$\text{Minimum surface area A} = \frac{3 \times \text{population}^{0.85}}{40}$$

Where A = Minimum area ($m^2$) at the top of the hopper (Nicoll, 1988) (see Figure 11.3).

# FINAL TANK DESIGN

**TYPICAL UNITED KINGDOM DESIGN PARAMETERS.** ***Retention Period.*** Traditional United Kingdom practice was to design on basis of retention time, 4 to 6 hours at one dry weather flow and surface loading of 1 to 3 m/h at peak flow.

***Mass Flux Theory.*** For the past twenty years, United Kingdom design practice has been based on flow, solids loading, and sludge settlement characteristics using mass flux theory and the stirred specific sludge index (SSVI) parameter (described below) for sludge settleability.

Typical parameters are as follows:

- SSVI: 100 to 130 mL/g.
- Maximum solids flux: 4 to 4.5 kg/m$^2$·h.
- Underflow velocity: 0.3 to 1 m/h.
- Maximum weir overflow rate: 10 m$^3$/m·h.
- Surface loading rate: 0.5 to 3 m/h.
- Return activated sludge (RAS) flow: 0.5 to 1.5 dry weather flow.

**SETTLEABILITY PARAMETERS.** The primary difference between the United States and European practice is the measure used to characterize the settleability of activated sludge. The SSVI and diluted sludge volume index (DSVI) tests have replaced the use of the sludge volume index (SVI) (Mohlman index) in design standards.

***Settlement Curve.*** Quiescent settlement of activated sludge is allowed for 30 minutes in a one-liter measuring cylinder, and the sludge–liquid interface is monitored at intervals. A plot of sludge height against time can be produced, and, from this, the initial settling velocity can be determined.

***Initial Settling Velocity.*** Measurement of the initial settling velocity (ISV) of activated sludge at various suspended solids concentrations can be used to model and optimize mixed liquor suspended solids (MLSS) concentrations and recycle sludge rates. A good settling sludge will have an ISV of 5 m/h or more, while a poor settling sludge could have a rate as low as 0.5 m/h.

***Sludge Volume Index.*** The SVI is defined as the volume (in milliliters) occupied by one gram of activated sludge solids after 30 minutes quiescent settlement in a one-liter measuring cylinder.

$$\text{SVI} = \frac{\text{Volume of settled activated sludge (as \% of total volume) after 30 minutes}}{\text{Concentration of MLSS (as \%)}}$$

An increasing SVI indicates decreasing settleability. An index of greater than 180 would indicate a bulking sludge. The SVI is dependent on sludge solids concentration and does not represent hydraulic conditions within the tank. In the United Kingdom, SVI is generally used on-site by operators as an indicator for controlling sludge inventory. It is not used as a design parameter.

***Stirred Specific Sludge Volume Index.*** The $SSVI_{3.5}$ is defined as the volume (mL) occupied by one gram of solids after 30 minutes of settling in a gently stirred (1 rpm) settling column at a standard initial concentration of 3500 mg/L.

Developed by the Water Research Centre (WRC) (United Kingdom)], the test uses standard stirred cylinders to reduce wall and bridging effects to better simulate conditions in a full-scale settlement tank. The mixed liquor is diluted using final effluent, and the test is typically conducted at three different MLSS concentrations. The results are interpolated and standardized to a MLSS concentration of 3500 mg/L.

The SSVI is the design parameter used in the WRC mass flux model, which is standard United Kingdom design practice. The typical final tank design uses a maximum SSVI of 120 mL/g. However, one water company designs clarifiers for a three-stage Bardenpho activated sludge plant at 100 mL/g.

The WRC procedure allows an error margin of ± 20% between the predicted solids loading and the actual applied solids loading.

The SSVI can have a theoretical range of 0 to 286 mL/g. The following ranges are found in practice:

- 40 to 60: Excellent settleability.
- 60 to 80: Good settleability.
- 80 to 100: Average settleability.
- 100 to 120: Poor settleability.
- Greater than 120: Bad settleability or bulking.

The WRC mass flux model is also commonly used as an operating tool to control sludge inventory and recycle rates. The model does not address settlement tank depth. United Kingdom final settlement tanks are typically shallow compared to United States design practice, with a side wall depth of 1.5 m being common.

***Sludge Density Index.*** This parameter is the inverse of SVI and is defined as the density of the settled portion of sludge after an unstirred period of 30 minutes.

$$\text{SDI} = \frac{\text{Concentration of mixed liquor suspended solids (as \%)} \times 100}{\text{Volume of settled activated sludge (as \% of total volume) after 30 minutes}}$$

The SDI for a good settling sludge may be 2.0 or greater, while a bulking sludge could have an index of 0.5 or less.

***Stirred Sludge Density Index.*** This index is measured in the same way as SDI, but using the stirred settling apparatus (CIWEM, 1997).

***Diluted Sludge Volume Index.*** This is defined as the volume (in milliliters) occupied by one gram of sludge after 30 minutes of settling in a one liter unstirred cylinder with the proviso that the sludge is diluted such that the settled volume after 30 minutes of settling is between 150 mL and 250 mL.

Both Dutch (STOWA) and German (ATV) design standards are based on DSVI. Table 11.1 is taken from ATV-DVWK-A 131 (May 2000), which recommends the DSVI to be used for different types of activated sludge plants (ATV, 2000).

***Differences between the Two Design Approaches.*** Both DSVI and SSVI developed from basic flux theory but have developed in different directions and give different results. The SSVI is the design parameter used in the WRC mass flux model, which is standard United Kingdom design practice (White, 1975). The WRC mass flux model is the Vesilind mass flux model incorporating an empirical relationship between SSVI and the Vesilind setting parameters. White's empirical function is valid only up to the critical underflow rate. Ekama and Marais established a relationship between the two (SSVI = 0.67 DSVI) (Ozinsky and Ekama, 1995). Their procedure is known as the modified WRC method.

The major difference between the modified WRC method and the ATV/STOWA procedures are the following:

- The ATV/STOWA procedures do not use the recycle rate to fix the permissible overflow rate.

**Table 11.1** Recommended DSVI to be used for different types of activated sludge plants.

| | **Industrial/commercial wastewater influence on DSVI** | |
|---|---|---|
| | No effect | Detrimental effect |
| Carbonaceous only | 100 to 150 | 120 to 180 |
| Nitrifying (and denitrifying) | 100 to 150 | 120 to 180 |
| Extended aeration | 75 to 120 | 100 to 150 |

- The ATV procedure includes a method to estimate the required depth of the settling tank.
- The ATV and STOWA procedures recognize compaction failure.

**DESIGN OF SLUDGE SCRAPERS.** While, in the United States, scrapers are driven from a spur gear at the center of the clarifier, scrapers in Europe are commonly driven by peripheral drives riding along concrete walls at the clarifier perimeter. This is less complex and expensive drive than the United States practice (Voigt, 2002).

Scrapers for the collection of sludges in European practice are typically "spiral" plow type scrapers. This is a more efficient method of collecting biosolids and is gaining popularity in the United States (Voigt, 2002).

The design of sludge scrapers is either proprietary or follows standard design practice. Both scraped and suction removal systems have been used successfully on circular and rectangular tanks. Pulsating scrapers (e.g., Zickert) have been used successfully on primary Lamella and conventional tanks.

One particular area of interest is the design of scrapers for circular secondary tanks. In the United Kingdom, the standard floor slope is 7.5 degrees, with a large central sludge hopper. The scraper is generally bought competitively against a standard specification. In other countries, where the floor slopes are flatter, the design of the sludge scraper is more critical. This can be seen from the Working Report on Sludge Removal Systems for Secondary Sedimentation Tanks of Aeration Plants (ATV, 2000).

**DESLUDGING SETTLING TANKS.** Sludge is removed from settling tanks either by hydrostatic bellmouths or direct pumping. Although direct pumping is the preferred method to desludge primary tanks, humus and final settling tanks are often desludged using bellmouths. Bellmouths feeding a RAS pump station is a more economical solution than direct pumping.

Bell mouths are generally set at commissioning to give the design RAS flowrate. As the flow to the tank increases, the increase in hydraulic head will increase the rate of sludge withdrawal. In many cases, on detection of a high blanket, the bellmouth will lower (increasing the flowrate) until the high blanket alarm is cleared.

Screw pumps or progressing cavity pumps are widely used to recycle activated sludge. This is because of concerns relating to "floc shear". The European practice is to avoid centrifugal or high-speed shearing devices on sludge from activated sludge basins (Voigt, 2002).

**SCUM REMOVAL.** Typically, scum is retained by peripheral scum boards, and there is a dipping scum box fitted to the peripheral wall of the tank. A strike bar submerges the scum box with each rotation of the scraper bridge and flushes the scum through with effluent to an alternative process stream. Prevailing wind direction must be taken into consideration when locating the scum boxes.

There is much debate as to the best return point for the scum stream. Ideally, it is completely separated and fed directly to the sludge treatment stream to prevent reinfection. There are, however, large volumes of liquid associated with scum removal systems.

Bridge mounted systems are frequently retrofitted to tanks with ineffective or non-existent scum removal. The scum collection box is fixed to the rotating bridge just below surface level. A pump returns the collected scum and water to the clarifier central stilling box with the intentions that the scum will cohere with the biomass and settle. The advantage of this system is that there is no requirement for pipework through the peripheral tank walls. The systems are typically operated by timers.

**DEEP SIDEWALL DEPTHS.** Sidmouth wastewater treatment works (WWTW), (Devon, United Kingdom) has a residential population of 15 000. However, it has an annual folk festival for eight days in July, and the population equivalent increases to 36 000 during this period. A new works was commissioned in 2000 to meet the Bathing Water Directive. The bioreactor and final clarifiers had to accommodate the upturn in population. The clarifiers were built with a 4-m sidewall depth, and the works performed within consent during the first folk festival.

**PARABOLIC FLOORS.** Mogden WWTW (London, United Kingdom) has four constant velocity radial flow tanks. Across the 43-m diameter, the depth is varied to provide a constant cross-section at all diameters. The depth varies from 0.9 m at the periphery to a central sludge storage hopper of 9 m deep. The single scraper blade is curved in the vertical plane to fit the shape of the floor.

Two similar tanks were built at Skelmersdale (Lancashire, United Kingdom) (Stanbridge, 1976).

## ALTERNATIVES

**TERTIARY TREATMENT.** Alternatives to providing additional final clarifiers to meet stringent consents include irrigation over grass plots or reed beds, settlement in lagoons, gravel bed filters, slow sand filtration, rapid downward and upward flow sand filtration, wedgewire filters, brush filters, microstrainers, and membranes.

In many cases, the effluent produced by the tertiary treatment is of a higher standard than required, and it is common practice to treat just part of the flow and blend tertiary treated effluent with secondary effluent to minimize capital cost. Tertiary treatment is often retrofitted and requires pumping. Minimizing throughput also minimizes operating and pumping costs.

**SEPTIC TANKS.** Septic tanks are generally found in small communities. Often there is no main drainage and each residence has its own dedicated septic tank. In this case, maintenance and emptying responsibilities lie with the householder. The discharge is generally to a soakaway and is not consented. In the United Kingdom, the potable water provider and the wastewater treatment agent tend to be the same company. The householder with a septic tank pays reduced rates to the potable water provider.

Where there is local drainage to a communal septic tank, the water company is responsible for its operation and maintenance, and the effluent is consented. The consents are generally quite relaxed and often descriptive. Most problems arise where septic tanks were installed for the local community but now receive peak loads because of visitors, for example, country inns receiving high nonresident dining trade.

The typical design criterion for septic tanks is capacity and expressed as follows:

$$C = 180P + 2000$$

Where

C = Capacity of the tank in liters, and
P = Design population with a minimum value of 4 (Nicoll, 1988).

**ALTERNATIVES TO CONVENTIONAL CLARIFIERS.** The first activated sludge works were operated in fill-and-draw mode. By the 1920s, many of these systems were converted to continuous flow. The three main objections to fill-and-draw systems at the time were the following:

- Instantaneous effluent discharges,
- Diffusers blocking after settlement phase, and
- Operator attention required to open and close valves.

In the 1970s, simple oxidation ditch-type batch processes were used, including triple ditches, split channel oxidation ditches, Biwater's BIFAD (continuous flow fill, aerate, settle by timer), and Kruger's biodenitro/biodenipho processes. By the early 1980s, small SBR works were built that used either coarse bubble or jet aeration.

By the end of the 1980s, most of the problems that had shifted technology away from batch systems were overcome either by computer technology or improvements in aeration devices (Halladey and Coleman, 2001).

The processes currently attracting interest in Europe are Lamella separators as clarifier enhancement, and SBRs, triple ditches, and dissolved air flotation (DAF) as alternatives. Conventional Lamellas are addressed in Chapter 3. Spiral Lamellas are a British innovation and are described below.

***Spiral Lamella Separators.*** The DeHoxar spiral separator was conceived and developed in the 1990s by David DeHoxar, Southern Water, United Kingdom. It is a novel, very compact, gravity settlement device. Spiral separators are in use on both water and wastewater treatment works.

Spiral separators are more compact than conventional Lamella separators and can require as little as 3% of the footprint requirement of conventional settlement tanks. Where odor control is an issue, the small footprint minimizes capital cost for covers.

Figure 11.4 is a diagram of a plate pack with six interleaved plates. Each plate is in the shape of a conical helix and is a full 360-degree helical turn. The plates are attached to a cylindrical core. The plate shape gives inherent strength and stiffness. No spacers or structural connections between the plates, other than the core, are required to keep the whole plate pack in shape.

Figure 11.5 is a section through a spiral separator designed for primary treatment of wastewater.

*THEORY OF THE SPIRAL SEPARATOR.* The theory of the spiral separator is shown diagrammatically in Figure 11.6.

Figure 11.6a shows a conventional rectangular settling tank. All particles that will settle through the whole depth of the tank ("H") in the retention time within the tank are removed.

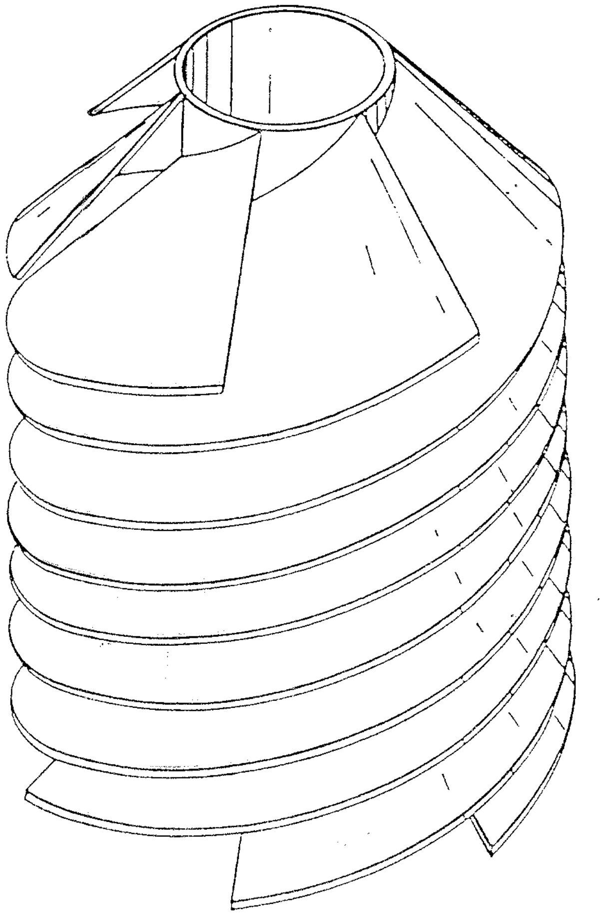

**FIGURE 11.4** Spiral plate pack.

Figure 11.6b shows a Lamella plate separator. All particles that will settle through the vertical distance between the plates ("h") in the retention time within the plate pack are removed.

The Lamella separator is a more compact device as, for the primary treatment of wastewater, "H" is typically 2 to 3 m, and "h" is typically 150 to 200 mm.

An alternative method for comparing performance is to consider the effective settlement area. For a conventional settlement tank, this is equivalent to the area of the tank, and, for a Lamella separator, it is the sum of the projected horizontal area of all the plates in the plate pack.

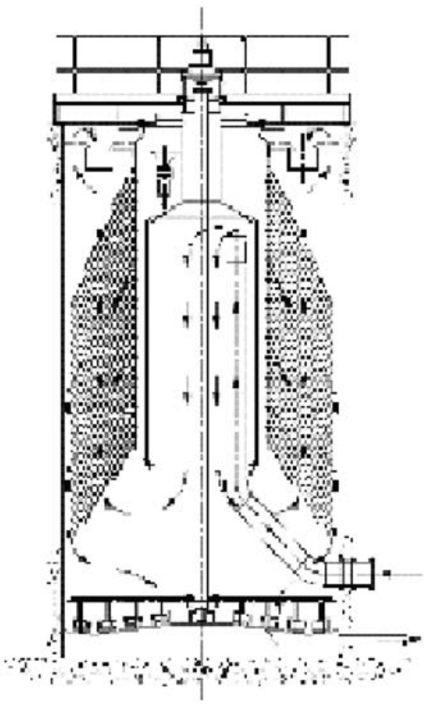

**Figure 11.5** Section through a spiral separator.

In a spiral separator, rotating the plate pack moves the plates relative to a settling particle. The effect of moving the plates is shown in Figure 11.6c, as a velocity triangle. If a particle is settling at a velocity "v", and the plate is moving at velocity "M", the relative velocity of the particle to the plate is shown in the velocity triangle by the arrow "r". The length of the arrow "r" is greater than the length of the arrow "v", showing that the relative velocity is higher than the absolute velocity.

Plate pack rotation is typically at or around "no swirl speed". This is typically approximately six revolutions per hour for a spiral separator operating at its hydraulic capacity.

*PROCESS DESIGN PARAMETERS.* The principal parameter for sizing spiral separators is hydraulic loading rate of plates. This is calculated by dividing the maximum flow out of the spiral separator by the horizontal projected plate pack area. Solids flux on the plate pack and the annulus between the plate pack and tank wall is checked to make sure this is within a reasonable operating range.

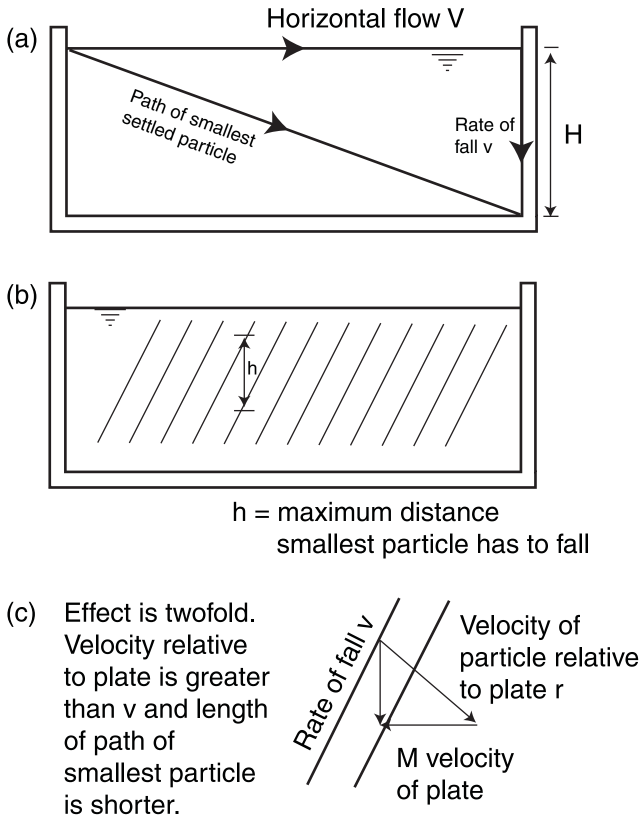


**FIGURE 11.6** Theory of the spiral separator.

Hydraulic plate loading rate =

$$\text{Max flow} / (\text{od} - \text{id})^2 \times \pi/4 \times n \times \theta/360 \text{ (typically expressed in } m^3/m^2{\cdot}h)$$

Where

od = Outside diameter of plate pack,
id = Internal diameter of plate pack,
n = Number of plates in the pack, and
$\theta$ = Degree of turn of each plate, typically in the range 180 to 270 degrees.

*DESIGN CONSTRAINTS.* Spiral separator size is the internal diameter of the tank in which the plate pack is installed.

For spiral separators 5 m and over, plate packs are supplied in "flat pack" style for assembly on-site. These are generally installed in in-situ cast reinforced concrete tanks. Platelets are moulded in a plastics factory, so that only a limited range of sizes are made. These are 5, 7, and 10 m (and, in the future, 14 m). Each size has double the capacity of the size below. Therefore, one 7-m platelet has the capacity of two 5-m platelets, and one 10-m platelet has the capacity of two 7-m platelet, and so on.

The number of "middle" platelets in each platelet can be varied to match plate area to the duty for each project. There is a specially shaped platelet at the top and bottom of each plate for structural reasons. The area of shaped top and bottom platelets is ignored in the calculation of hydraulic plate loading.

Spirals up to 4 m are fully assembled in a workshop and supplied with the steel process tank. These can be made in any size, but, for standardization, the number of different sizes will be restricted. Currently, 3 m is a "normal" size.

Plate manufacturing techniques are currently being reviewed. It is proposed to move from resin transfer moulded glass reinforced plastic to vacuum formed polypropylene plates.

*UPSTREAM REQUIREMENTS.* A 6-mm, two-dimensional screen is required upstream of spiral separators as a minimum. Effective grit removal is also required. For primary treatment of wastewater, the inlet dissolved oxygen should be 4 mg/L to avoid anaerobic growth on the plate packs. This requirement can be relaxed if secondary sludges are cosettled.

*DESLUDGE REQUIREMENTS.* Continuous sludge removal is best, but little, and, often, sludge removal is acceptable. There must be a positive indication that sludge is being removed from each spiral on site.

A dedicated progressing cavity pump for each spiral separator with a common plumbed in standby is a recommended arrangement.

*MAINTENANCE AND INSPECTION.* Each spiral must be drained down and inspected every three months. With sufficient operating experience on a particular site, the frequency of drain down and inspection can be reviewed.

***Dissolved Air Flotation.*** Flotation techniques have been popular in many industries, including mining, food, oil, and water/wastewater. Induced and dissolved air, froth, and electroflotation have been used over the years. In the water and wastewater industry, DAF has been the most successful technique, with hundreds of plants now in operation. Manufacturers have developed their own proprietary approaches to air dissolution and release method, sludge removal technique, basin geometry, and pretreatment approach. However, the underlying advantages remain as follows:

- High rate process—small footprint,
- Low residence time,
- Thick sludges can be produced, and
- Rapid startup and shutdown.

While the process is now very common for water treatment, wastewater examples are less common. The use of DAF for wastewater was popular in the 1960s and 1970s at some locations for thickening surplus activated sludge. More recently, there has been some resurgent of interest in the use of DAF thickening, particularly in Germany, either to overcome the solids-limiting factors of conventional final clarifiers and therefore operating at significantly higher MLSS concentrations, within the aeration basin, or in the reduction of the return activated sludge/surplus activated sludge volumes. The introduction of some novel proprietary concepts have seen that recycle volumes and, hence, energy input can be reduced. However, the use of belt, drum, and centrifuge thickeners continues to predominate thickening applications. Most of the DAF uses to date have been in Scandinavia for tertiary treatment, particularly Sweden, and, on occasions, primary treatment with chemicals in Norway. The Swedish applications have generally been associated with chemical phosphorous removal, typically to achieve a 0.5 mg/L phosphorus limit. In this application, the process also serves to reduce solids typically to less than 10 mg/L and to act as a final guard in the event of an upstream plant problem. Surface loading rates were typically 5 to 10 $m^3/m^2 \cdot h$, with chemical flocculation times of 5 to 10 minutes, with sludge concentrations of 3 to 4 % being produced. Plant capacities of up to 650 000 population equivalent have been built. The DAF process continues to show itself as flexible, robust, reliable, and easy to operate.

More recently, the development of compact moving bed biological reactors and other biofilm processes has brought about a resurgence of interest in DAF as a compact final solids separation stage. Recent wastewater plants at Gardemoen, Sweden, and Lowestoft, United Kingdom, have adopted DAF as a final solids separation stage (see Figure 11.7). Both of these sites are totally enclosed for aesthetic and environmental reasons, and, again, it is the compact nature of the DAF process that makes it attractive in whole life cost terms.

**FIGURE 11.7** Saturator.

**SINGLE TANK REACT/SETTLE.** Several SBRs, triple ditches, and split ditches have been constructed in the United Kingdom. The SBRs and triple ditches have produced excellent results, but the split ditch has proved unreliable at high flows. The split ditch differs from the triple ditch in that it does not fully separate the settlement modes by using isolated tanks for each zone. This causes currents in the split ditch, which reduce the efficiency of settlement.

***Sequencing Batch Reactors.*** In Australia and North America, small extended aeration SBR plants were built using either coarse bubble aeration or jets. Some deficiencies specific to SBRs were the following:

- Unequal distribution of the sludge inventory among the basins; and
- Operators' unfamiliarity with the operation of batch processes and discontinuous discharges.

The United Kingdom overtook North America in the application of SBR technology because United Kingdom works treat combined storm and foul wastewater to satisfy consents based on spot rather than composite samples. This is a more stringent design requirement. United Kingdom designers refined the design to operate in storm cycle 20 to 25% of the year rather than a few times a year, as is the case in many United States works.

The transition from the American to the United Kingdom market was not without problems, the most recent being the relationship between the catchment characteristics, decant depth, and cycle time on the works treatment capacity.

Today, robust and advanced systems are available from numerous suppliers, including CASS (EarthTech) (South Yorkshire, United Kingdom), ITT Sanitaire (United Kingdom), Fluidyne (Ontario, California), Degremont (Walla Walla, Washington), AquaAerobics (Rockford, Illinois) Purac (Netherlands), Waterlink (United Kingdom), and others.

United Kingdom process consultants and contractors have developed methods to check and refine designs supplied by others to ensure that the design suits the catchment being served. This is possible using activated sludge kinetics and computer simulation.

***Advantages and Disadvantages of Sequencing Batch Reactors.*** Activated sludge is a robust method to remove BOD, oxidize ammonia, and reduce nitrate. All the variants of activated sludge, including SBRs, share these features. The relative advantages and disadvantages of these technologies are primarily related to their mode of operation and are depicted in Table 11.2.

***Triple Ditches.*** Recent interest in SBRs and the batch settling characteristics of activated sludge generates interest in a development of the 1980s—the triple ditch. This form of treatment was adopted on several sites in the United Kingdom in the 1980s and has performed successfully to date, after some initial teething problems with suspended solids in the effluent. Recent experiences with SBRs could prompt a re-examination of the validity of the triple ditch as a compact solution.

*OPERATIONAL BASICS.* It can be that the batch nature of the SBR will provide excellent effluent qualities, but operators may prefer the continuous flow character of a conventional plant.

Attempts to combine the best of both types of design have resulted in such designs as the split and triple ditch. In these designs, no separate clarification stage is provided. Instead, the basic tank is split into three separate zones. The central zone is continuously aerated, but the outer zones are controlled to alternate in aerate, batch

**Table 11.2** Relative advantages and disadvantages of SBRs as they are primarily related to their mode of operation.

| | **Sequencing batch reactor** | **Continuous flow reactor** |
|---|---|---|
| Flexibility | The decant depth and cycle time can be adjusted to meet seasonal flows or meet seasonal consents. The capacity of the system can be increased by increasing the decant depth or constructing more reactors. | To upgrade a continuous flow works, the diffuser pattern must be altered and the pipe work changed. The capacity is increased by constructing new process units. |
| Footprint | The shape of the reactors can be changed to fit the site much more effectively than bioreactors and final settling tanks. The footprint can be reduced by stacking the SBR cells. | The footprint of a continuous flow works can be reduced by using deep aeration tanks and Gould type II rectangular final settling tanks. |
| Capital costs | Simpler civil structure may reduce capital costs, particularly if ground conditions are poor. | Circular final settling tanks and associated piping and pumping are expensive. |
| Diffusers | Requires twice the diffusers of a continuous flow reactor. Settling in the SBR precludes the use of ceramic diffusers. | Requires less diffusers than an SBR, but diffuser density must be tapered. |
| Phosphorus removal | Chemical and some biological. | Can achieve a stringent phosphorus standard (1 mg/L) biologically if a volatile fatty acid source is available. |
| Nitrification capacity | Lower than plug flow plant. Not recommended for stringent spot ammonia consents (3 mg/L). | Achieves almost complete nitrification. The final tanks buffer ammonia peaks. |
| Control | More complex than a continuous-flow reactor because aeration, wasting, and discharge must be scheduled. Also, the controller must switch system from normal to storm cycles. Interlocks must be established to prevent failures, e.g., decanting during aeration. | Aeration and wasting are the main control loops. |
| Sampling | The solids in the SBR must be measured when the reactor is aerated and the volume is known. The effluent must be sampled during decant. | For the most part, sampling is independent of the operation of the system. |

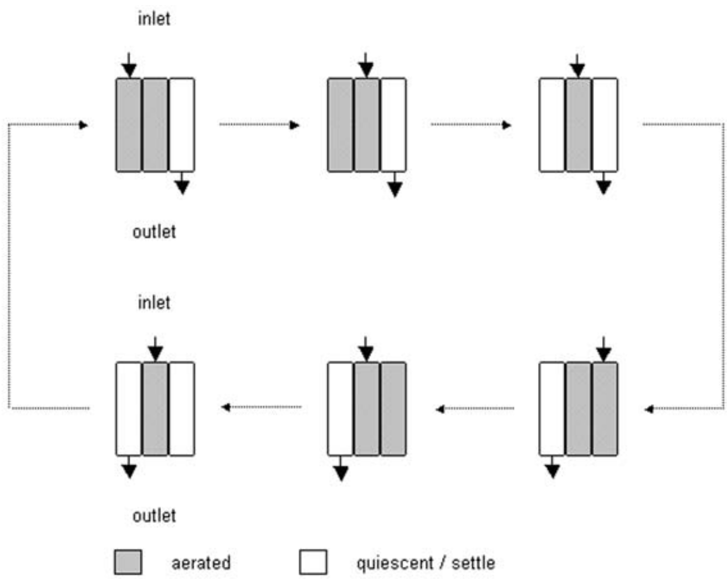


**FIGURE 11.8** Triple ditch schematic.

settle, and dynamic settle modes, very closely mimicking the operation of the SBR without actually stopping the process. Flow is fed to each tank in turn, depending on the point in the cycle. (Figure 11.8).

*TRIPLE DITCH DESIGN PARAMETERS.* The triple ditch was derived from conventional extended aeration configured as a "race track", where the mixed liquor was continuously circulated around the aeration tank using mechanical aeration via rotors to provide both oxygen for the process and propulsion for the mixed liquor. In the examples constructed in Anglian Water, United Kingdom, all use horizontal rotors. Further refinements to this design have resulted in the aeration and propulsion functions being separated: using fine-bubble diffused air grids in conjunction with large diameter rotating blades.

The original design parameter for an extended aeration plant was based on a long sludge age to confer stability of sludge plus high quality effluent. This resulted in a long retention time, as follows:

$$R = \frac{B \times 1}{C \times F{:}M}$$

Where

$R$ = Retention time at dry weather flow, d;
$B$ = BOD concentration applied to tank, mg/L;
$C$ = MLSS, mg/L; and
F:M = BOD load per unit sludge mass per day.

Substituting typical United Kingdom values and the classical F:M chosen for extended aeration of 0.055/d:

Retention at dry weather = 37 hours.

Incorporating settlement to the design without using external clarifiers, one-third of this volume will be used as a clarifier. The total volume of the triple ditches built within Anglian Water were based on one of the following:

Simple calculation based on F:M as above (actual F:M = 0.07/d).

Calculation as normal, but volume increased by 50%.

F:M reduced by ratio of aerated/mix contact time ($t_a$) to total contact or residence time ($t_c$)

$$\text{Minimum biomass, kg/d} = \frac{B \times f \times t_c}{F{:}M \times t_a}$$

It can be seen that whatever method is used, the biomass present will be adequate for nitrification. Once the basic volume is determined, one-third is taken for each tank and the outer tanks checked for performance as a clarifier. This is determined by considering the following:

- Maximum depth of tank to suit mechanical aerator, typically 3 m.
- Length-to-breadth ratio fixed by maximum rotor width, typically 6 m.
- Batch settlement rate.
- Dynamic flow.
- Movement of solids.
- Decanting weir length.

Flow across the tanks is by large ports at low level in the internal tank walls positioned near the inlet end to reduce potential disturbance of the sludge blanket during decanting. When the aerators are switched off in one of the outer tanks, a period of up to 10 minutes is required to obtain quiescent conditions. Flow will be completely isolated in this phase for up to 30 minutes, allowing any denitrification to occur and allowing for batch settlement. The aim is to lower the sludge blanket so that it is approximately 1 m below the outlet weirs.

*PREDICTING SLUDGE BLANKET DEPTH.* The usual formulae used to predict the depth are zone settling velocity equations.

$$U_{max}\text{, m/h} = V\, e^{-kC}$$

Where

$e = 2.72$,
$V = 9.32 - 0.039$ SSVI,
$k = 0.269 + 0.00122$ SSVI, and
$C$ = MLSS, kg/m$^3$.

The depth of sludge blanket $h = V_o\, e^{-kC} \times t_s$ where $t_s$ is the settling time in hours.

The WRC settlement equation has been used above, but there are many equations that may be used. Computer modeling systems, such as Hydromantis GPS-X (Hydromantic, Inc., Hamilton, Ontario, Canada), use the Vesilind equation. This has the same form as the WRC, but the constants are determined from the SVI, and not SSVI (which is "normalized" at MLSS = 3500 mg/L). The constants are empirically the following:

$$V_{max}\text{, m/h} = \frac{1\,[\,703.7 - 4.67\,\text{SVI} + \text{SVI}^2\,]}{24}$$

$$k = 2.6556 \times 10^{-4} - 2.847 \times 10^{-6}\,\text{SVI} + 2.5011 \times 10^{-8}\,\text{SVI}^2$$

This formula generally gives greater settlement velocities.

Practical observation of the triple ditch constructed at Upminster, Essex, United Kingdom, demonstrated that, if the WRC zone settling velocity was used on the initial MLSS concentration, then the actual depth of sludge blanket was overestimated. However, the settlement period can be divided into several equal periods and the calculation made for each, assuming that the total sludge mass occupies the resulting sludge blanket, hence increasing the concentration and decreasing the zone settling velocity. The individual depths are then summated to give the predicted total depth to the sludge blanket. This method gave a better agreement with reality and may easily be applied to a computer spreadsheet.

Stirred specific sludge indices of approximately 87 mL/g were found. The sludge blanket was measured to fall approximately 0.75 m in 55 minutes, continuing to fall during decanting to approximately 1.2 m. The WRC equation would predict a fall of 1.45 m; however, a 10-minute period of turbulence preventing settlement occurred because of the switching off of the horizontal mechanical aerators. Correcting the prediction for this gives 1.2 m depth $>$ 0.75 m actual. This is a single set of measurements, and more results need to be published to gain confidence that design predictions can be realized; however, it seems to indicate that a more cautious approach is advisable, at present.

It can be seen from the above that most of the settlement that occurs is during the 30-minute quiescent period. A dynamic flow period then follows by lowering the decanting weirs and allowing flow through the tanks. In this period, which can be up to 4 hours, further settlement occurs at an unpredictable and slower rate. The main objective in this period is to avoid washing the sludge blanket over the weirs.

*DESIGN OF DECANTING WEIRS.* The length of weirs required in rectangular settlement tanks is generally determined by specifying a weir loading rate. Experiments in Southend, United Kingdom, showed that good results were obtained using a loading rate of 5 $m^3/m{\cdot}h$, while acceptable results for meeting a 35-mg/L suspended solids consent standard could be obtained at 8 $m^3/m{\cdot}h$. It can be seen that, at either of these rates, very long weirs will result. The weirs on the triple ditches constructed were located at either side of the tank and decant initiated by rotating the tilting weir a short distance. To follow these conservative rules will result in an inoperable system.

In practice, the weirs provided are 26 m in total length, and the typical maximum head was designed to be 30 mm at 3 dry weather flow (DWF) and 44 mm at 6 DWF. The works sees no more than 3 DWF typically, but exceptional storms can pass 6 DWF through the plant.

*PERFORMANCE OF WEIRS.* Rotating the tilting weirs to initiate decant is a critical process. Observations on-site showed that mixed liquor was splashed over the weirs into the outlet channel during aeration and thus washed out as final effluent upon decant, resulting in very high suspended solids being experienced for up to 10 minutes. With a head of between 25 and 30 mm, the suspended solids when decanting had stabilized at less than 35 mg/L and, as flows varied, could be very low.

During aeration, scum was trapped between the cylindrical floating scum barrier, and deformation of the tubes allowed scum to contaminate the final effluent. A steel scum plate was subsequently fitted to the weir mechanism so that it rose and fell as the weir rotated, thus allowing the plate to prevent scum from splashing into the outlet channel or coating the weir during aeration, but retracting it from the water during decanting.

These problems caused during aeration were more detrimental to performance than any theoretical local increases in velocity resulting from short weirs and were basically the consequence of adopting horizontal mechanical aeration brushes. This problem was overcome by diverting the first flows from the decanting operation back to the head of the works for a short interval of time, then directing flow to the outfall.

*MIGRATION OF MIXED LIQUOR SUSPENDED SOLIDS.* Another effect that determined the actual settlement of the MLSS was distinct movement of solids from the inner tank to either outer tank while decanting. This increased the suspended solids concentration in the outer tank during quiescent settlement, reducing the rate of descent of sludge blanket and disturbing the already settled blanket. The design MLSS for the system was 3500 mg/L, but the outer tanks operate at more than 500 mg/L greater MLSS.

*SUMMARY OF PERFORMANCE.* In the 1980s, several split and alternating ditch systems were built in Anglian Water for approximately 30 000 population equivalent each. The split ditch had settlement problems and a clarifier was added. The triple ditches at Wickford, Basildon (two constructed), and Upminster remain operational in their original form. All experienced operational difficulties associated with solids washing out into the effluent as the cycle changed from batch settlement to decant. The main problem was scum and MLSS trapped in the weir and channels of the decanting system, combined with high local velocities at the weir. The operators have learned to live with them, and, overall, the solutions are as good as any other oxidation ditch. The other triple ditch in Anglian Water at Heacham experienced solids washout during storm flows and again has been modified to overcome this. It is worth noting that the triple ditches were found to be cost-effective when compared to standard oxidation ditches, but not necessarily when compared to other activated sludge processes that met the same quality requirements.

The numerical performance of the plants installed is summarized below.

- Upminster + grass storm plot suspended solids = 5 mg/L average long-term
- Basildon suspended solids = 7 mg/L average long-term
- Heacham before modification suspended solids = 5 mg/L average

# OPERATIONAL PROBLEMS

**MICROBIOLOGICAL CONSIDERATIONS.** The microbiology in the aeration tanks and biological filters is dependent on the conditions within the unit process and the characteristics of the feed to the works. These include industrial contributions, nutrient levels, soluble organic content, temperature, pH, septicity, and type of sewer (separate or combined). When settling deteriorates, the normal response is to examine the sludge under microscope (Eikleboom, 2000).

If filamentous bacteria are the cause, then they are identified as per Eikleboom or Jenkins. This is important because chlorination, as a method of controlling filaments, is not an acceptable option at many sites in Europe (Jenkins et al., 1993).

**DIAGNOSIS.** With traditional staffing levels drastically reduced, few sites are permanently staffing and are visited on a rotating basis. Instrumentation and telemetry are largely used to monitor works performance. This leads to rather "blind" diagnosis, and problems such as uneven distribution between tanks may not be identified.

**COMMON UNITED KINGDOM PROBLEMS.** ***Primary Tanks.*** Traditionally, the main problems with primary tanks have been rag and grease blockage and sludge compaction blocking sludge draw off pipes. These problems have largely been resolved by the installation of inlet works fine screens, scum traps, and autodesludging. However, many of the screen installations are underdesigned or do not have sufficient washwater supply, and it is common to see the screens being bypassed.

***Humus Tanks.*** Seasonal growth of duckweed on humus tanks imposes a high BOD on the effluent and can cause problems with downstream tertiary treatment processes. Various solutions have been installed, from mechanical "gulpers", covering of tanks to exclude sunlight, to encouragement of wildfowl to feed on the duckweed.

***Final Clarifiers.*** United Kingdom final clarifiers are traditionally shallow. It is common to see the sludge blanket rise close to surface of the tank under normal daily flow conditions. When works were highly staffed, operators would frequently increase the RAS return rate to lower the blanket. This often exceeded the limited flux and exacerbated the problem. The operators no longer have the luxury of time to watch the sludge blanket rise and fall and blanket detectors' monitor to control the system. Retrofit of selector zones before the activated sludge basin has generally improved activated sludge setting characteristics.

Floating biomass and scum can be a problem. Mechanical "gulpers" have been installed, with spray bars on the scraper bridges to drown the solids. Buildup can be directional on large diameter tanks because of prevailing wind direction. In some cases, the scum has formed a solid surface on which vegetation has become established.

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# Chapter 12

# Interaction of Clarifiers with Other Facilities


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# INTRODUCTION

Primary and secondary clarifiers are an inseparable and integral part of every conventional wastewater treatment plant. Their performance efficiency is affected by the upstream wastewater collection and treatment facilities and has a significant effect on downstream biological treatment and solids handling facilities. This chapter addresses the interaction of plant clarifiers and the other wastewater treatment and solids handling processes.

# CLARIFIERS AND WASTEWATER COLLECTION SYSTEMS

**EFFECT OF WASTEWATER COLLECTION SYSTEMS ON CLARIFIER DESIGN.** Wastewater collection system type has a pronounced effect on the wastewater treatment plant influent. Combined sewer systems are subject to wider flow variations compared to separate sanitary sewers. With a combined sewer system, wet weather plant influent flow could reach several times the average plant dry weather flowrate. The enforcement of more stringent regulations limiting combined sewer overflows (CSOs) and increased requirements for stormwater treatment would ultimately result in elevated plant influent flows and a higher potential for a negative effect on clarifier performance. Recent developments of CSO regulations induced the wider use of wastewater storage during wet weather events and real-time control of the CSOs to maximize the use of the sewer storage capacity and minimize the overflows. These CSO control measures typically result in a plant clarifiers being subjected to peak wet weather conditions more frequently and for longer periods of time (Ekama et. al., 1997).

Under wet weather conditions, the increased and diluted influent plant flow typically stirs up the clarifier sludge blanket and intensifies the transient currents in the clarifiers. This affects both clarifier effluent quality and the density and quality of the clarifier sludge. Transient flows have a negative effect on both primary and secondary clarifiers and on the overall secondary treatment process. Cooler stormwater deteriorates activated sludge settling characteristics and the overall hydraulic performance of the clarifiers. Prolonged wet weather events may also result in significant washout of grit from the sewer and grit chambers to the primary clarifiers and in reduction of the sludge volume index in the secondary clarifiers to very low levels.

Another industry-wide trend that has a measurable effect on plant capacity and clarifier performance is the implementation of comprehensive wastewater collection system infiltration and inflow reduction programs. Infiltration and inflow could contribute significantly to plant influent quality and quantity, especially in areas with highly permeable soils, high groundwater tables, and old wastewater collection systems. As an effective infiltration and inflow reduction program is implemented, the plant influent flowrate would typically decrease between 5 and 25%, which, in general, would have a positive effect on clarifier performance. However, plant influent wastewater strength is also likely to increase significantly, which would result in increased sludge production and sludge blanket depth in the primary and secondary clarifiers.

Wastewater treatment plant hydraulic design flows used to reflect the effect of the type of sanitary sewer on clarifier design are daily average flow; maximum daily flow; peak hourly flow, and peak instantaneous flow. Each of these design flows is important for different reasons. The peak instantaneous flow is used for the design of the plant influent pumping capacity and for determining clarifier design provisions needed for handling sewer system flow surges during wet weather conditions. The peak instantaneous flow is also considered when selecting sludge blanket depth control strategy in secondary clarifiers during transient flows. Average and peak daily flows are used to determine clarifier average and maximum daily hydraulic and solids loading rates and to select the type, size, and configuration of the clarifier sludge collection and withdrawal systems. Peak hourly flow is used to estimate the maximum depth of the clarifier sludge blanket. Peak daily and hourly flows are also used to size plant equalization basin and/or other on- or off-site wastewater storage equipment.

Traditionally, peak daily flow is estimated by applying a peaking factor to the average daily flow. However, when available, actual flows provide a more accurate representation of the plant peaking factors and should be used for determining peak design flows. Computer models based on actual wastewater collection system data and existing flow patterns are recommended to be used for large complex sewer systems to establish key design plant flows. These models typically incorporate key sewer system characteristics, such as tributary area served, rainfall duration and intensity, and time of concentrations, location and volume of sewer system retention basins (if any), and existing CSOs, which allow one to accurately determine plant peak instantaneous flow and its effect on clarifier design.

**MITIGATION OF TRANSIENT FLOW EFFECT ON CLARIFIER PERFORMANCE.** ***Transient Flow Reduction Measures in the Wastewater Collection System.*** The effect of transient loads on the plant clarifier performance can be decreased by a number of sewer system peak-flow-reducing measures, such as implementation of comprehensive infiltration and inflow flow reduction programs; more frequent sewer system cleanings and repairs, aimed to restore collection storage system capacity and to remove flow obstructions that decrease sewer retention volume; minimizing industrial wastewater discharger peak flows by enforcing construction of discharge flow and load equalization measures; enlarging key bottlenecked sections of the wastewater collection system; and providing sewer system retention tanks.

***Reduction of Transient Flow Effect by Equalization.*** Plant influent equalization is an effective transient flow reduction measure. The use of equalization basins in facilities with wide variations of diurnal plant influent flow (peaking factor higher than 2.5) would allow one to significantly decrease the size of the plant primary and secondary clarifiers. Another benefit of flow equalization is improved primary clarifier performance because of the influent preaeration in the equalization basin. Reduced peak flows would also allow increasing the mixed liquor suspended solids (MLSS) concentration in the aeration system and, at the same time, maintaining acceptable solids loading of the secondary clarifiers. For systems where achieving complete nitrification is essential, the increased MLSS concentration would allow increasing activated sludge system solids retention time (SRT) and decreasing the food-to-microorganism ratio, which would facilitate nitrification. Shifting treatment from high-peaking factor periods during the day to off-peak periods would also help reduce plant energy costs.

The positive effect of flow equalization on clarifier capacity has been demonstrated at the Lake Buena Vista, Florida, 76 000-$m^3$/d (20-mgd) nutrient removal plant (Hubbard et al., 2001). This plant has four biological nutrient removal (BNR) trains, each with capacity of 19 000 $m^3$/d (5 mgd), and designed to meet the following effluent discharge limits: 5 mg/L of carbonaceous biochemical oxygen demand (BOD) and total suspended solids (TSS), 3 mg/L of total nitrogen, and 1 mg/L of total phosphorus. The actual plant average influent wastewater flow was approximately 44 000 $m^3$/d (11.5 mgd). Using offline equalization in combination with online process control instrumentation enabled the plant staff to treat the entire influent plant flow of 44 000 $m^3$/d (11.5 mgd) with only one BNR train and to increase the design capacity of the existing secondary clarifiers from 35 000 to 44 000 $m^3$/d (9.3 to 11.5 mgd). Flow equalization allowed for maintaining the flowrate to the activated sludge system within 15% of the average daily flow. Flow exceeding the predicted average flow volume for a given day was directed to the offline equalization basins through an automatic slide gate and directed back into the treatment train at night, when flows drop below the average daily flowrate.

***Transient Flow Handling Using High-Rate Solids Separation.*** Performance of primary clarifiers is closely related to their surface overflow rate (SOR), as previously discussed in Chapter 2. In wastewater treatment plants with high wet weather peaking factors, oversizing primary clarifiers to handle transient flows could be avoided by using various ballasted flocculation processes, which combine addition of coagulant and settling ballast (generally microsand) to the primary clarifier influent with installation of inclined tubes (lamellas) in the clarifiers. A portion of the settled sludge or recovered ballast is recycled to the primary clarifier influent to seed the process. The addition of ballast increases the density of the floc particles by agglomeration. This results in a three- to five-fold increase of the design clarifier SOR. Typically, conventional clarifiers are designed for SORs of 33 to 49 $m^3/m^2 \cdot day$ (800 to 1200 gpd/sq ft). The use of high-rate ballasted solids separation technology allows increasing design clarifier SOR to at least 160 $m^3/m^2 \cdot d$ (4000 gpd/sq ft). Because the ballast enhances solids removal, its use in primary clarification reduces the solids and organic loading of the downstream biological treatment processes.

Currently, there are more than 50 plants worldwide using ballasted floc settling, with the largest units treating a peak flow of $1.89 \times 10^6$ $m^3/d$ (500 mgd). This type of high-solids separation facility can be designed with built-in flexibility to operate as a primary clarifier during wet weather conditions and as an effluent polishing clarifier for enhanced phosphorus removal during dry weather flows. Lamella settlers have been used with and without chemical and ballast addition to handle high-magnitude transient flows and achieve enhanced TSS and BOD removal. Additional discussion of high-rate solids separation processes is presented in Chapter 3.

***Transient Flow Handling by Increasing Clarifier Depth.*** Clarifier depth increase can effectively reduce the negative effect of transient flows on facility performance. Deeper clarifiers provide more room for sludge blanket buildup within the clarifier's thickening zone and protect the clarification zone from sludge blanket incursions. A full-scale primary clarifier performance study, completed by Albertson (1992), concluded that the maximum hydraulic overflow rate that can be processed by the primary clarifiers is proportional to the clarifier sidewater depth. Studies on full-scale circular secondary clarifiers, completed by Parker (1983) and Voutchkov (1993), indicate that deeper clarifiers are better suited to accommodate hydraulic surges and maintain desired effluent water quality.

Plants with high wet weather peaking factors (typically higher than 2.5) are more prone to clarifier sludge blanket washouts and are recommended to be designed with a sidewater depth of at least 4.3 to 5 m (14 to 16 ft). Detailed design recommendations for clarifier sidewater depth selection are presented in Chapters 8 and 9.

In conventional activated sludge plants, under daily average dry weather flow conditions, secondary clarifiers should be designed to maintain a 0.3- to 0.6-m- (1- to 2-ft-) deep sludge blanket. For BNR plant clarifiers, the sludge blanket is not recommended to exceed 0.5 m (1.5 ft) under average conditions. For municipal plants with separate sanitary and stormdrain sewer systems, clarifier blanket depth during transient flows should be allowed to temporarily rise to up to 1.0 m (3 ft). For combined sewer systems with wet weather peaking factors higher than 2.5, a transient solids blanket depth allowance of up to 1.8 m (6 ft) is suggested. In any case, a buffer distance of a minimum of 1 m (3 ft) should be provided between the sludge blanket level and the clarifier surface to maintain consistent effluent water quality.

Depth is not the only clarifier design variable that can be adjusted to accommodate transient wet weather flows. The design engineer has to consider the tradeoffs between higher clarifier depth and lower surface loading rate (Parker, 1983; Tekippe, 1986; Voutchkov, 1993; Wahlberg, 2001) and the potential advantages of activated sludge system process modifications (i.e., contact stabilization and step-feed aeration) to determine the optimum aeration basin secondary clarifier system design for handling transient flows. A hypothetical example of the potential tradeoffs between clarifier depth and SOR is depicted on Figure 12.1.

***Mitigation of Transient Flow Effect by Reducing Overall Solids Inventory.*** Transient flows often result in temporary transfer of significant amounts of activated sludge solids from the aeration basins to the secondary clarifiers. This solids transfer could quickly build a sludge blanket high enough to result in solids carryover and deterioration of clarifier effluent water quality. An alternative to providing deeper clarifiers for handling solids blanket buildup during transient flow events is to reduce the sludge blanket depth buildup in the clarifiers by decreasing the total amount of solids in the aeration basin activated sludge system (overall solids inventory). In practical terms, this means designing and operating the aeration basins at lower MLSS concentration and SRT.

For example, if an activated sludge system operates at 2500 mg/L and, during wet weather conditions, generates a transient sludge blanket of 1.8 m (6 ft), reduction of activated sludge system SRT by 40% and of the MLSS concentration in the aeration basin to 1500 mg/L would typically result in a reduction of the transient sludge blanket to approximately 1.1 m (3.6 ft) under similar operational and sludge settleability conditions. In a 2.5- to 3.0-m- (8- to 10-ft-) deep clarifier a transient solids

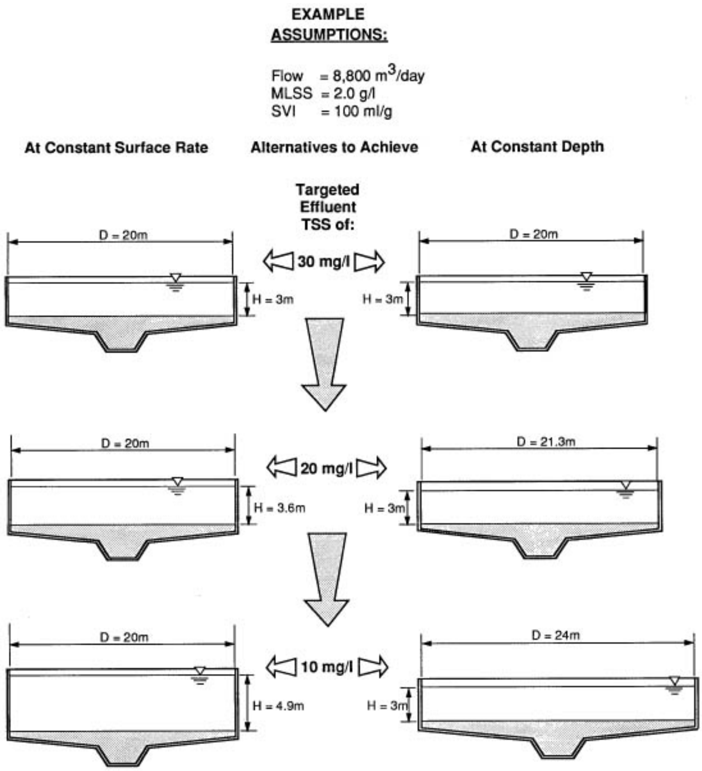


**FIGURE 12.1** Example of tradeoffs between clarifier depth and surface overflow rate (Voutchkov, 1993).

blanket of 1.8 m (6 ft) is likely to result in deterioration of effluent water quality, while a 1.1-m (3.6-ft) solids blanket would not significantly affect clarifier effluent quality.

The concept of improving existing clarifier performance by operating at lower SRTs is especially applicable to conventional wastewater treatment plants with relatively shallow clarifiers and conservatively designed solids handling systems. If such a system is currently operated at a higher sludge age and MLSS concentration and experiences frequent solids washouts because the system clarifiers cannot retain the solids that are transferred to them from the aeration basins during transient loads, switching to a new operational mode at a lower solids inventory (lower SRT and MLSS concentration) could resolve the clarifier solids washout problem and would overall improve the clarifier effluent water quality.

The example above illustrates the tradeoff between clarifier depth and reduced solids inventory as two alternative approaches for handling transient flows. Designing activated sludge systems to operate at a low solids inventory allows using shallower secondary clarifiers to achieve effluent quality comparable to that of high solids inventory systems with deeper clarifiers under transient flows.

Operation at a lower solids inventory or MLSS is often the key reason why shallow clarifiers produce effluent quality comparable to deeper clarifiers at similar or sometimes higher surface loading rates. Therefore, when comparing the effect of side water depth and SLR on secondary clarifier effluent water quality, activated sludge solids inventory is one of the key parameters that must be taken into consideration in the comparison. Otherwise, shallow clarifier performance may appear better or sometimes superior to the performance of deeper clarifiers, which may lead to the misleading general conclusion that higher side water depth provides little or no benefit for improving clarifier performance under transient loads.

Another benefit of operating an activated sludge system at a lower overall solids inventory (lower SRT and MLSS concentration) is a potential improvement of the clarifier's overall performance. Higher MLSS concentrations typically contribute to the formation of density currents in clarifiers and generally result in lower mixed liquor settling velocities (Wahlberg, 1996).

Mitigation of negative transient flow effects on clarifier performance by reducing activated sludge solids inventory is very beneficial when upgrading existing plants with shallow clarifiers and when adequate aeration basin capacity is available to achieve plant secondary treatment goals. This approach, however, may have a limited application for BNR plants targeting high levels of nitrogen removal, where maintaining a high solids inventory or SRT in the activated sludge system is needed to achieve stable nitrification and consistent effluent water quality.

***Transient Flow Control by Increase of Return and Waste Activated Sludge Rates.*** In secondary clarifiers, the effect on transient flows could also be partially mitigated by increasing the waste activated sludge (WAS) removal rate and return activated sludge (RAS) recycle rate. However, increasing the RAS recycle rate could only be effective for controlling the effect of relatively short transient events on the clarifiers (4 to 8 hours) and is limited by the capacity of the sludge collection and withdrawal systems. This strategy, however, has a limited benefit for long-lasting transient flow conditions. The main reason is that the increased RAS recycle rate only transfers sludge temporarily from the clarifiers to the aeration basins, and, after being retained for a short time in the aeration tanks, the RAS solids return back to the clarifiers. The increased RAS recycle flow will, therefore, ultimately increase the hydraulic loading of the clarifiers, and the sludge blanket will begin to rise again.

The RAS flowrate increase must be gradual and coordinated with the rate of sludge collection. Sudden increases in the RAS recycle rate may result in sludge blanket channeling ("rat-holing"). In addition, an abrupt change of the RAS recycle rate may create a hydrodynamic shockwave that may propagate quickly to the clarifier's clear effluent zone and cause excessive turbulence in the clarifiers.

The optimum design RAS recycle rate and control strategy to accommodate transient loads can be determined using clarifier solids flux state-point analysis (Keinath, 1985). Under this state-point concept, the design RAS recycle rate is established as the rate at which the clarifier is in a critically loaded condition, corresponding to a stable steady-state sludge blanket level. The design range of the RAS recycle rate is typically between 25 and 75%. The total RAS pump capacity is recommended to be designed for 120% of the average dry weather flow or 50% of the peak wet weather design flow (whichever is higher). Additional discussion of the state-point concept and its use for secondary clarifier design is presented in Chapter 4.

Installation of variable frequency drives on the RAS pumps may be warranted if the plant is exposed to frequent transient flows of a magnitude exceeding 2 to 2.5 times the daily average flow. It is recommended to provide separate RAS pumps and flowmeters for the individual plant clarifiers rather than using a common suction header and RAS flowmeters for all units. The above-described RAS recycle measure can be combined with an increase in the WAS withdrawal rate to mitigate transient load effect on the clarifier performance during extended peak flow conditions.

***Handling of Transient Flows by Activated Sludge Contact Stabilization.*** An additional measure for successful control of transient loads is temporary transfer and storage of some of the activated sludge in the aeration tanks, rather than in the clarifiers, by using a portion of the aeration tank volume as a zone of contact stabilization (sludge reaeration) fed only with RAS. The contact stabilization (sludge reaeration) zone of the aeration tanks generally is located ahead of the main aeration zone. Return activated sludge is added to the tank inlet separately and aerated for a period of time before being blended with the primary effluent, which is introduced directly to the aeration zone.

The solids balance between the aeration zone and contact stabilization zone is controlled by the RAS recycle rate. As the RAS recycle rate is increased, a greater portion of the activated sludge solids is transferred from the clarifier blanket to the contact stabilization zone of the aeration basin. These solids will be retained in the contact stabilization zone for a certain period of time (typically, 4 to 6 hours), effectively allowing for the reduction of the clarifier sludge blanket depth.

Taking into consideration that the clarifier sludge solids originated in the aeration zone of the aeration basins, increasing the RAS recycle rate will also decrease the amount of MLSS in the aeration zone of the basins (i.e., the higher rate of return will shift solids from the aeration to the contact stabilization zone of the aeration basins also). Therefore, as the RAS recycle rate is increased, the detention time of the MLSS in the aeration zone is lowered. This solids inventory shift will proportionally increase the food-to-microorganism ratio in the aeration zone, which may have a negative effect on aeration basin BOD removal and nitrification and on sludge setleability.

If the RAS recycle rate is increased to such an extent that aerobic zone MLSS and contact time are reduced significantly (to gain a rapid reduction of clarifier sludge blanket depth), the increased food-to-microorganism ratio may result in deterioration of sludge settleability, negating the positive effect of this control measure on clarifier performance. Therefore, the use of contact stabilization coupled with an increased RAS recycle rate for transient flow control has to be optimized against aeration zone contact time, secondary effluent BOD and nitrogen water quality, and sludge settleability.

Conversion of a conventional activated sludge system to a contact stabilization system has been successfully implemented at the Camp Creek Water Pollution Control Plant in Fulton County, Georgia (Danco and Dickens, 1994). At this facility, which has two aeration tanks and four shallow circular secondary clarifiers, one aeration tank has been converted to an aeration zone and the other to a contact stabilization zone. Conversion to contact stabilization allowed for improving facility nitrogen removal, minimizing floating sludge problems in the clarifiers, and simplifying activated sludge process control. Before the conversion, secondary clarifier performance has been affected by frequent solids overloading and sludge blanket denitirification.

***Handling of Transient Flows by Step-Feed Aeration.*** Step-feed aeration basin configuration allows influent flow feed at two or more locations along the length of the aeration basin. Under this configuration, the entire RAS flow is recycled to the inlet of the aeration basin, and MLSS concentration decreases along the length of the basin because each of the influent entries dilutes the mixed liquor. By maintaining the majority of the solids load to the inlet end of the aeration basin and diluting MLSS towards the outlet, the clarifier solids loading is reduced and sludge blanket level is controlled at transient-flow conditions. In effect, the step-feed configuration allows for shifting the solids inventory from the clarifier to the front end of the aeration basin, thereby reducing clarifier solids flux.

***Mitigation of Transient Flow Effects by Aeration Basin Adjustable Effluent Weirs.*** Installing adjustable effluent weirs on the aeration basins coupled with providing extended aeration basin freeboard can further reduce the transient-flow effect on the secondary clarifiers. When a flow surge occurs, the adjustable weirs are elevated, and the additional aeration basin volume allows for the retention of some of the excessive flow in the aeration tanks and, thereby, dampening of the transient effect on the secondary clarifiers. This approach is typically applicable for aeration basins with diffused bubble aeration and may have limited use for aeration tanks with surface aerators. In addition to the extra costs for constructing deeper aeration tanks, this approach for reducing transient flow effect on the secondary clarifiers may result in excessive activated sludge flock breakup resulting from elevated effluent weir drop. An effective measure for mitigating the effect of the flock breakup effect, when using adjustable aeration basin effluent weirs, is the addition of a small dosage (0.5 to 1.5 mg/L) of cationic polymer to the secondary clarifier feed. Polymer addition generally strengthens the flock structure and, at the same time, improves clarifier effluent quality.

***Mitigation of Transient Flows by Temporary Shutdown of Aeration.*** A measure that could be used as a last resort in controlling clarifier blanket depth and preventing solids carryover with the final effluent is to shut off the aeration, the internal recycle, and the mixing equipment in the activated sludge basins. This will immediately prevent additional solids from reaching the clarifiers and will allow the biomass that has been conveyed to the clarifiers to be returned back to the aeration basins (Randall et al., 1992). This measure, however, is typically applicable only to aeration basins equipped with mechanical aerators or coarse-bubble diffusers. Plants using fine-bubble diffuser systems for activated sludge tank aeration may implement this transient flow mitigation approach only for a very short period of time (typically not more than 30 minutes) without exposing the aeration diffusers to excessive fouling. If the aeration system type is not a constraint, this mode of operation can be used for 3 to 4 hours without a significant negative effect on plant effluent quality.

Hydrodynamic modeling takes into account the effect of wide influent water quality and quantity fluctuations during wet weather events and identifies the most efficient and cost-effective combination of design and control measures to handle wet weather conditions and produce a flow of target water quality. More detailed information on use of hydrodynamic models for clarifier design and performance analysis is provided in Chapter 6.

# CLARIFIERS AND PRETREATMENT FACILITIES

**EFFECT OF PLANT INFLUENT PUMPING STATION DESIGN ON CLARIFIER PERFORMANCE.** Plant influent pump size, configuration, and type of motor controls have a significant effect on clarifier performance. Wide and sudden changes in plant influent flowrate typically create hydraulic transients, which degrade clarifier effluent quality and overall clarifier performance (Collins and Crosby, 1980; Maskell and Lumbers, 1974; Porta et al., 1980). Therefore, frequent and abrupt starts and stops of large influent pumps and direct pumping into the clarifier units have to be avoided.

Installation of variable speed drives on the plant influent pumps would allow mitigating abrupt changes in clarifier influent flowrate and hydraulic loading. The use of screw pumps is recommended if feasible, because the intake configuration and mode of operation of these pumps allows dampening plant influent flow variations, if the upstream wastewater collection facilities have available flow retention capacity.

**EFFECT OF SCREENING FACILITIES ON CLARIFIER PERFORMANCE.** Plant influent wastewater contains a variety of large suspended or floating materials that must be removed to protect the structural integrity and treatment performance of the downstream treatment facilities. The type and performance of screening pretreatment equipment have a measurable effect on the performance of the primary clarifiers and, to a lesser extent, of the secondary clarifiers. There are two different types of screens: (1) fine and coarse screens, which are typically used to retain and remove large solid materials from the influent wastewater, and (2) grinders, which only reduce the size of the influent debris to smaller settleable particles and leave the grinded materials in the influent for further removal in the primary clarifiers.

The most widely used mechanically cleaned screens have bar openings between 6 and 38 mm (0.25 to 1.5 in.). The amount of screenings removed at the mechanically cleaned screens is typically in a range of 0.0037 to 0.082 $m^3$/ML of treated wastewater (0.5 to 11 cu ft/mil. gal) and averages 0.02 $m^3$/ML of treated wastewater (2.7 cu ft/mil. gal) (Qasim, 1985). The screenings typically contain 10 to 20% solids and weigh between 600 and 1100 kg/m$^3$ (40 to 70 lb/cu ft) and typically average 960 kg/m$^3$ (60 lb/cu ft).

Comminuting devices (grinders) are sometimes installed in the plant influent channel to screen and shred material to sizes from 6 to 19 mm (0.25 to 0.75 in.). The use of these devices is intended to reduce odors, flies, and cumbersome operations related to screenings removal, handling, and disposal.

If grinders are used and the grinded screenings are left in the plant influent flow, they would contribute an additional 3 to 77 mg/L (average of 20 mg/L) of TSS to the design plant influent TSS concentration. This would typically result in an average of 5 to 10% increase in primary sludge quantity. This increase could be several times higher during wet weather periods. In addition, peak daily screening quantities may vary considerably from average conditions (as much as 20:1 on an hourly basis). The sludge increase resulting from the use of grinders is measurable and must be reflected in the design of the primary clarifier sludge collection and withdrawal equipment. This sludge increase also must be taken into consideration in the solids handling facility design.

If left in the influent flow, most of the settleable screenings would be removed in the primary clarifiers. However, some of the screenings may reach the aeration basins, where they may aggregate and increase in size because of the vigorous aeration in the basins and, subsequently, may clog secondary clarifier sludge collection orifices if the clarifiers are equipped with suction sludge collection systems. Therefore, if grinders are installed as screening facilities, the use of clarifier suction sludge collection systems is not recommended. If grinders are the only viable screening process for a given application, the design of the clarifier sludge suction system and the influent grinding system must be carefully coordinated to avoid clogging of the suction system orifices and pipes. Screenings left in the plant influent may also pose settling Lamella tube clogging problems if Lamella blocks are installed for enhanced settling.

**EFFECT OF GRIT REMOVAL SYSTEM TYPE AND DESIGN ON CLARIFIER PERFORMANCE.** The main purpose of primary clarifiers is to remove mostly fine organic suspended solids settleable by gravity. Plant influent contains a relatively large amount of coarse inorganic solids, such as sand, cinders, and gravel, which are called *grit*. Grit must be removed upstream of the primary clarifiers in grit chambers to protect treatment plant equipment from excessive wear and abrasion, prevent obstruction of channels and pipes with heavy deposits that reduce their conveyance capacity, prevent cementing effects on the bottom of the primary clarifiers and digesters, and reduce the amount of inert materials in the solids handling facilities. Grit chambers are typically designed to remove particles with a specific gravity of 2.5 and retained over a 65-mesh screen.

The grit quantity and quality are important factors that must be taken into consideration in designing primary clarifiers. The quantity of grit removed in grit chambers varies significantly, depending on the type and condition of the wastewater collection system, proximity to the sea or beach areas, and type of industrial waste dischargers. The grit amount typically ranges between 0.0052 and 0.21 $m^3$/ML of treated wastewater (0.7 to 28 cu ft/mil. gal) and averages 0.03 $m^3$/ML of treated wastewater (4 cu ft/mil. gal) (Qasim, 1985). Grit typically contains 35 to 80% solids and has a specific weight in a range of 400 to 1800 kg/$m^3$ (90 to 110 lb/cu ft).

If grit chambers do not operate adequately, the excessive amount of grit left in the primary influent may cause an overload of the clarifier sludge collection equipment, may increase the amount of primary sludge, and may have a negative effect on the facilities and equipment for handling primary sludge. This excessive grit carryover may result in a measurable increase of the primary sludge quantity. If the primary sludge contains such a large amount of grit, sludge degritting before conveyance of the primary sludge to the solids handling facilities is warranted. Degritting devices (hydrocyclones and centrifuges) separate grit from the organic materials in the primary sludge and provide beneficial effect on downstream solids handling facilities. Primary sludge degritting is generally recommended as an improvement measure in existing plants with poorly performing grit chambers. In new plants, grit chamber design must be focused on effectively removing grit before it reaches the primary clarifiers, rather than on providing equipment for degritting of the primary sludge.

Aerated grit chambers have a positive effect on the primary clarification process because they reduce the potential for primary clarifier sludge septicity. Uncontrolled sludge septicity generally affects the overall clarifier performance. In addition, plant influent aeration ahead of the primary clarification reduces hydrogen sulfide concentration of the raw wastewater and, thereby, diminishes the rate of corrosion of clarifier equipment and structure.

Typically, aerated grit chambers are designed for hydraulic retention time of 2 to 5 minutes. However, if aerated grit chambers are used for preaeration or septicity control or to remove fine grit, their retention time is suggested to be increased to 10 to 20 minutes. In addition, installation of coarse-bubble aeration systems in the channels connecting the grit chamber and the clarifiers is recommended. All aerated channels must be covered and ventilated for odor and corrosion control. In case the plant influent contains a significant amount of oil and grease, aerated grit removal reduces the amount of floatables reaching the primary clarifiers. Aerated grit chambers can also be used for chemical addition, mixing, and flocculation ahead of the primary clarifiers.

## CLARIFIERS AND BIOLOGICAL WASTEWATER TREATMENT

**EFFECT OF PRIMARY CLARIFICATION ON NUTRIENT REMOVAL IN CONVENTIONAL ACTIVATED SLUDGE SYSTEMS.** The plant influent organic substrate-to-nutrient ratio is a fundamental factor affecting performance of the biological wastewater treatment systems. Generally, this ratio is measured as BOD-to-nitrogen-to-phosphorus ratio (BOD:N:P). Typically, conventional biological removal systems require a BOD:N:P ratio of 100:5:1. Primary clarification reduces the organic substrate-to-phosphorus ratio in the plant influent, thereby reducing the amount of phosphorus and nitrogen that can potentially be removed in the conventional biological treatment process (WEF, 1998a). Typically, primary clarifiers remove higher percentage of organic materials (BOD and COD) than they do nutrients (nitrogen and phosphorus). In industrial plants where the influent BOD:N:P ratio could be unbalanced, primary clarification may further negatively affect the activated sludge system BOD removal efficiency because of an inadequate amount of nutrients in the wastewater. Under such conditions, additional sources of soluble nitrogen and phosphorus may need to be added to the primary effluent to compensate for substrate-to-nutrient reduction in the primary clarifiers. This effect of primary clarifiers on the organic substrate-to-nutrients ratio must be taken into consideration when designing activated sludge systems.

Because of the negative effect of the primary clarifiers on the substrate-to-phosphorus ratio, some BNR plants have been designed without primary clarification (Randall et al., 1992). However, when primary clarification is eliminated, the downstream treatment facilities must be designed to accommodate the solids typically retained by primary clarification. Because of the significantly higher secondary sludge production (50 to 70%) without primary clarification, the aeration basins of the BNR systems must be increased in size. The elimination of the primary clarifiers also produces sludge that overall is more difficult to handle. Therefore, the decision on using primary clarifiers ahead of BNR systems must be made based on the site-specific intake wastewater characteristics, nutrient removal goals, and the type and size of the solids handling facilities of the wastewater treatment plant.

**USE OF PRIMARY CLARIFIERS FOR CHEMICAL PHOSPHORUS REMOVAL.** Phosphorus removal by addition of chemicals to primary clarifier influent is easy to implement and simple to operate. Chemicals (typically iron or aluminum salts) are added upstream of the clarifier in locations providing conditions for good mixing with the plant influent. The influent phosphorus reacts with the metal salt, forming phosphate precipitate, which is removed as sludge in the primary clarifiers. Chemical addition in primary clarifiers removes up to 90% of the particulate phosphorus in the plant influent. Chemical clarification processes, such as contact clarifiers, sludge blanket clarifiers, and claricones, have been successfully used for chemical phosphorus removal. More detailed discussion of the use of primary clarifiers for enhanced phosphorus removal is presented in Chapter 3.

A key disadvantage of chemical phosphorus precipitation is that this treatment process produces significant amounts of sludge, which results in increased solids handling and disposal costs. Typically, 2.9 mg of solids are produced per milligram of aluminum, if alum is used, and 1.9 mg of solids are generated per milligram of iron (Fe), if iron salts are applied (WEF, 1998a). However, aluminum is more efficient than iron in terms of the amount of metal needed to remove one pound of phosphorus. Theoretically, precipitating one milligram of phosphorus requires 1.8 mg of iron and only 0.87 mg of aluminum. Therefore, the total amount of solids generated from the removal of one milligram of phosphorus using aluminum salts is only slightly (10 to 15%) higher than that produced by iron salt precipitation.

In addition, the chemical phosphorus precipitation process consumes a significant amount of plant influent alkalinity (5.8 mg as calcium carbonate [$CaCO_3$]/mg aluminum and 2.7 mg as $CaCO_3$/mg iron). The use of plant alkalinity upstream of the BNR system typically has a negative effect on the nitrogen removal efficiency of the system because a significant amount of alkalinity is needed for wastewater nitrification.

If plant influent phosphorus is almost completely removed in the primary clarifiers, its concentration in the primary clarifier effluent may be insufficient to maintain an adequate biomass growth in the activated sludge system, as discussed in the previous section of this chapter. A nitrogen removal study at the Washington, D.C., Blue Plains wastewater treatment plant (Bailey et. al, 1997), where iron salts were added to the primary and secondary treatment processes for enhanced phosphorus removal, indicates that enhanced primary clarifier phosphorus removal can result in inadequate soluble phosphorus concentrations available for the denitrifying microorganisms. This deficiency at the Blue Plains wastewater treatment plant resulted in reduced BNR, erratic denitrification rates, filament growth, increased sludge yields, and inefficient use of methanol.

The type of chemical coagulant used for phosphorus precipitation must be carefully selected, because it may have a significant effect on some of the downstream treatment facilities. For example, iron compounds, unlike those containing aluminum and calcium, can also effectively control septic odors. However, if the treatment plant has a UV disinfection system, overdosing iron salts can foul the UV tubes and measurably reduce their disinfection efficiency. Residual iron also interferes with the disinfection process, because iron absorbs the UV portion of the spectrum.

Aluminum salts produce precipitates that do not re-dissolve under anaerobic conditions as iron phosphates would (Lind, 1998). This is a key consideration in wastewater treatment plants with anaerobic digesters. The use of aluminum salts, although producing slightly higher amount of solids, would typically minimize phosphate release in the anaerobic digesters and related solids handling sidestreams.

The use of chemically enhanced primary clarification may also have an effect on the final sludge quality and its disposal options. Along with phosphorus, coagulants would also precipitate heavy metals from the plant influent, thereby increasing the metal content in the plant sludge. Aluminum and iron salts used for precipitation would also contribute to the increased content of heavy metals in the sludge, because these commercial products generally contain trace amounts of metal impurities. If the plant sludge is planned to be beneficially used, the effect of chemical precipitation on final sludge quality must be carefully assessed for compliance with applicable regulatory requirements.

The costs for chemical and biological phosphorus removal are affected significantly by the plant influent five-day-BOD-to-total phosphorus ($BOD_5$:TP) ratio and the target level of effluent phosphorus concentration. Chemical phosphorus removal becomes less cost-effective as the $BOD_5$:TP ratio decreases and the effluent phosphorus target level decreases. The use of biological phosphorus removal is also favored when the incremental sludge handling and disposal costs are relatively high.

**USE OF PRIMARY CLARIFIERS FOR SOLIDS PREFERMENTATION.** Volatile fatty acids (VFAs) play a key role in the metabolism of bacteria, such as *Acinetobacter*, capable of enhanced phosphorus removal in the BNR systems. The accumulation of VFAs gives the phosphorus removal organisms a competitive edge for growth and survival in the activated sludge system.

Typically, VFAs are created during the natural fermentation process occurring in the wastewater collection system upstream of the wastewater treatment plants. This phenomenon has been observed at several plants in the United States and is typical for wastewater plants in tidewater areas, where the sewers are long and the slopes are small (WEF, 1998a). However, VFA generation in the sewers varies with temperature and may be quite low in the winter. Therefore, to provide optimum conditions for enhanced biological phosphorus removal in the downstream anaerobic zones in the BNR system, additional VFA can be generated by pre-fermentation in the primary clarifiers.

The fundamentals of design of primary clarifiers and other pre-fermentation facilities for enhanced biological phosphorus removal have been described in detail by Barnard (1984) and Randall et al. (1992). The primary clarifiers can be used to ferment organic carbon available in the plant influent to generate short-chain VFAs. Ideally, acid fermentation would provide sufficient amount of VFAs to remove phosphates biologically to levels below 0.1 mg/L as phosphorus, if the BNR system is followed by tertiary filtration (WEF, 1998a). This is achieved by operating the primary clarifier to carry a sludge blanket and slowly recycling this sludge to the clarifier inlet. Figure 12.2 illustrates the concept of "activated primary sedimentation tank" operated to maximize VFAs production.

The constant recycling of sludge seeds the incoming clarifier influent with fermenters, elutriates the VFAs from the sludge blanket, and prevents the formation of methane and hydrogen sulfide through the constant exposure to air with every recycle. Because the sludge recycle leads to a slow buildup of methane organisms, the primary sludge must be completely removed from the clarifiers regularly. The frequency of complete clarifier blanket removal is site-specific and varies seasonally.

When using two primary clarifiers, as shown on Figure 12.2, there are a number of possible operational scenarios for sludge recycle. The sludge can be separately recycled back to the influent of each tank, the sludge withdrawal pumps are connected directly to the underflow of each tank, and the sludge lines are interconnected using two-way valves. By providing the operational flexibility indicated on Figure 12.2, the underflow from one of the tanks can be pumped to the other, while the underflow of the second tank is pumped to the digesters. This configuration maintains continuous fermentation process in the clarifiers while completely removing the sludge out of one of them.

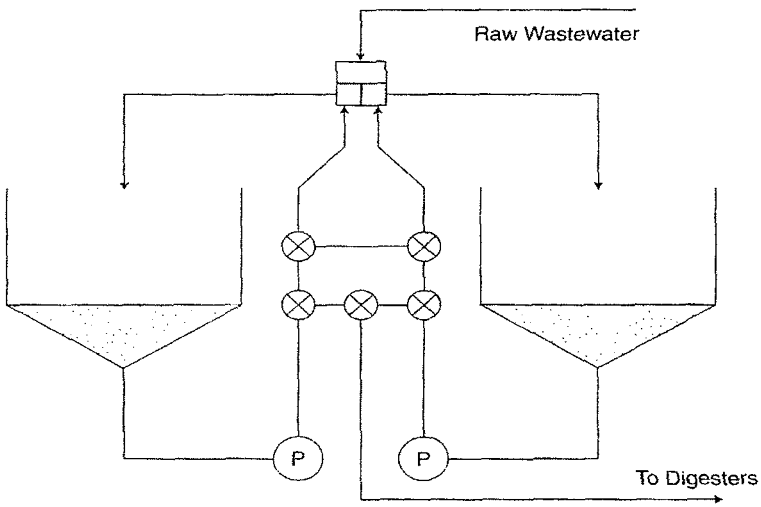


**Figure 12.2** Arrangement of two activated primary tanks (Barnard, 1984).

The key disadvantage of using primary clarifiers as pre-fermenters is that the recycled primary sludge increases the organic and solids load of the primary clarifiers and thereby reduces the available clarifier capacity. One problem experienced with primary clarifiers used as pre-fermenters is the additional load on the scraper mechanism resulting from the high sludge blanket required for this process. Therefore, in existing plants, the size and capacity of the sludge collection mechanisms must be carefully assessed to establish if primary clarifier modification to the pre-fermenter is viable. In new clarifiers, the sludge collection mechanisms must be carefully selected to maintain 0.6 to 0.9 m (2 to 3 ft) of sludge blanket.

**SECONDARY CLARIFIER DESIGN FOR ENHANCED NUTRIENT REMOVAL.** Secondary clarifier design is paramount for the successful operation of BNR systems. The clarifiers must be designed to produce effluent TSS concentration below 10 mg/L to effectively remove total phosphorus below 2.0 mg/L (Morales et al., 1991; Voutchkov, 1992). This typically requires secondary clarifiers to be designed for relatively conservative surface loading rates in a range 0.5 to 1.0 $m^3/m^2{\cdot}h$ (300 to 600 gpd/sq ft) (Sedlak, 1991).

Resolubilization of phosphorus in the sludge blanket and subsequent phosphorus release with the final effluent is a problem that typically occurs in shallow clarifiers that carry relatively deep sludge blankets. The effect of phosphorus resolubilization can be reduced by increasing clarifier side water depth so that the clarifier can be operated with a minimum upflow velocity through the sludge blanket. It is recommended to design the clarifiers with a side water depth in a range of 4.3 to 5.5 m (13 to 16.5 ft) to prevent significant upflow through the sludge blanket. The clarifier upflow and phosphorus elutriation can further be minimized by increasing the RAS recycle rate and sludge waste rate of the clarifier. The need to minimize upflow through the clarifier sludge blanket renders the use of rimflow-type clarifier (see Chapter 3) undesirable when the treatment plant includes BNR facility targeting production of effluent with low phosphorus concentration.

Biological nutrient removal systems are susceptible to induce growth of filamentous and scum-producing organisms (Sedlak, 1991). Therefore, the secondary clarifiers are recommended to be designed with scum collection and removal mechanisms. The collected scum should be conveyed to the solids handling systems.

In addition, the BNR clarifiers should be designed with provisions to handle bulking sludge. Anaerobic and anoxic selectors have a positive effect on the sludge settling characteristics and typically effectively control excessive filamentous organism growth. Alternative bulking sludge control measures and design provisions are discussed in detail in Chapters 8 and 9.

Biological nutrient removal systems, which incorporate anaerobic and anoxic selectors, have a positive effect on sludge settling characteristics and effectively control excessive filamentous organism growth. Incorporation of anaerobic or anoxic selectors to the activated sludge system generally results in improved secondary clarifier performance. A number of examples of full-scale nutrient removal systems using selectors for filamentous growth control and their effect on clarifier performance are presented elsewhere (Jenkins et al., 1993; Randall et. al, 1992; WEF, 1998a).

Biological phosphorus removal can further be enhanced by the addition of phosphorus precipitating salts (alum, ferric sulfate, ferric chloride, etc.) to the activated sludge. The phosphorus polishing chemical can be added at several locations within the activated sludge system. Typically, coagulant is recommended to be added to the secondary clarifier influent because it reduces interference with the BNR process in terms of alkalinity consumption. At this stage of treatment, phosphorus is predominantly in the form of orthophosphates, which can be precipitated and settled in the clarifiers. The most suitable points of chemical addition are locations where flash mixing can be achieved effectively (clarifier splitter boxes, flocculation wells (if such are provided), aerated distribution channels, etc.). Coagulant addition can also partially compensate for activated sludge flock sharing resulting from excessive mixing and turbulence in the aeration basins.

Metal salts addition increases the nonvolatile portion of the activated sludge system, and, therefore, higher MLSS concentration must be maintained in the aeration basins to provide the same amount of active biomass. An increased amount of solids in the activated sludge system would require the clarifiers to be designed to maintain higher sludge blanket depth and would need installation of larger RAS and WAS pumps.

Denitrification in activated sludge secondary clarifiers caused by creation of anaerobic conditions in the clarifier blanket leads to uncontrolled flotation of solids, because the gaseous nitrogen produced in the process becomes entrapped into the activated sludge flocks and floats to the clarifier surface. In addition, anaerobic conditions in the secondary clarifier sludge blanket may result in soluble phosphorus release from the biomass in the clarifier, which would lead to an increase in plant effluent phosphorus concentration. This unwanted condition could be prevented by limiting the amount of nitrates entering the clarifier, by maintaining aerobic conditions in the secondary clarifier, and by minimizing the time the sludge is retained in the clarifier. For plants targeting enhanced phosphorus removal only, the amount of nitrates entering the secondary clarifier can be reduced by designing the activated sludge system to operate at a low SRT (typically less than 6 days), thereby minimizing the presence of nitrifiers in the activated sludge biomass and eliminating nitrification. For BNR plants that have to comply with both nitrogen and phosphorus removal requirements, the amount of nitrates entering the clarifier can be achieved by denitrification in the anoxic zones of the activated sludge system.

Maintaining aeration basin effluent dissolved oxygen concentration in a range of 2.5 to 3 mg/L at all times, operating at an RAS recycle rate higher than 50%, and maintaining clarifier sludge blanket lower than 0.5 m (1.5 ft) are effective measures to prevent denitrification and phosphorus release in the secondary clarifiers.

Typically, high solids blankets tend to deteriorate effluent water quality in BNR plants. Therefore, BNR system secondary clarifiers are recommended to be designed with sludge collection and withdrawal systems that have adequate capacity to remove the sludge in a relatively short time and to maintain the sludge blanket level between 0.35 and 0.5 m (1.0 and 1.5 ft). The sludge blanket level, however, should not be allowed to drop below 0.20 m (0.6 ft) during minimum plant flow conditions, or channeling may occur in the sludge blanket, resulting in low RAS concentrations. To accommodate the sludge blanket control measures suggested above, secondary clarifiers of BNR plants with significant diurnal flowrate fluctuations are recommended to be provided with two-speed or variable speed controls of the sludge collection mechanism drive.

Another criterion that can be used to determine the acceptable clarifier sludge blanket depth and capacity of the clarifier sludge collection and withdrawal systems in BNR plants is the SRT of the clarifier sludge blanket. This parameter is calculated by dividing the mass of solids in the clarifier blanket by the rate of withdrawal of RAS and WAS solids from the clarifier. In general, sludge blanket retention time in BNR system clarifiers, estimated for daily average conditions, should not exceed 2 to 3 hours to avoid potential denitrification in the sludge blanket. The optimum time for solids retention in the clarifiers depends on a number of factors, including the type and configuration of the BNR system, concentration of oxygen in the activated sludge entering the clarifiers, wastewater temperature, SRT, and the target clarifier effluent water quality.

**OPTIMIZATION OF CLARIFIERS—AERATION BASIN SYSEM.** It is well-established that aeration basin and secondary clarifier size are interrelated, and their design can be optimized to achieve minimum life-cycle cost of the entire activated sludge system. The optimization of the aeration basin secondary clarifier system typically focuses on selection of the most cost-effective design MLSS concentration in the aeration basins. Typically, there is an optimum design MLSS that will minimize the combined capital costs of the aeration basins and the clarifiers when the clarifier capacity is thickening-limited (van Haandel, 1992). An activated sludge secondary clarifier is thickening-limited whenever the MLSS concentration exceeds some minimum concentration, when the MLSS concentration is less than the concentration at the minimum of the solids flux curve, and when the RAS concentration is greater than the critical concentration. It should be noted that the optimum MLSS concentration typically increases with the increase in system SRT. Because BNR systems generally have to operate at higher SRTs, the optimum design MLSS concentration for these systems is generally higher than that of conventional activated sludge systems. Another important factor for determining the optimum MLSS concentration is the activated sludge settling characteristics. As sludge settling improves, the optimum MLSS increases, and the overall cost of the activated sludge system is reduced.

A number of activated sludge models are currently available to determine optimum activated sludge system design. More detailed discussion of optimization models is presented in Chapter 6.

# INTERACTION WITH SOLIDS-HANDLING FACILITIES

**CLARIFIERS AND SLUDGE THICKENING.** The sludge generated during the sedimentation process is initially thickened in the primary and secondary clarifiers. The extent of sludge thickening that can be achieved in the clarifiers depends on numerous factors, such as plant influent wastewater quality and quantity, capability of the clarifiers to carry the sludge blanket, type and capacity of the sludge collection and withdrawal system, sludge settleablity, sludge septicity in the primary clarifiers, and type of biological treatment process in the activated sludge system.

***Thickening in Primary Clarifiers.*** Ideally, if the plant influent is not septic (short sewer collection system with septicity or odor control provisions, located in an area of cold to moderate climate), the grit removal facilities operate well, the primary clarifiers have adequate depth to carry 0.6 to 0.9 m (2 to 3 ft) of sludge blanket, and sludge collection and withdrawal systems are adequately sized and automated to maximize sludge concentration, then the primary sludge in the clarifiers can be consistently thickened to 3 to 6% solids. This primary sludge concentration is considered optimum from a point-of-view of sludge collection and conveyance. Thicker sludge (typically with a solids concentration higher than 6% solids) would be more difficult to remove from the clarifiers and convey to the solids handling facilities. If the primary clarifiers are designed to perform both sedimentation and thickening function, then further downstream thickening facilities are not required.

If the plant influent and primary sludge are prone to septicity, the plant experiences frequent transient flows and dry weather daily peaking flow and/or TSS load factors are consistently higher than 2, the primary clarifiers must be built relatively shallow because of site-specific constraints, and the primary sludge does not settle well, then the primary clarifiers must be designed to perform only sedimentation function and to continuously withdraw sludge. In case the primary clarifiers cannot carry a sludge blanket and sludge withdrawal is continuous, the primary sludge concentration typically varies between 0.5 and 1.5%. This sludge contains a significant amount of water and must be thickened further for cost-effective and efficient solids handling.

Successful design of primary clarifiers for maximized thickening has been reported for the City of Memphis, Tennessee, 302 800-$m^3$/d (80-mgd) Maxson wastewater treatment plant (Collins and Jenkins, 1999). The new 60-m- (180-ft-)

diameter plant primary clarifier raised primary sludge solids concentrations from 4 to 7% solids.

***Thickening in Secondary Clarifiers.*** The level of activated sludge thickening that can be achieved in secondary clarifiers depends on a number of variables, most of which are related to the type of activated sludge system and the biological treatment processes. These factors have a direct effect on the sludge settleablity, compressibility, and side effects affecting the clarification process, such as occurrence of uncontrolled denitrification in the sludge blanket, filamentous growth and sludge bulking, and pin flock. In addition, factors related to clarifier configuration, hydraulics, and sludge current distribution have a significant effect on the waste activated sludge concentration. The key factors affecting activated sludge settleability and secondary clarifier thickening performance are further discussed in detail in Chapter 4.

Typically, the TSS concentration of the activated sludge wasted from the secondary clarifiers ranges between 4000 and 8000 mg/L (0.4 to 0.8% solids). Under best-case settling and activated sludge system performance conditions, the WAS concentration may reach 10 000 to 14 000 mg/L (1 to 1.4% solids). The WAS, even under the best-case clarifier thickening scenario, contains a very large quantity of water that must be reduced before further processing in the downstream solids handling facilities.

***Co-thickening of Primary and Secondary Sludge in Primary Clarifiers.*** Co-thickening of primary and secondary sludge in primary clarifiers includes conveyance of the secondary sludge to the primary clarifiers, blending with plant influent, and cosettling this sludge with the plant influent suspended solids. Several key benefits of the co-thickening in the primary clarifier are enhanced primary sedimentation caused by the flocculating effect of the secondary sludge on the influent suspended solids, cost reduction resulting from elimination of secondary sludge thickening facility, and simplified solids-handling operations.

Successful co-thickening of primary sludge and trickling filter secondary sludge has been reported at a number of wastewater treatment facilities (Kemp and MacBride, 1990). In these plants, the primary clarifiers were designed for relatively low loading rates and were equipped with sludge collection and withdrawal systems, allowing relatively rapid sludge removal. Rapid sludge removal prevented an increase in primary clarifier effluent soluble BOD caused by the biological activity in the clarifier sludge blanket. The co-thickened concentration was in a range of 2 to 3.5% solids.

At present, co-thickening of primary and WAS in the primary clarifiers is not practiced widely, because past full-scale experience shows that this approach had a detrimental effect on the overall primary clarifier performance (WEF, 1998b). Key disadvantages of co-thickening of primary sludge and WAS in the primary clarifiers are elevated soluble BOD concentration of the primary effluent, reduction of the clarifier treatment capacity by 40 to 50%, and production of primary sludge of 1 to 3% lower solids concentration.

***Sludge Thickening Facilities.*** Generally, additional post-clarification sludge thickening is applied to minimize the volume of the solids-handling facilities. Thicker sludge requires smaller piping and pumping equipment to convey and chemicals and digester capacity to stabilize. Commonly used methods for sludge thickening are gravity thickening, dissolved air flotation (DAF), and mechanical thickening (centrifugation and gravity belt thickening).

Gravity thickening is generally accomplished in circular sedimentation basins, similar to those used for primary or secondary clarification. Gravity thickening is used to concentrate low-solids primary sludge, trickling filter sludge, and activated sludge. Thickeners are also used for combined and chemical sludge. The level of thickening achieved by gravity is typically 2 to 5 times the concentration of the feed sludge. The gravity thickeners are most suitable for low-solids primary and trickling filter sludge. Waste activated and chemical sludge are difficult to thicken by gravity. This sludge is most cost-effective to concentrate by DAF and mechanical thickening.

Primary clarifier performance has a significant effect on the downstream thickening facilities. Septic primary sludge generally thickens at a lower rate and requires special provisions for thickener gas release and odor control. Activated sludge age and settleability affect the size of the DAF thickeners and mechanical thickening equipment and the amount of chemicals needed to condition the sludge before thickening. The lower the concentration of the primary and secondary sludge, the proportionally higher the volume of the thickening facility would need to be. Additional information on sludge thickening can be found in the Water Environment Federation's® (formerly the Water Pollution Control Federation [WPCF]) Manual of Practice No. FD-1, *Sludge Thickening* (WPCF, 1980).

**CLARIFIERS AND ANAEROBIC DIGESTION.** ***Effect of Clarifier Performance on Digester Operation.*** Performance of primary and secondary clarifiers has a significant effect on a plant's anaerobic digestion process. This process is very sensitive to changes in sludge volatile organic content, quantity, and concentration. Therefore, primary and secondary sludge removal frequency, quantity, and quality must be closely monitored and controlled to avoid digester process upsets and failures. If the treatment plant is prone to frequent transient loads and significant daily variations of influent water quality and quantity, construction of sludge storage tanks ahead of the anaerobic digesters is recommended to dampen daily fluctuations of sludge quantity and quality and to provide homogenous feed to the anaerobic digesters.

To optimize digester performance, primary sludge concentration is recommended to be maintained in a range of 4 to 6% solids. Lower concentrations would result in conveying an unacceptably high amount of water to the anaerobic digesters and would affect the acid formers to methane formers ratio, which ultimately will result in destabilization of the anaerobic digestion process. Therefore, a primary sludge concentration of 1% solids or less requires thickening before digestion. Primary sludge concentrations higher than 6% solids are achievable. However, sludge at this concentration is difficult to pump and is likely to result in a negative effect on clarifier performance because of septicity.

Primary sludge contains more readily biodegradable organic compounds than secondary sludge and, therefore, yields a higher volatile suspended solids removal rate and a higher digester gas production rate. Digester foaming problems also tend to occur less frequently and are less severe when digesting primary sludge.

Currently, it is a common practice to combine primary and secondary sludge for anaerobic digestion. In this case, it is most desirable to maximize the influent TSS and BOD removal in the primary clarifiers and to minimize the amount of activated sludge production.

***Digester Hydrogen Sulfide Control by Chemical Addition to Primary Clarifiers.*** Adding oxidants, such as ferrous chloride, ferric chloride, ferric sulfate, or chlorine, could be used effectively to control the content of hydrogen sulfide in the digester gas. Typically, hydrogen sulfide emissions from digesters are limited by pertinent air quality management regulations. Without the addition of oxidizing chemical to the primary clarifiers, anaerobic digesters typically generate 1000 to 4000 mg/L of hydrogen sulfide.

Chemically enhanced primary clarification typically reduces the hydrogen sulfide concentration to lower than 40 mg/L. If both ferric chloride and chlorine are used for hydrogen sulfide control, chlorine must be added upstream of the point of ferric chloride addition. Generally, the primary sludge would float if ferric chloride and chlorine are added at the same point because of the formation of iron sulfide, which forms black fine particles that are difficult to settle. Ferrous chloride is more effective in hydrogen sulfide control in digesters than ferric chloride. However, ferric chloride more effectively removes suspended solids and phosphorus from the wastewater. The use of aluminum sulfate for hydrogen sulfide control is not as effective as the application of iron salts.

The effectiveness of hydrogen sulfide control by the addition of iron salts depends on the point of their addition and how effective the coagulant is mixed with the plant influent. Coagulants may be added before the grit chambers to take the benefit of grit chamber contact time for mixing. Other potentially appropriate locations are ahead of plant influent Parshall flumes (if used) or in the grit chamber splitter boxes, where wastewater creates adequate turbulence for efficient mixing.

***Effect of Enhanced Primary Clarification on Digester Capacity.*** Chemically enhanced primary clarification will result in enhanced clarifier suspended solids, phosphorus, and BOD removal efficiency and, therefore, will also increase the amount of sludge generated in the clarifiers. This primary sludge quantity increase must be taken into consideration in the digester design. See Chapter 3 for a detailed discussion of chemically enhanced clarification.

**CLARIFIERS AND AEROBIC DIGESTION.** Aerobic digestion is most commonly used in relatively small plants of design capacity of 18 900 $m^3$/d (5 mgd) or less (WEF, 1998b). Aerobic digesters generally process sludge from extended aeration activated sludge facilities with or without primary clarifiers. If primary clarifiers are not used, the amount of the secondary sludge increases measurably, and this extra sludge must be taken into consideration when sizing the aerobic digesters.

The aerobic digester retention time and oxygenation requirements for stabilization of a mixture of primary sludge and WAS are significantly higher than those needed for WAS stabilization only. Because of the high energy costs associated with aerobic digester aeration, for plants larger than 37 850 to 56 775 $m^3$/d (10 to 15 mgd), it is more economical to treat primary sludge separately in anaerobic digesters while aerobically digesting only the WAS. However, in small wastewater treatment plants, overall system simplicity considerations may benefit elimination of the primary clarifiers and aerobically digesting all plant sludge. Aerobic digesters may be more cost-effective when treating WAS from extended aeration facilities operating at a very high SRT because partial aerobic stabilization is already completed within the aeration basins.

Because aerobic digesters are not as sensitive as anaerobic digesters to fluctuations of plant influent quality and quantity, sludge storage in equalizing day tanks is not typically required. However, providing high-efficiency sludge thickening facilities ahead of the anaerobic digesters is recommended to minimize aerobic digester volume and associated power costs for digester mixing. While, typically, the optimum sludge feed concentration to anaerobic digesters is 3 to 4% solids, aerobic digestion favors feed sludge concentrations in a range of 4 to 6% solids. These high levels of thickened WAS can be achieved cost-effectively only by mechanical thickening equipment (centrifuges or gravity belt thickeners) and sludge conditioning before thickening. Because the mechanical thickening process is also energy-intensive, the most cost-effective level of thickening must be determined based on the life-cycle cost analysis of costs of sludge thickening and sludge stabilization.

Aerobic digesters, similar to the activated sludge systems, may frequently experience foaming problems caused by excessive growth of filamentous bacteria. If secondary clarifier WAS contains a significant amount of filaments, these microorganisms will seed the biomass of the aerobic digester and will cause or contribute to digester foaming problems. Therefore, incorporating provisions for effective control of filamentous growth to the design of the activated sludge system is of even greater importance when the sludge is stabilized by aerobic digestion.

## EFFECT OF PLANT SIDESTREAMS ON CLARIFIER PERFORMANCE.

Sidestreams from various solids handling facilities (thickener supernatant, DAF subnatant, anaerobic digester supernatant, and waste streams from sludge dewatering) are typically returned upstream of the primary clarifiers. The BOD, TSS, COD, ammonia, and phosphorus concentration of these sidestreams is several times higher than that of the plant influent. Sidestreams, such as tertiary filter backwash, may cause surges in flow. Therefore, the primary clarifiers, biological treatment system, and secondary clarifiers must be designed to handle these additional organic loads and flows. Sidestream recycle load and flow fluctuations must be minimized and, if possible, sidestreams should be recycled during low influent flow or low influent load periods.

A key advantage of aerobic digestion compared to anaerobic sludge stabilization is that it produces significantly lower strength supernatant, which minimizes the additional load on primary clarifiers and the activated sludge system. The organic strength of the aerobic digester supernatant is comparable to that of the plant influent and typically does not exceed 1% of the total plant flow.

The effect of sidestreams on the receiving primary or secondary clarifiers could be reduced significantly by their treatment before recycle. The type and size of treatment technology depends on the sidestream quality and quantity—the additional solids nutrient loads and toxic compounds the sidestream carries compared to the plant influent and the capacity of the existing clarifiers to handle the additional waste loads that would be contributed by the sidestream. If the clarifiers to which the waste streams are recycled do not have adequate solids retaining and/or handling (collection and removal) capabilities, the sidestream solids removal before recycle would be warranted.

Often, sidestreams from solids handling facilities (such as anaerobic digester supernatant, supernatant from gravity thickeners, or subnatant from DAF thickeners) contain solid particles that are much finer than those in the raw plant influent wastewater and, therefore, are more difficult to settle. The removal of these finer solids by gravity sedimentation without prior chemical conditioning may not be efficient and cost-effective. If a given sidestream has to be preconditioned by the addition of chemicals (coagulant and flocculant) to achieve measurable solids removal by sedimentation or DAF, it may be more cost-effective to add conditioning chemicals to only this sidestream and treat it separately, rather than to condition the entire blend of the plant influent and the sidestream.

On the other hand, some of the sidestreams (such as centrate from dewatering centrifuges or filtrate from dewatering pressure filter presses) generally contain a residual amount of polymer and, therefore, after blending with the other plant waste streams that do not contain conditioning chemicals and/or with the plant influent, may enhance the overall clarifier performance. This is especially true for wastewater treatment plants in which actual influent strength, in terms of TSS, is significantly lower than the concentration for which the clarifiers were designed.

Taking into consideration the potential effects of plant sidestreams on clarifier performance discussed above, the decision for cosettling of some or all of the plant waste streams in the clarifiers must be made based on a detailed site-specific cost-benefit analysis.

# CASE STUDIES

**USE OF PRIMARY CLARIFIERS FOR SOLIDS FERMENTATION AND ENHANCED PHOSPHORUS REMOVAL.** The 75 700-$m^3$/d (20 mgd) BNR plant, located in Lake Buena Vista, Florida, and operated by Reedy Creek Improvement District, has successfully tested recirculation of a portion of the primary clarifier sludge to enhance plant phosphorus removal. The plant activated sludge system applies a five-stage Bardenpho process to achieve discharge permit limits of 5, 5, 3, and 1 mg/L for carbonaceous BOD, TSS, total nitrogen, and total phosphorus, respectively.

Currently, facility effluent total phosphorus concentration is 0.8 mg/L using biological treatment with minimal enhancement by chemical precipitation. Chemical precipitation was found to be necessary because the plant's wastewater collection system is relatively short and does not provide adequate time for formation of short-chain VFAs, which are needed and used as carbon source in the anaerobic selectors of the BNR plant to achieve enhanced biological phosphorus removal. To improve phosphorus removal, the plant staff experimented with enhancing the formation of VFAs in the primary clarifiers by recirculating a portion of the primary sludge to the clarifiers. Primary sludge recirculation increased the SRT in the clarifiers, induced solids fermentation, and thereby increased the concentration of VFAs in the primary effluent. As expected, the VFA-enriched primary effluent significantly increased the overall phosphorus removal in the BNR process. However, solids recirculation reduced the treatment capacity of the primary clarifiers and limited plant capacity expansion as influent plant flow increased. Therefore, the staff decided to continue their experiments with primary sludge fermentation in a separate tank. Pilot testing completed in a 606-L (160-gal) continuously mixed fermenter, with an SRT of 10 days, indicated that VFA concentration can be increased several times (VFAs in the unfermented solids ranged between 300 and 500 mg/L, and VFA after fermentation reached as high as 1700 mg/L). Another beneficial effect of the primary sludge fermentation was the fact that it also significantly reduces the primary sludge volume. As a result of the fermentation process, the average primary sludge concentration decreased from 2.3 to 1.1%.

**OPTIMIZATION OF CLARIFIERS—AERATION BASIN SYSTEM DESIGN.** Optimization of the clarifier-aeration basin system at the 16 860 $m^3$/d (4.5 mgd) Preston wastewater treatment plant in the regional municipality of Waterloo, Canada, allowed reduction of overall system tankage requirements of up to 25% compared to the original conventional design approach (Ross et al., 1997). The clarifier aeration basin system optimization was a part of plant upgrade efforts to accommodate future plant flow increase, achieve year-round nitrification (effluent ammonia winter and summer limits of 10 and 5 mg/L, respectively), and meet relatively stringent phosphorus effluent limits of 0.6 mg/L.

The Preston wastewater treatment plant treats a combination of municipal and industrial wastewater in a conventional activated sludge process. The plant has two major contributors of industrial wastewater: (1) a potato chip factory, and (2) an automotive manufacturing facility. The potato chip industrial discharger contributes approximately 25% of the hydraulic load and 75% of the organic load of the wastewater plant. The automotive manufacturer contributes approximately 25% of the plant hydraulic load and a low portion of the organic load.

At the time of the optimization study, the plant was operating at approximately 60% of its rated capacity of 18 860 $m^3/d$ (4.5 mgd) and without a nitrification requirement. The facility treatment processes include grit removal, primary clarification, secondary treatment in an activated sludge process, and disinfection using sodium hypochlorite. The activated sludge treatment system consists of two parallel aeration tanks equipped with mechanical aerators followed by four parallel circular secondary clarifiers. The treatment plant schematic is presented in Figure 12.3. Table 12.1 indicates plant key influent water quality characteristics.

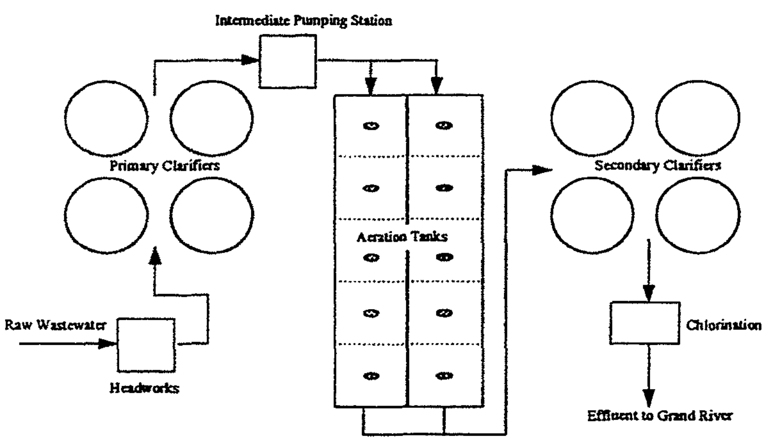


**FIGURE 12.3** Schematic of Preston wastewater treatment plant (Ross et al., 1997).

**TABLE 12.1** Average raw wastewater and primary effluent quality at Preston wastewater treatment plant (Ross et al., 1997).

| | Raw wastewater | | |
|---|---|---|---|
| Parameter | Preston wastewater treatment plant | Typical Ontario municipal | Primary effluent |
| $BOD_5$, mg/L | 397 | 170 | 222 |
| TSS, mg/L | 348 | 200 | 126 |
| Total Kjeldahl nitrogen (TKN), mg/L | 39 | 35 | 35.5 |
| Total phosphorus, mg/L | 5.9 | 6 | 4.2 |

A dynamic biological simulation model, coupled with extensive field studies, was used to determine the capacity of the existing clarifiers and aeration basins and to identify plant capacity and process upgrade measures. A commercially available plant clarifier aeration basin optimization software package that operates in the Windows® environment (BIOWIN, EnviroSim Associates Ltd., Flamborough, Ontario, Canada) was selected to complete system optimization. A comprehensive primary effluent and final effluent monitoring program was implemented to generate data needed to calibrate the model. The site-specific values of the biological growth kinetics coefficients in the model were determined using a bench-scale sequencing batch reactor that was fed with primary effluent from the treatment plant. These values are summarized in Table 12.2. Historical dynamic peak event data were applied to accurately describe key system process parameters during transient conditions.

The maximum available capacity of the existing clarifiers (maximum hydraulic and SLRs) and aeration basins (hydraulic retention time under maximum MLSS concentration of 3000 mg/L) were determined by implementing field stress tests. Combining stress testing and dynamic modeling results, the maximum daily capacity of the activated sludge system was determined to be 32 000 $m^3/d$ (8.5 mgd) and peak instantaneous capacity to be 47 460 $m^3/d$ (12.5 mgd).

The clarifier aeration basin system was optimized taking under consideration three influent sources: (1) influent source representing current flows and loadings to the plant, (2) influent source representing growth within the wastewater treatment

**Table 12.2** Calibration parameters for biological model of Preston wastewater treatment plant (Ross et al., 1997).

| Parameter/fraction | Abbreviation | Value |
|---|---|---|
| **Influent COD fractions** | | |
| Soluble readily biodegradable | $f_{bs}$ | 0.25 |
| Soluble unbiodegradable | $f_{us}$ | 0.10 |
| Particulate unbiodegradable | $f_{up}$ | 0.04 |
| **Influent TKN fractions** | | |
| Ammonia | $f_{na}$ | 0.52 |
| Soluble unbiodegradable | $f_{nu}$ | 0.04 |
| Particulate fraction of total organic biodegradable nitrogen | $f_{nox}$ | 0.56 |
| **Modified kinetic parameters** | | |
| Heterotrophic growth rate | $O_{m,H}$ | 7.0 $d^{-1}$ |
| Nitrifier growth rate | $O_{m,N}$ | 0.55 $d^{-1}$ |
| Nitrifier half saturation constant | $K_{s,NH3}$ | 0.6 mg/L |

plant service area, and (3) stormwater flow component used to simulate plant operations during transient flows. To reflect the worst-case scenario for nitrification in the aeration basins, the minimum aeration tank wastewater temperature was assumed to be 10°C (50°F).

The size of aeration tanks and secondary clarifiers are directly related by a number of process parameters, including the following: influent flow, MLSS, RAS concentration and rate, WAS concentration and rate, SRT, and the clarifier SOR and SLR. The following design boundaries were applied during the optimization process for the aeration basin and secondary clarifiers: maximum MLSS concentration of 3000 mg/L, peak instantaneous clarifier SOR of 190 kg/m$^2$·d (39 lb/d/sq ft), peak day SLR of 35 kg/m$^2$·d (859 gpd/sq ft).

Using the boundary conditions defined above, the dynamic clarifier aeration tank model simulation was applied over a range of aeration tank volumes to identify optimal secondary clarifier surface area requirements as a function of the aeration tank volume. The relationship between the aeration tank volume and secondary clarifier surface area is shown in Figure 12.4. The plot shows that the plot aeration tank

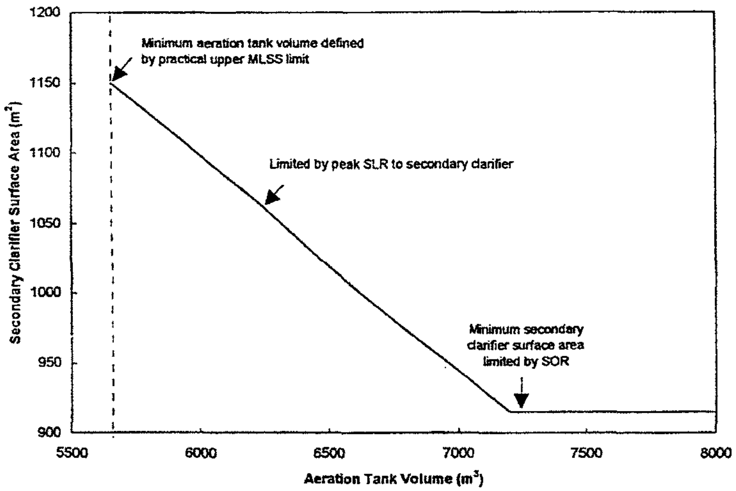


**FIGURE 12.4** Tradeoffs between aeration volume and clarifier surface area for Preston wastewater treatment plant (Ross et al., 1997).

volume must be at least 5650 $m^2$ (60 820 sq ft), based on maintaining practical average MLSS concentration of 3000 mg/L. The secondary clarifier surface area requirement for this case is 1150 $m^2$ (12 380 sq ft) and is limited by the peak SLR. With larger aeration basins, the secondary clarifier surface area can be reduced. However, the clarifier size is limited by the peak SLR. The minimum clarifier size was established as 915 $m^2$ (9850 sq ft), at which point clarifier design is limited by the peak SOR. At the minimum clarifier surface area, the clarifier tank volume that results at both peak secondary clarifier SLR (i.e., 190 $kg/m^2{\cdot}d$ [39 lb/d/sq ft]) and peak clarifier SOR (i.e., 35 $m^3/m^2{\cdot}d$ [869 gpd/sq ft]) is 7200 $m^3$ (1902 000 gal).

The optimal design secondary clarifier aeration basin configuration for upgrading the Preston wastewater treatment plant lies on a point along the sloped line of Figure 12.4, where conditions are limited by the SLR. The actual selection of aeration tankage and secondary clarifier sizing will be dependent on a number of other site-specific factors, such as site requirements and constraints, capital costs, and the ultimate site-capacity requirements.

Table 12.3 provides a comparison between the clarifier aeration basin conventional plant design based on general industry guidelines and optimal design extremes, discussed above.

**Table 12.3** Comparison between conventional and optimized activated sludge system design (Ross et al., 1997).

| Parameter | Conventional design | Application of dynamic modeling and stress testing: Minimized aeration volume | Application of dynamic modeling and stress testing: Minimized secondary clarifier size |
|---|---|---|---|
| Aeration volume, $m^3$ | 7680 | 5650 | 7200 |
| SRT, d | 12 | 10 | 10 |
| MLSS, mg/L | 2500 | 3000 | 3000 |
| Clarifier surface area, $m^2$ | 1340 | 1150 | 915 |
| Peak SLR, $kg/m^2{\cdot}d$ | 120 | 190 | 190 |
| Peak SOR, $m^3/m^2{\cdot}d$ | 24 | 28 | 35 |

Analysis of the results presented in Table 12.3 indicates that the optimization of secondary clarifier aeration basin system achieves significant reduction of the overall system capacity and cost compared to conventional design using general industry guidelines and practices. Depending on the scenario considered for the site-specific conditions of the Preston wastewater treatment plant, aeration tank size could be reduced by 6 to 27%, and secondary clarifier size could be decreased between 15 and 31%, compared to conventional plant design.

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# Appendix A

# Settling Test Procedure

Based on the procedure of Larsson (1986), the need for clarification area can be predicted from jar settling tests.

(1) The tests should be made at the intended operational temperature. Ensure thermal equilibrium in sample and test cylinders before tests are started.
(2) Fill up a 1000-mL graduated cylinder with the suspension to be tested.
(3) If the suspension is to be pretreated by chemicals before settling, then chemicals should be added in the test cylinder and the pretreatment should be carried out. After pretreatment is complete, the sample should be stirred gently.
(4) When the convective eddies from the stirring have ceased, start the stopwatch.
(5) To simulate the clarity of the overflow at a specific loading rate, pipette or siphon out the upper 100 mm of the suspension column in the cylinder after a time $t_1$, chosen to simulate a suitable surface load. Be careful to withdraw the sample just under the liquid surface. To determine an optimum loading rate, several tests (each in a separate graduated cylinder) should be made at different loading rates and at corresponding times $t_1$, $t_2$, $t_3$, etc.
(6) The following analyses should be made: concentration of suspended solids in feed and concentration of suspended solids in the pipette 100 mm from each graduated cylinder.

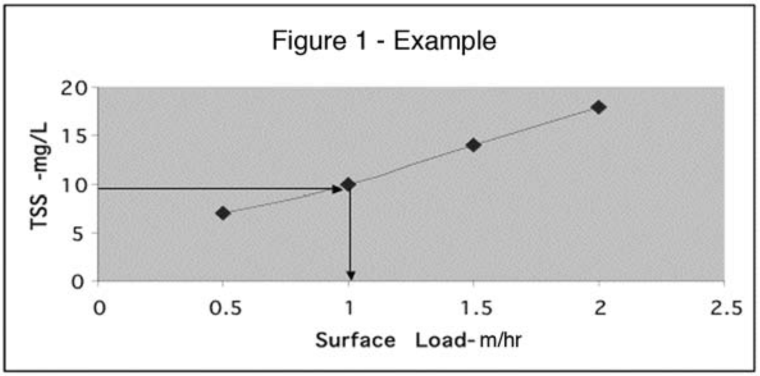
## EVALUATION OF CLARIFICATION TEST

The surface loading rate simulated by the cylinder tests will be

$$\frac{Q}{A} = \frac{0.1}{t}$$

where $t$ is in hours and $Q/A$ is in meters per hour (multiply by 589 to get gpd/sq ft) (if using a 2000-mL cylinder, use 200 mm and the formula is $0.2/t$).

Because a number of siphoning off tests are made, a set of corresponding values for surface overflow rate and settling time will be obtained. From this diagram, the proper surface load on settling area can be chosen according to requirements of overflow clarity.




## EXAMPLE

Data for a liquid to be clarified is shown in Figure A.1. Find the surface area needed to each 10 mg/L ESS. From the diagram, it can be seen that the maximum surface load (surface overflow rate) should be 1.0 m/h. This means that the (projected if Lamella clarifier) clarification area can be obtained when the feed flow is known. If flow is 10.0 $m^3$/h, area needed would be $Q$/surface load = 10 /1 = 10 $m^2$.

Note: originally developed for Lamella separators but applicable to any type system treating type 1 or type 2 suspensions.

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