

Harvesting of marine microalgae by electroflocculation: The energetics, plant design, and economics



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HIGHLIGHTS

- Cost estimation of electroflocculation for harvesting marine microalgae.
- Energy consumption was optimized by mixing and electrode separation.
- Cost estimation includes plant construction, energy and metal dissolution.
- Processing cost is calculated to be \$0.19 kg⁻¹ of algal mass.

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ABSTRACT

Microalgae have the potential to be the feedstock for biofuels and laboratory scale electroflocculation was studied as a harvesting technique for marine microalgae. The effects of the electrode separation and mechanical mixing on the energy consumption were also assessed. Results were used to design a commercial scale electroflocculation plant for the estimation of the harvesting cost. By combining electroflocculation with mixing and settling, an overall energy consumption of 0.33 MJ m⁻³ has been achieved. On a large scale, the mixing can be made energy efficient by the use of a baffled hydraulic mixer. The total cost for the harvesting, including electrical energy, electrode metal dissolution and capital depreciation, is estimated to be \$0.19 kg⁻¹ of the ash free dry mass. Hence, electroflocculation has the potential to be more economical than other harvesting techniques for marine microalgae.

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1. Introduction

Some species of marine microalgae, such as *Pleurochrysis carterae*, offer advantages such as high specific growth rate (0.54 d⁻¹), high lipid yield (21.9 t ha⁻¹ y⁻¹) and do not compete with farm produce or native vegetations for valuable resources such as fresh water or arable land [1]. These advantages make marine microalgae attractive as a feedstock for the production of biodiesel. Industrial processes suitable for the separation of microalgal biomass from the culture media include flocculation, centrifugation, filtration or sedimentation; these processes can be used individually or in combination [2]. Despite some physical limitations such as small sizes of microalgae (of cell diameters typically <15 μm) and their low biomass concentrations (ash free dry mass concentrations typically ranged from 0.2 to 0.5 kg m⁻³ under autotrophic conditions), these separation methods are quite capable of separating the biomass from the surrounding media [3]. The major

difficulty is the high production cost of microalgal biomass which has been estimated to be about \$10 kg⁻¹ [4], while plantation oil such as canola can be available at about \$1 L⁻¹ [5]; hence, the success of microalgal biofuels depends very much on the economy of production. Harvesting cost has been estimated to be about 25% of the total production [6]; hence, it is important to include cost estimation as part of the evaluation of the harvesting process.

Flocculation increases the settling rate by congregating suspended particles; this process has lower energy requirements than centrifugation and is able to increase the biomass concentration from about 0.4 kg m⁻³ in the feed stream to about 10–20 kg m⁻³ after this process. Flocculation may be induced through the application of various forms of flocculants such as inorganic, organic, polymeric [7] or through processes such as auto-flocculation [8], electroflocculation [9–11] or microbial flocculation [2,12]. The flocs collected may contain a certain amount of flocculant; therefore, flocculation may not be suitable for some purposes.

Electroflocculation is a process that uses electric currents to dissolve sacrificial metal to supply the ions required for the flocculation. In comparison with auto-, bio- or microbial flocculation, electroflocculation is a physical/chemical process that has the

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Nomenclature

A	surface area (m^2)	Q_m	medium flow rate ($\text{m}^3 \text{h}^{-1}$)
B	channel length (m)	q	overlap ratio
Ca	Camp number	RE	recovery efficiency
C_{op}	operating cost ($\text{\$/m}^{-3}$)	Re	Reynold's number
D_{al}	dissolution efficiency for aluminium	R_h	hydraulic radius (m)
d	channel depth (m)	r	depth ratio
d_h	hydraulic diameter (m)	V	applied voltage (V)
d_s	distance between the electrodes (m)	v	linear flow velocity of media (m s^{-1})
G	velocity gradient (s^{-1})	W_b	baffle thickness (m)
I	current (A)	W_c	channel width (m)
M	molar	∇	volume of fluid in the flocculator (m^3)
m	million	K	specific conductivity (S m^{-1})
N	number of channels	η_{IR}	internal resistance potential drop (V)
N_{cell}	cell count per unit volume	η_k	kinetic overpotential (V)
N_o	cell count per unit volume at time 0	η_m	mass transfer overpotential (V)
N_t	cell count per unit volume at time (t)	η_n	total overpotential (V)
P	power (W)	ρ	density of media (kg m^{-3})
pB	slot width ratio	θ	fluid retention time (s)
Q_L	medium flow rate (L min^{-1})	μ	dynamic viscosity (Pa s)

advantages of being non-species specific, simpler to operate and results are more predictable. Unlike chemical flocculation such as the use of alum or ferric chloride, electroflocculation does not introduce unnecessary anions such as SO_4^{2-} or Cl^{-1} which can result in the lowering of pH [13]. The construction of the electroflocculation cell is also relatively simple; it consists of a container with electrode plates and a direct current power supply, and hence involves modest capital investment. For these reasons, electroflocculation has been selected as a harvesting technique for the marine microalgae, *Tetraselmis* sp., in this study.

1.1. Aim

As stated earlier, the success of microalgal biofuels depends very much on the economy of production, but available literatures on microalgae recovery by electroflocculation have only discussed the energy consumption with the cost of metal dissolution neglected. By investigating the effects of: (i) mixing, and (ii) electrode separation on energy consumption during electroflocculation, this study aims to estimate the cost of capital investment, energy consumption and electrode dissolution for such a process; subsequently, by incorporating these major costs, determine the cost of microalgae harvesting on a $\text{\$/kg}^{-1}$ basis.

1.2. Electroflocculation

Aluminium electrodes were chosen for this study as they were shown to be more effective in flocculation than iron or stainless steel [14]. A simplified electrode dissolution process for aluminium is shown below:

Reactions occurring at the anode surface are:



The major products are H^+ , Al^{3+} and O_2 gas.

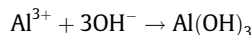
Reactions occurring at the cathode surface are:



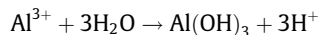
The major products are OH^- , and H_2 gas.

In addition to the electrochemical reactions listed above, anodic aluminium will dissolve to form hydroxides depending on the pH as below:

Under alkaline conditions



Under acidic conditions



The Al^{3+} ions are hydrated and react spontaneously to produce hydroxides and/or polyhydroxides such as $\text{Al}(\text{H}_2\text{O})_6^{3+}$, $\text{Al}(\text{H}_2\text{O})_5\text{OH}^{2+}$, or $\text{Al}(\text{H}_2\text{O})_4^{2+}$, and the hydrolysis of these products forms a range of mono- and polymeric species such as, $\text{Al}(\text{OH})_2^+$, $\text{Al}(\text{OH})_2^+$, $\text{Al}_2(\text{OH})_4^+$, $\text{Al}_6(\text{OH})_{15}^{3+}$, $\text{Al}_7(\text{OH})_{17}^{4+}$ etc., over a wide range of pH [15]. These polymers are hydrated and able to trap suspended particles.

From these reactions, it can be seen that major operating costs of electroflocculation are the cost of electricity and the replacement cost of the aluminium electrodes. These two costs are discussed in Sections 1.2 and 1.3.

1.3. Energy requirement

The energy requirement for electroflocculation ranges from 0.976 MJ m^{-3} for the removal of 0.95 mass fraction of a mixture of microalgae from rinsing water [9], to 9 MJ m^{-3} for the removal of a 0.90 mass fraction of a mixture of hydrocarbon, microalgae or suspended solids from wastewater [11,16]. One way to increase the efficiency of floc formation is to combine mechanical mixing and flocculation followed by settling [17–19]. Mixing provides charged particles with the momentum necessary to overcome the electrical double layers and also increases the probability of contact among particles for the floc formation. The mixing mechanisms are well understood and widely practised in the chemical flocculation in wastewater treatment, but have rarely been discussed in the electroflocculation of microalgae as a harvesting of technique. Harif and Adin [18] and Ofir et al. [17] discussed the effects of mixing in terms of flocculation efficiency and floc formation in wastewater

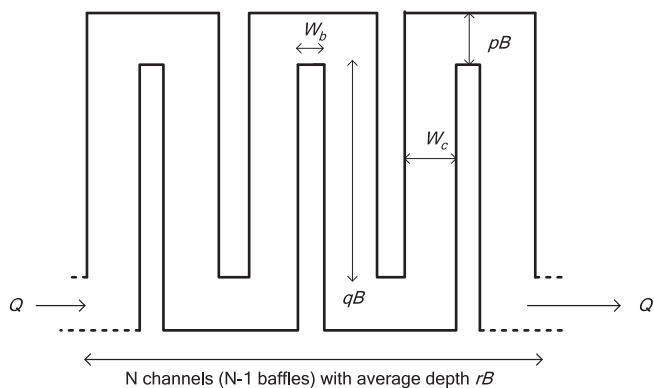


Fig. 1. Top view of a schematic layout of a baffled hydraulic flocculator, where N is the number of channels; B is the channel width between baffles; p is the slot width ratio (with respect to B); q is the overlap ratio (with respect to B); W is the baffle thickness and Q is the fluid flow [22].

treatment, but the effect on energy optimisation, which is essential to the economics of microalgal harvesting, has not been investigated. Finally, all studies mentioned in this section were based on essentially fresh water environments, whereas the culture medium for marine microalgae has a salinity of over 0.5 M and an electrical conductivity of about 5 S m^{-1} [20]. This higher conductivity may affect the energy required for the flocculation.

The energy requirement for electroflocculation will comprise of two major components; namely, mixing and electrical (resistance, electrolysis and electrode dissolution). These two components are discussed in Sections 1.3.1 and 1.3.2 respectively.

1.3.1. Mixing

Both the demand for energy and the necessary economy of scale require the microalgal culture to be processed in the order of thousands of cubic meters per hour [21]. On such a scale, energy efficient mixing is essential and can be provided by the use of a baffled hydraulic flocculator [22]. An optimal flocculator has a slot ratio 0.9–1.1, overlap ratio of 4–5 and depth ratio of 1–3 [22] and a schematic presentation of such a flocculator is shown in Fig. 1.

This type of flocculator has the advantages of simplicity in construction, no moving parts, low maintenance and do not induce short circuiting during mixing [22–24]. Furthermore, the baffles and electrodes can be integrated into one single unit and such design has been proposed in wastewater treatment plants [25].

There are two criteria for the mixing power consumption of the design of a baffled hydraulic flocculator: a velocity gradient, G , in the range $100\text{--}10 \text{ s}^{-1}$ and a Camp number, Ca , in the range $10^4\text{--}10^5$ [26,27].

The velocity gradient, G , is represented by:

$$G = \sqrt{\frac{P}{\mu \nabla}} \quad (1)$$

where P is the power consumed by the media, W ; μ is the dynamic viscosity, Pa s ; and ∇ is the total volume of fluid in the flocculator,

m^3 . The velocity gradient can therefore be interpreted as the root mean of the power dissipation per unit volume of the flocculator.

Camp number, Ca , is represented by:

$$Ca = G\theta \quad (2)$$

where θ is the fluid retention time in seconds; Ca indicates the total energy dissipated by each individual volume as it flows through the flocculator.

Lee et al. [21] calculated that by using a baffled hydraulic system, an algal culture of volume $3.375 \times 10^4 \text{ m}^3$ has a daily mixing energy requirement of 43.2 MJ [21]. In comparison, the minimum energy required for the electroflocculation of the same volume of algal culture is calculated to be 18.2 GJ (Table 1). This shows that mixing energy has an order of magnitude 10^{-3} of that by electroflocculation and can be neglected. For this reason, only the design and capital cost of the hydraulic mixer, but not the mixing energy evaluation, will be discussed in this study.

1.3.2. The effects of electrode separation on the energy requirement

The total overpotential, η_n , that is necessary for the flow of electric current between the electrodes in the flocculation tank, comprises of three components as indicated by [28]:

$$\eta_n = \eta_k + \eta_m + \eta_{IR} \quad (3)$$

where η_k is the kinetic overpotential due to contributing factors such as flow dynamics or gas evolution, η_m the mass transfer overpotential due to the diffusion of ions, and η_{IR} is the internal potential drop due to the resistance of solutions and electrode deposits. The resistance due to the electrolytic solution can be determined by:

$$\eta_{IR} = \frac{Id_s}{A\kappa} \quad (4)$$

where I is the current, A ; d_s is the distance of separation between the electrodes, m ; A is the active electrode surface area, m^2 and κ is the specific conductivity, S m^{-1} . η_{IR} decreases with the reduction in distance between cathodes and anodes and hence the energy requirement. In an industrial scale electroflocculation plant, a larger number of electrodes correspond to higher installation and maintenance costs. For this reason, it is necessary to optimize the number of electrodes required.

Some data are available from electroflocculation literature with respect to d_s , removal efficiencies and energy consumption; they are presented in Table 1 [9,10,29–31]. Due to the absence of a common algal species and cell concentration in the feed stream, it is not possible to make a strict comparison among these results. Nevertheless, these data do provide an idea of the magnitude of energy required by electroflocculation for the algae removal.

The values of d_s range from 2 or 3 mm to 26 cm and were chosen quite arbitrarily by their respective authors. Table 1 shows no apparent correlation between the energy consumption and values of d_s . Finally, except for Vandamme et al. [31], all these experiments were performed in an essentially fresh water environment and the energy requirement in a marine environment is expected to be lower due to the higher electrical conductivity.

Table 1
Electrodes separation d_s and energy consumptions.

	Electrode separation, d_s (cm)	Removal products and efficiencies (%)	Energy consumption (MJ m^{-3})	References
1	0.3	Industrial sewage (95–99%)	1.44	[29]
2	1	Algae in drinking water (100%)	1.44	[10]
3	2–4	Algae in lake water (100%)	36	[30]
4	4.4	Marine microalga <i>Phaeodactylum tricornutum</i> (80%)	0.54	[31]
5	26	Algae in rinsing water (95%)	1.19	[9]

1.4. The dissolution cost of electrodes

In addition to the electrolytic dissolution as determined by Faraday's Law, many metal can be chemically dissolved during electrolysis, these electrodes are corroded by chemical reactions where the electrons released by the sacrificial metal are not part of the electrolytic current. The current dissolution efficiency for aluminium, D_{al} has a value between 1.4 and 2, i.e., it will be dissolved at a rate 1.4 to twice of that determined by Faraday's Law [32].

Eq (5) was developed by Donini et al. [33] to estimate the operating cost of electroflocculation, C_{op} , in $\$/\text{m}^{-3}$:

$$C_{op} = 0.03917(0.02127V + D_{al}) \times \frac{I}{Q_L} \quad (5)$$

where V is the applied voltage, V; D_{al} is the aluminium dissolution efficiency, no unit; I is the current, A and Q_L is the volumetric flow rate, L min^{-1} .

Donini et al. developed this model base on an electrical energy cost of $\$0.05 \text{ kW h}^{-1}$ and aluminium cost of $\$7 \text{ kg}^{-1}$. The first term on the right hand side of Eq. (5) represents the electrical energy cost while the second term represents the aluminium dissolution cost. This equation suggests that for a typical applied voltage between 10 V and 20 V and a typical D_{al} value of 1.5, the aluminium dissolution cost is 3.5–7 times more than the energy cost and such cost cannot be neglected.

2. Materials and methods

2.1. Microalgae

Tetraselmis sp. was used for this electroflocculation experiment. The starting culture was obtained from the Algae R&D Centre, Murdoch University, Western Australia. The microalga was grown in $f/2$ medium at 22°C on a shake table in a culture tube, $f/2$ medium was prepared by the addition to 1 L of filtered (pore size $0.45 \mu\text{m}$) sea water using the following chemicals [34]: NaNO_3 , 75 mg; $\text{Na}_2\text{HPO}_4 \cdot 3\text{H}_2\text{O}$, 5 mg; Na_2EDTA , 4.36 mg; $\text{FeCl}_3 \cdot 6\text{H}_2\text{O}$, 3.15 mg; $\text{CuSO}_4 \cdot 5\text{H}_2\text{O}$, 10 μg ; $\text{ZnSO}_4 \cdot 7\text{H}_2\text{O}$, 22 μg ; $\text{CoCl}_2 \cdot 6\text{H}_2\text{O}$, 10 μg ; $\text{MnCl}_2 \cdot 4\text{H}_2\text{O}$, 18 μg ; $\text{NaMoO}_4 \cdot 5\text{H}_2\text{O}$, 6.3 μg . The culture medium has a salinity of 0.5 M and the pH was adjusted to about 8.2 by the addition of sodium bicarbonate prior to the introduction of the culture. The contents were later introduced to a 250 mL Erlenmeyer flask on a shake table and finally to a 20 L carboy photo-bioreactor. Mixing was provided by air sparging from the bottom of the drum; lighting was supplied by tri-phosphorus fluorescent tubes with an intensity of $150 \mu\text{mol photons m}^{-2} \text{ s}^{-1}$ with a 12 h/12 h light/dark cycle. Upon stationary phase, the medium salinity was raised by 0.1 M every 3 days and sub-cultured until the culture medium had a final salinity of 1 M. The culture was then transferred to an open

raceway pond of length 2.0 m, width 1 m and depth 0.2 m situated in the University of Adelaide.

2.2. Electroflocculation

2.2.1. Effects of mixing on energy requirement

The experiment on the effects of mixing on energy consumption was carried out in a clear acrylic tank as shown in Fig. 2. The tank had an electrolytic compartment of 0.0048 m^3 (length: width: height of 0.21 m: 0.095 m: 0.24 m respectively) with two vertical flat aluminium electrodes ($0.19 \text{ m} \times 0.07 \text{ m}$). These two electrodes were placed with one side against the wall as close as practicable with the separation between them d_s , varied by rectangular boxes of various sizes. The compartment was lined with a sheet of clear LDPE to prevent leaking. The algal culture has a pH of 8.4 and the temperature was kept constant at 22°C .

The electrolytic compartment was filled with the culture medium to a pre-determined level. Three runs were performed. In Run 1, the supernatant from the overnight settling was used, so that the effect on the difficult to settle microalgal cells could be observed. In Runs 2 and 3, microalgal cells were obtained directly from the open raceway pond. An electric current (typically about 10 V, 5 A) was passed for 60 s and 30 s respectively with an electric stirrer (20 rad s^{-1}) to achieve a uniform cell suspension. After the predetermined durations of electroflocculation, 150 mL aliquots were transferred to 250 mL conical flasks and were left on an orbital shaking table (7.5 rad s^{-1}) for 15 min. The flocculated algae were then left to settle under gravity for 30 min. Samples were then taken from the supernatant and the cell concentration was determined by the number of cell count per ml, N_{cell} . All tests were performed in quadruplicates with electrodes scrubbed and the polarity reversed after each run.

2.2.2. Electrode separation

The same rig depicted in Fig. 1 but without the stirrer was used for the electrode separation experiment. The distance between the two electrodes was varied between 5 cm and 30 cm in steps of 5 cm by using boxes of various lengths. Electrodes were sandpapered gently and polarities were reversed between each run. The current was set as close as practicable to 1.33 A so that current densities were approximately 100 A m^{-2} . A black and white Secchi disk was placed against the back of the tank during electroflocculation and suspended microalgae were allowed to be carried to the top by the rising air bubbles generated from the electrode surface. The applied current was stopped as soon as the marking on the disk became visible; the applied current, voltage and duration were recorded and the underflow samples were taken at 6 cm below the medium surface. Test for each electrode separation was repeated three times and test runs were randomized. Flocculation efficiencies were determined by cell counting. The visibility of

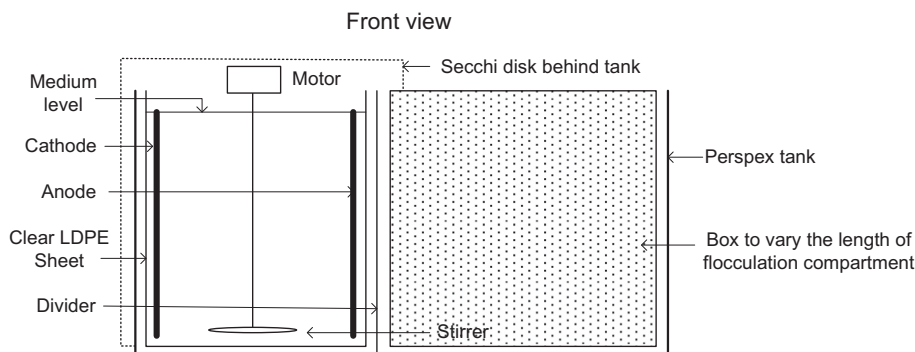


Fig. 2. Front view of the tank with adjustable distance between cathode and anode.

the disk provided a reasonably consistent indication of the cell concentrations and hence the electrical energy required to achieve such a predetermined level of recovery can be calculated. There is no additional mechanical mixing of the culture medium other than that generated from the rising of the gas bubbles; therefore, this series of test runs also served as controls for the experiments on mixing.

2.2.3. Determination of flocculation effectiveness

The effectiveness of flocculation was determined by the recovery efficiency, RE , which is defined as the ratio of the recovered biomass to the total biomass. The recovery efficiency is determined by:

$$RE = 1 - (N_t/N_o) \quad (6)$$

where N_o is the cell count per unit volume of the original cell suspension and N_t is the cell count per unit volume of the clarified cell suspension after a electroflocculation time of t s.

2.2.4. Operating cost of electroflocculation

The operating cost estimation according to Eq. (5) is based on the cost of energy and materials in 1994 [33]. By substituting the average current price of electricity at $\$0.25 \text{ kW h}^{-1}$, remade aluminium at $\$1.992 \text{ kg}^{-1}$ [35] and the flow rate, Q_m , converted to $\text{m}^3 \text{ h}^{-1}$, Eq. (5) is modified to the following equation:

$$C_{op} = 6.687 \times 10^{-4} (0.3738V + D_{al}) \frac{I}{Q_m} \quad (7)$$

Eq. (7) shows that for an average D_{al} value of 1.5 and a voltage of approximately 4 V or less, the aluminium dissolution cost will be higher than the energy cost.

3. Results and discussion

3.1. Results on effects of mixing

The results for the mixing experiment were summarized in Table 2. Test samples for Run 1 consisted of supernatants from overnight settling and consisted of cells that were difficult to settle. Test samples for Run 2 and 3 were obtained directly from the open pond prior to the experiments. Samples for Run 1 had lower initial cell counts and lower average recovery efficiency, RE , of 0.85 when comparing with those for Run 2 and 3.

Run 2 had an electroflocculation time t of 60 s but the recovery efficiency, RE , of 0.95 is only marginally better than that from Run 3 which has a electroflocculation time of 30 s and a recovery of 0.92. The electrical energy consumption for this run is 0.328 MJ m^{-3} . The parameters for Run 3 are chosen as basis for the design of the large scale electro-flocculator.

3.2. Results on the effect of electrode separation on energy consumption

The visibility of the Secchi disk provided a reasonably consistent indication of the cell concentrations and hence the electrical

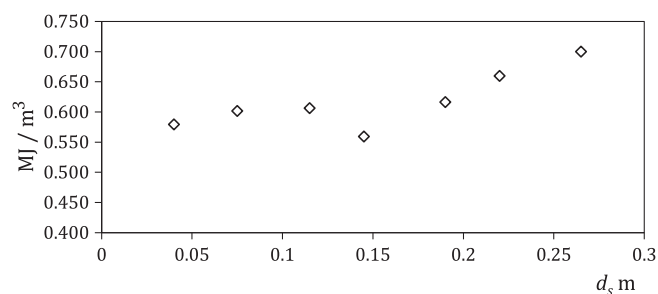


Fig. 3. Plot of energy requirements (MJ m^{-3}) vs. electrode separation.

Table 3

Comparison of the two electroflocculation processes, A and B.

	A. electroflocculation + mixing + settling	B. electroflocculation + flotation
Minimum energy requirements, MJ m^{-3}	0.328	0.559
Recovery	0.91	0.87
Product stream dry mass concentration, kg m^{-3}	10	30–40
Overall processing time, min	45.5	5.75

energy required to achieve a predetermined level of recovery for each separation. The applied voltage ranged from 1.4 V to 3.4 V with flocculation time between 169 s and 550 s depending on d_s ; the recovery efficiency as determined by the visibility of the disk has an average RE of 0.87 and a standard deviation of 0.03. The overall uncertainty was approximately 4% mainly due to the determination of the end point by using the disk.

Fig. 3 shows electroflocculation energy requirements in MJ m^{-3} for various electrode separations in a culture medium of salinity 1 M. The energy consumption is virtually flat up to a d_s value of 0.15 m followed by a steady increase; therefore, an electrode separation distance of 0.15 m was chosen for the design of the electro-flocculation plant.

A comparison on the effects of the two electroflocculation processes, A (mixing + settling) and B (flotation) is summarized in Table 3:

Comparing Process A and B, Process B has a much shorter processing time and a higher product concentration, these two advantages will result in a processing plant with a smaller foot print area and a more concentrated product; however, Process B also has an energy requirement that is about 1.7 times higher than that of Process A. The higher energy required is mainly due to the generation of gas bubbles necessary for the floc floatation. Another consideration is the requirements from further downstream processes, for example: microalgal slurry with a dry mass concentration of about 3% has a creamy consistency and will be difficult to be handled by cell disruption processes such as high pressure homogenizers [36]. As processing energy cost and cell disruption are important consid-

Table 2

Recoveries and energy requirements from the electroflocculation.

Run	Voltage (V)	Current (A)	Flocculation time, t (s)	N_o^a ($\times 10^4 \text{ mL}^{-1}$)	N_t^a ($\times 10^4 \text{ mL}^{-1}$)	Recovery ^a RE	Energy requirement ^b (MJ m^{-3})
1	5.0	9.9	60	31.5	4.5	0.85	0.623
2	5.2	9.9	60	135.5	6	0.95	0.652
3	5.3	9.9	30	149.5	11.5	0.92	0.328

^a N_o, N_t : cell counts per ml at time, $t = 0, t$ s respectively; approximately 4×500 cells were counted during each run and the maximum uncertainty is plus or minus 5%. [46].

^b The uncertainties in voltage, current and time were approximately 2%, 1% and 3% respectively, therefore, the energy requirement has a maximum uncertainty of 6%.

Table 4
Growth parameters for typical *Tetraselmis sp.*

Growth and design parameters	Values
Biomass concentration, kg m ⁻³	0.25 [37,38]
Lipid fraction	0.25 [39,40]
Litres per barrel	159
Lipid production, m ³ d ⁻¹	39.37
Biomass production, kg d ⁻¹	1.575 × 10 ⁵
Processing capacity, m ³ d ⁻¹	6.3 × 10 ⁵
Processing capacity, m ³ s ⁻¹	7.29

erations in the production of biofuels, Process A is chosen as the harvesting technique and the basis for this plant design.

3.3. Design and cost estimation of the electroflocculation harvesting plant

The commercial scale microalgae harvesting plant will consist of three modules, namely, electroflocculation, baffled hydraulic mixing and gravity settling. The cost estimation is based on a plant with an annual production capacity of 100,000 barrels of lipids. Base on typical microalgal growth parameters presented in Table 4 [37–40], the plant requires a pond surface area of approximately 1 km² and culture medium processing capacity of 6.3 × 10⁵ m³ d⁻¹ (7.29 m³ s⁻¹).

3.3.1. Electroflocculation module

The minimum linear flow velocity to prevent flocs settling is 0.075 m s⁻¹ [23], and based on the electroflocculation time of 30 s (Section 3.1), the length of the electro-flocculation channel is: 0.075 m s⁻¹ × 30 s = 2.25 m.

The dimensions of commercially available aluminium sheet are 1.2 m × 2.4 m × 1 mm with a weight of 7.8 kg. To minimise fabrication costs, each electrode can be made of one aluminium sheet with a separation of 0.15 m between electrodes; therefore each channel is 2.4 m × 1.2 m × 0.15 m. The linear flow velocity will be increased slightly from 0.075 m s⁻¹ to 0.08 m s⁻¹ to keep the electroflocculation time for the same duration of 30 s.

The cross-sectional area can be obtained by dividing the volumetric flow rate with flow velocity. The required cross-sectional area is therefore:

$$7.29 \text{ m}^3 \text{ s}^{-1} / 0.08 \text{ m s}^{-1} = 91.125 \text{ m}^2.$$

The working width of the combined electroflocculation module can be determined based on the required cross sectional area and electrode height.

The working width is cross sectional area divided by channel height:

$$91.125 \text{ m}^2 / 1.2 \text{ m} = 75.937 \text{ m}$$

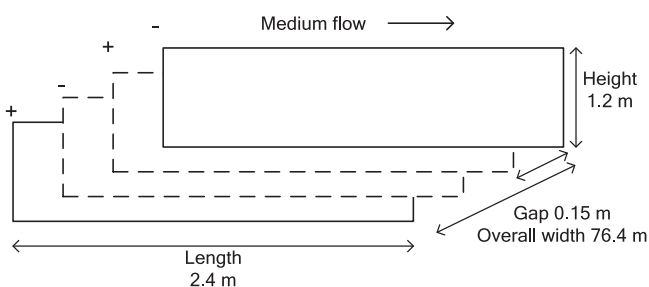


Fig. 4. Sketch showing dimensions of the proposed electroflocculator.

Based on the experimental results on the electrode separation of 0.15 m, the number of channels required is:

$$75.937 \text{ m} / 0.15 \text{ m} = 506.2 \text{ or } 506 \text{ channels}$$

Therefore, the number of electrodes is 507.

The overall width of the combined electroflocculation module equals to total width of channel plus the total thickness of electrode i.e.

$$506 \times 0.15 \text{ m} + 507 \times 0.001 \text{ m} = 76.4 \text{ m}$$

The overall dimensions for the electroflocculation module will therefore be 2.4 m × 76.8 m × 1.2 m as shown in Fig. 4:

Such module has a volume of 221.2 m³ with a net footprint area of 184.3 m².

The amount of aluminium required is 507 × 7.8 kg = 3.954 tonnes. The spot price for remade aluminium is \$1992 per tonne [35]; therefore, total cost of aluminium for the electrodes is approximately AU \$7876.

For the ease of construction, the electroflocculation module could consist of 9 above ground tanks each of 9 m × 2.4 m × 1.2 m. Each tank costs approximately \$660,000 [41]. Nine of such tanks with aluminium electrode will require a total cost of approximately \$6,100,000. Above ground tanks are preferred for the ease of inspection, as any unknown leaks may cause safety problems with stray-electricity.

3.3.2. Mixing modules

The required mixing time in the baffled hydraulic mixer is 15 min; therefore, the required total volume of the baffled hydraulic mixers is:

$$7.29 \text{ m}^3 \text{ s}^{-1} \times 15 \text{ min} \times 60 \text{ s} = 6561 \text{ m}^3.$$

Since the minimum flow velocity to prevent floc settling is 0.075 m s⁻¹, the length of the baffled hydraulic mixer with a residence time of 15 min is:

$$0.075 \text{ m s}^{-1} \times 15 \text{ min} \times 60 \text{ s} = 67.5 \text{ m}$$

A practical limit of the channel width to depth ratio is 1.0 and the average overlap ratio is 4–5 [22], also, using a typical mixing tank depth of 2 m for the ease of maintenance; therefore, the channel width is 2 m and the tank width is 8.44 m. The total number of channels is 8. The schematic of one baffled hydraulic mixers is shown in Fig. 5. As the dimensions are different from those available from literature [23,42], therefore, to ensure the flow remains turbulent and offers good mixing, the Reynold's number, *Re*, is calculated from Eq (8):

$$Re = \frac{\rho d_h v}{\mu} \quad (8)$$

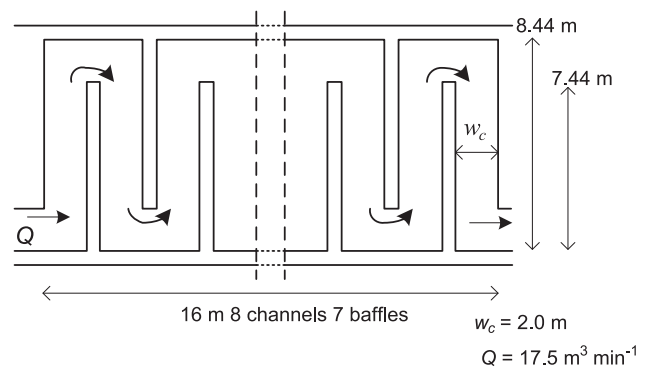


Fig. 5. Schematic presentation of the proposed baffled hydraulic mixer.

Table 5

Cost comparison for different separation processes.

Separation processes	Reference	Year	Cost (\$kg ⁻¹ biomass)		Items included in the processing cost
			Original ^c	2012	
Centrifugation – self cleaning plate	[3]	1995	1.71	2.58	Plant depreciation, maintenance and energy
	[7]	1988	0.86 ^a	1.68	Equipment depreciation, maintenance and energy
Flocculation – sedimentation	[7]	1988	0.37 ^a	0.72	Plant depreciation, maintenance, flocculant and energy
	[4]	1996	1.25	1.83	Plant depreciation, maintenance, flocculant and energy
Flocculation – flotation	[3]	1995	1.39	2.10	Plant depreciation, maintenance, flocculant and energy
	[7]	1988	0.91 ^a	1.78	Plant depreciation, maintenance, flocculant and energy
Electro-flocculation	[9]	1997	0.22	0.31	Energy only
Microbial flocculation	[21]	2010	0.29 ^{b,d}	0.31	Plant depreciation, maintenance, raw materials and energy
Electro-flocculation		2012		0.19	Plant depreciation, electrode dissolution and energy

^a US \$1 = DM 1.85 in 1988.^b US \$1 = A \$1.1 in 2010.^c US consumer price index in 1988, 1995, 1996, 1997, 2009 and 2012 are 115.7, 150.3, 154.4, 159.1, 211.1 and 226.6 respectively.^d Typical biomass concentration is assumed to be 0.5 kg m⁻³.

Where: ρ is the density of media, 1070 kg m⁻³ (salinity of 1 M); μ is the viscosity of media, 1.31×10^{-3} Pa s [43]; v is the linear velocity, 0.075 m; d_h is the hydraulic diameter. The hydraulic diameter d_h can be calculated from the hydraulic radius, R_h , from Eq (9) where:

$$d_h = 4 R_h \quad (9)$$

For open channels, the hydraulic radius, R_h , is calculated from Eq (10) [44].

$$R_h = \frac{dW_c}{2d + W_c} \quad (10)$$

Where: W_c is the channel width of 2 m; d is the channel depth of 2 m, therefore R_h is 0.666 m and d_h is 2.667 m.

$$Re = \frac{1070 \times 2.667 \times 0.075}{1.31 \times 10^{-3}} = 1.63 \times 10^5 \quad (11)$$

As $Re \gg 4000$, the flow is turbulent for proper mixing.

The volume of a single baffled hydraulic mixer is:

$$16 \text{ m} \times 8.44 \text{ m} \times 2 \text{ m} = 270.08 \text{ m}^3$$

As the required total volume of the baffled hydraulic mixers is 6561 m³, then the number of baffled hydraulic mixers is:

$$6561 \text{ m}^3 / 270.08 \text{ m}^3 = 24.3 \text{ tanks}$$

Therefore, 25 tanks are required with a footprint area of

$$25 \times 16 \text{ m} \times 8.44 \text{ m} = 3240 \text{ m}^2$$

The cost of the tanks is estimated by in-ground tanks of similar sizes at a construction cost of \$850 m⁻² surface area [41] and the cost of one baffled hydraulic mixer (excluding baffles) is:

$$\$850 \text{ m}^{-2} \times 18 \text{ m} \times 8 \text{ m} = \$122,400$$

For 25 tanks, the total cost is:

$$\$122,400 \times 25 = \$3,060,000$$

3.3.3. Settler

The required residence time for this settling tank is 30 min; therefore, the volume of the settling tank is:

$$7.29 \text{ m}^3 \text{ s}^{-1} \times 30 \text{ min} \times 60 \text{ s} = 13,122 \text{ m}^3$$

The construction cost, C , of a gravity settler can be estimated by [45]:

$$C = 2630 \times A^{0.678} \quad (12)$$

where A is the area of the clarifier/settler in m²; C is the cost in € in the year 1998. This cost estimation is valid for settler with surface areas from 175 m² to 1250 m².

For a typical circular settler of average depth 1.2 m, the surface area is:

$$13,122 \text{ m}^3 / 1.2 \text{ m} = 10,935 \text{ m}^2.$$

This area is bigger than the typical size range specified by Eq. (11); therefore, 9 smaller settlers, each with a surface area of 1215 m² (total footprint area of 10,935 m²) could be used.

The construction cost, C , for the 9 settlers is:

$$= 9 \times 2630 \times (1215)^{0.678} = \text{€}2,921,063$$

The accumulative Consumer Price Index in Australia is 120 in 1998 and 172.2 in 2012, the exchange rate is \$1 to €0.77; therefore, total cost for the settlers in 2012 is:

$$2,921,063 \times (172.2/120) \times (\$/\text{€}0.77)$$

Or approximately \$5,500,000.

The total capital cost for electroflocculators, baffled hydraulic mixers and gravity settlers is therefore:

$$\$6.1 \text{ m} + \$3.06 \text{ m} + \$5.5 \text{ m} = \$14.4 \text{ m}$$

The net footprint area for all the harvesting modules is

$$184.3 \text{ m}^2 + 3240 \text{ m}^2 + 10935 \text{ m}^2 = 14359 \text{ m}^2 \text{ (net)}$$

This area is approximately 1.4% of the total algal pond area.

3.4. Economics

3.4.1. Operating cost

By inserting laboratory electroflocculation results of a voltage of 5.3 V, a current of 9.91, aluminium dissolution efficiency of 1.5 and a flow rate of 0.576 m³ h⁻¹ into Eq. (7), the operating cost (electrical and aluminium dissolution) C_{op} in \$m⁻³ is:

$$C_{op} = 6.687 \times 10^{-4} (0.3738 \times 5.3 + 1.5) \frac{9.91}{0.576} = \$0.040 \text{ m}^{-3}$$

For a harvesting rate of $6.3 \times 10^5 \text{ m}^3 \text{ d}^{-3}$, the operating cost will be \$25231 d⁻¹ or \$9.21 m per annum; of which \$5.24 m is due to the cost of energy and \$3.97 m is due to the cost of aluminium dissolution. If electroflocculation/flotation is chosen as the harvesting method, then the energy cost will be approximately \$5.24 m × 1.7 or \$8.90 m.

3.4.2. Harvesting cost by electroflocculation

Allowing for a depreciation rate of 10% per annum, the depreciation cost is \$1.44 m per annum. The annual raw material, energy and capital depreciation cost will therefore be:

$$\$9.21 \text{ m} + \$1.44 \text{ m} = \$10.65 \text{ m}.$$

For a production rate of 5.748×10^7 kg per annum, the average harvesting cost for the algal dry biomass will be $\$0.19 \text{ kg}^{-1}$. The breakdown of the operation cost is: electrical energy at 49.2%; aluminium dissolution at 37.3% and plant depreciation at 13.5%. Comparing Process A and B, it can be seen that the extra capital cost incurred for mixers and settlers can be compensated by the lower energy requirement and this extra cost can be recovered within a period of 4 years.

3.4.3. Comparison with literature data

Table 2 shows that among the tests conducted by various authors, those performed by Vandamme et al. [31] have the lowest electroflocculation energy consumption of 1.08 MJ m^{-3} . By incorporating mixing and settling into the electroflocculation process, a lower energy consumption of 0.328 MJ m^{-3} has been achieved. This lowering of electrical energy used can reduce the amount of aluminium dissolution and hence the overall cost. A comparison of the harvesting costs (exclude drying) is presented in Table 5.

Table 5 shows that electroflocculation has the potential to be an economical harvesting technique; however, the estimated cost of $\$0.19 \text{ kg}^{-1}$ is still too high for biofuels as other costs such as cell disruption, extraction and refining need to be considered. For this reason, the co-production of other high valued microalgal products such as proteins or colour pigments may improve the overall production economy.

4. Conclusion

By incorporating mixing and settling into the harvesting process, electroflocculation has the potential to be a low cost microalgal harvesting technique when compare with other microalgal harvesting techniques. The mixing and settling can be made energy efficient by the use of a baffled hydraulic mixer and gravity clarifier. The processing cost, including electrical energy, aluminium dissolution and capital depreciation cost is estimated to be $\$0.19$ per kg of the microalgal biomass and may have the potential to be further optimised.

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