Process optimization of microalgae cultivation in a bubble-column photobioreactor

Y Muharam\textsuperscript{1,2}, Dianursanti\textsuperscript{1} and A S Wirya\textsuperscript{1}

\textsuperscript{1} Department of Chemical Engineering, University of Indonesia, Depok 16424, Indonesia
\textsuperscript{2} E-mail: muharam@che.ui.ac.id

Abstract. A process optimization was performed in this study to obtain the values of the geometry and process parameters that provide an optimum average microalgae concentration in an internally illuminated bubble-column photobioreactor. A phenomenological model of the photobioreactor is used to simulate the process in the reactor. The model considers gas-phase mass balance and liquid-phase mass balances accompanied by the rates of CO\textsubscript{2} and nutrient intakes in liquid phase. The process optimization gives the values of the small and large cylinder diameters, the culture height, the inlet air flowrate, the CO\textsubscript{2} mole fraction in the inlet air and the initial microalgae concentration that promotes the microalgae growth to 0.773 g/L from 0.2 g/L for less than four days.

1. Introduction
As one of the third-generation renewable energy sources that does not compete with the needs of other industries, microalgae can be a solution in sustainable production. Oil contained in microalgae is amazing, reaching 40-85\% of the dry weight [1]. That is much higher when compared to oil palm, which is only about 20\%. Microalgae can be processed into several derivates of bioenergy products such as biodiesel, bioethanol, biobutanol, straight vegetable oil and green fuels.

Mass production of microalgae will be efficient if reactors used for microalga cultivation are designed as good as possible. This is achieved when the design considers process and geometry parameters. A design method considering those parameters is simulation using theoretical or phenomenological models. By simulation, the process is optimized so that the best performance of the reactor is reached.

Several studies on the modelling of microalgae cultivation reactors have been reported. Seo et al. modeled a bubble-column photobioreactor with computational fluid dynamics to learn mixing in the reactor [2]. The microalgae growth rate was not introduced into the model. Pegallapati and Nirmalakhandan developed the model of an internally illuminated bubble-column photobioreactor that has been validated with the experimental data for 16-day cultivation [3]. They assumed that gas phase and bulk culture medium were perfectly mixed. Sforza et al. modeled a flat-plate photobioreactor that assumed as a perfectly stirred-tank reactor without recycle [4]. The material balance was modeled at macroscopic level. Modeling of an airlift-raceway photobioreactor has been performed by Kethesan and Nirmalakhandan [5]. The model is macroscopic level and includes CO\textsubscript{2} transfer, biomass growth, light effect and nitrogen consumption. Nauha and Alopaeus combined computational fluid dynamics
and compartmental models to predict mixing, mass transfer characteristics and light effect in a photobioreactor [6]. The models have not been validated with experimental data. Zhang et al. modeled a flat-plate photobioreactor describing algae growth, the consumptions of CO$_2$, phosphate, nitrate, and ammonium by *Chlorella kessleri* cultured in waste water for 21 days [7]. Recently, Muharam et al. developed the axisymmetric phenomenological model of an internally illuminated bubble-column photobioreactor for microalgal cultivation [8]. The model considers mixing in gas phase and bulk culture medium using dispersion approach and has been validated using the experimental data by Pegallapati and Nirmalakhandan [3].

The objective of the present study is to obtain an optimum process for *Nannochloropsis salina* microalga cultivation in an internally illuminated bubble-column photobioreactor through simulation. The internal illumination is chosen because it provides more homogeneous intensity to the angular direction of the reactor. The axisymmetric phenomenological model that has been developed by Muharam et al. [8] was used.

2. Methodology

As described in [8], the internally illuminated bubble-column photobioreactor model comprises a gas-phase mass balance and liquid-phase mass balances. The uptakes of CO$_2$, phosphate, nitrate and sulphate take place in liquid phase. The photobioreactor is illustrated in Figure 1. Two cylinders are arranged concentrically. The microalgae culture is placed in the annular space. A light with constant intensity is situated inside the small cylinder. The sparger holes to spray CO$_2$-containing air locate at the bottom of the cylinder.

Process optimization is to obtain the values of decision variables in terms of geometry and process parameters that provide the optimum value of an objective function, which in this case is the average microalgal concentration in the photobioreactor. Therefore, the culture height as well as the small and large cylinder diameters as the geometry parameters vary. The inlet air flowrate and pressure, as well as the CO$_2$ mole fraction in the inlet air and the initial microalgae concentration as the process parameters vary. Nutrient concentrations do not vary.

![Figure 1. Model of internally illuminated bubble column photobioreactor.](image)

The gas-phase CO$_2$ mass balance is as follows:

$$\frac{\partial C_{CO_2,G}}{\partial t} = \left( \frac{1}{r} \frac{\partial}{\partial r} \left( r D_{r,G} \frac{\partial C_{CO_2,G}}{\partial r} \right) \right) + \frac{\partial}{\partial z} \left( D_{z,G} \frac{\partial C_{CO_2,G}}{\partial z} \right)$$

$$+ \left( u_r \frac{\partial C_{CO_2,G}}{\partial r} + u_z \frac{\partial C_{CO_2,G}}{\partial z} \right) = -R_{CO_2,GL}$$

(1)

where $C_{CO_2,G}$ is the gas-phase CO$_2$ concentration, $u_r$ is the $r$-component velocity, and $u_z$ is the $z$-component velocity.

The mass balances of CO$_2$, microalgae and nutrients in liquid phase are described by:
\[
\frac{\partial C_{j,L}}{\partial t} - \left( \frac{1}{r} \frac{\partial}{\partial r} \left( r D_{r,L} \frac{\partial C_{j,L}}{\partial r} \right) + \frac{\partial}{\partial z} \left( D_{z,L} \frac{\partial C_{j,L}}{\partial z} \right) \right) = R_{j,L}
\]  

(2)

where \(C_{j,L}\) is the liquid-phase \(j\)-component concentration, and \(R_{j,L}\) is the \(j\)-component reaction rates.

The gas-phase axial dispersion coefficient, \(D_{z,G}\) and the gas-phase radial dispersion coefficient, \(D_{r,G}\) are in accordance with Joshi [9]:

\[
D_{z,G} = 50D_{G}^{0.35}\left( \frac{u}{e_{G}} \right)^{2}
\]  

(3)

and

\[
D_{r,G} = D_{z,L}
\]  

(4)

The liquid-phase axial dispersion coefficient, \(D_{z,L}\) and the liquid-phase radial dispersion coefficient, \(D_{r,L}\) are in accordance with Krishna et al. [10]:

\[
D_{z,L} = 0.0632 \bar{D}_{G}^{0.5} D_{R}^{1.5} \left( \frac{u^{3}}{g_{water}} \right)^{0.125}
\]  

(5)

and

\[
D_{r,L} = 0.01D_{r,L}
\]  

(6)

The gas-liquid mass transfer, \(R_{CO_{2,G}}\) is formulated by

\[
R_{CO_{2,G}} = k_{L}a_{CO_{2}}(C^{*}_{CO_{2}} - C_{CO_{2,L}})
\]  

(7)

where \(k_{L}a_{CO_{2}}\) is correlated to the liquid-phase diffusion coefficients of \(CO_{2}\) and \(O_{2}\), \(d_{CO_{2}}\) and \(d_{O_{2}}\) by

\[
k_{L}a_{CO_{2}} = k_{L}a_{O_{2}} \sqrt{\frac{d_{CO_{2}}}{d_{O_{2}}}}
\]  

(8)

as reported by Grima et al. [11].

The \(CO_{2}\) concentration in liquid phase that is in equilibrium with the one in gas phase, \(C^{*}_{CO_{2}}\) is related to the \(CO_{2}\) mole fraction, \(y_{CO_{2}}\) in gas phase by

\[
C^{*}_{CO_{2}} = \frac{py_{CO_{2}}}{RH_{CO_{2}}}
\]  

(9)

where \(H_{CO_{2}}\) the Henry’s dimensionless constant of \(CO_{2}\) in water.

The microalgae growth rate is expressed by

\[
R_{algae} = \mu C_{algae} = k_{d}C_{algae}
\]  

(10)

where \(\mu\) is the specific growth rate of microalgae as the function of \(CO_{2}\) and nutrient intakes, as well as temperature and light intensity, and \(k_{d}\) is the death rate of microalgae.
3. Results and discussion
The simulation results are shown in figure 2 to figure 7. The simulation was carried out with the small cylinder diameter of 5 cm, the large cylinder diameter of 30 cm, the culture height of 60 cm, the inlet air flowrate of 1 L/min, the inlet air pressure of 1 atm, the inlet CO₂ mole fraction of 0.02, the initial microalgal concentration of 0.2 g/L, and the inlet and ambient temperatures of 30 °C. When a process or geometry parameter varies, all other parameters do not change.

Figure 2 shows the relationship between the average microalgae concentration in the reactor and the large cylinder diameter. From the figure, it can be seen that the increase in the large cylinder diameter leads to the decrease in the average microalgae concentration, but the change is not significant. This indicates that the agitation by rising gas bubbles provides good mixing in the axial and radial direction of the reactor. The effect is significant if the gas flowrate is very low, which is within the regime of less than 0.5 L/min. In this regime, the increase in the large cylinder diameter decreases significantly the average microalgae concentration. This is because the light penetration inside the reactor is obstructed by microalgae. The culture that is away from the light source receives low-intensity light.

The microalgae concentration increases almost linearly during the first few days, then tends to be constant after reaching a maximum value. In this steady state, the growth rate and the death rate from overcrowding are comparable. The time needed to achieve the steady state depends on the large cylinder diameter. It takes longer time for the large cylinder with larger diameter to reach the maximum value.

Figure 3 exhibits the effect of the small cylinder diameter on the average microalgae concentration. The simulation was performed with the same values of the parameters set above, except that the large cylinder diameter was set to 50 cm. As shown in the figure, the highest maximum concentration occurs in the reactor with the small cylinder diameter being 5 cm.

The change in the culture height does not affect significantly on the average microalgae concentration as shown in figure 4. This happens when the bubble velocity can be preserved along the reactor. Indeed, the model does not consider the change in the bubble velocity when bubbles rise because the culture height is not very high. In the case of very high culture height, the bubble velocity decreases as a result of part of bubble mass moves into liquid phase along its way to axial direction.

Figure 5 depicts the microalgae growth as the function of the initial concentration. The initial concentration of 0.05 g/L results in the steady-state concentration of 0.68 kg/m³ or 10-fold increase. The initial concentration of 0.2 kg/m³ promotes the concentration to 0.75 kg/m³ at steady state or fourfold increase. With constraints such as self-shielding factor, the ratio of the maximum concentration to the initial concentration for higher initial concentration is smaller than for the lower

---

**Figure 2.** The average microalgae concentration vs the large cylinder diameter.

**Figure 3.** The average microalgae concentration vs the small cylinder diameter.
one. The ratio for higher initial concentration are certainly smaller, but their steady state is achieved faster than lower one, and more microalgae are yielded.

**Figure 4.** The average microalgae concentration vs the culture height.

**Figure 5.** The average microalgae concentration vs the initial microalgae concentration.

Figure 6 shows the effect of the inlet air flowrate on the average microalgae concentration in the reactor. The inlet flowrate varies within the range that complies the superficial velocity requirement for homogeneous mixing in bubble-column reactors. It can be seen from the figure that the change in the inlet flowrate does not significantly affect the average microalgae concentration. For low flowrate regime, the effect is clear, as happened for 0.01 L/min.

Figure 7 displays the effect of the CO$_2$ mole fraction in the inlet air. The increase in the CO$_2$ mole fraction leads to the increase in the CO$_2$ intake by microalgae. Thus, the microalgae concentration goes up. However, due to the limitation of CO$_2$ solubility in water, the maximum CO$_2$ mole fraction in the inlet flow is 0.05.

**Figure 6.** The average microalgae concentration vs the inlet flowrate.

**Figure 7.** The average microalgae concentration vs the CO$_2$ mole fraction.

From the above simulation results, the values of the geometry and process parameters giving the optimum microalgae growth are listed in table 1. These values provide the maximum average microalgae concentration of 0.773 g/L achieved in less than four days as shown in figure 8. This information is important to design an internally illuminated bubble-column photobioreactor with a desired production capacity. For high capacity, the culture height may be increased to several meters. In order to anticipate the reduction of CO$_2$ content in rising bubbles, some spargers may be installed at certain elevations. If one reactor column is not sufficient, identical reactor columns may be arranged in parallel and treated equally to meet the desired capacity.
Table 1. Optimization result of the bubble-column photobioreactor.

| Parameters                                    | Value |
|-----------------------------------------------|-------|
| Large cylinder diameter                       | 30 cm |
| Small cylinder diameter                       | 5 cm  |
| Inlet air flowrate                            | 1 L/min|
| CO\textsubscript{2} mole fraction in inlet air| 0.05  |
| Initial microalgae concentration              | 0.2 g/L|
| Inlet and ambient temperatures                | 30 °C |

Figure 8. The optimum average microalgae concentration.

4. Conclusion
The process optimization gives the values of the geometry and process parameters as listed in Table 1 with the optimum average microalgae concentration being 0.773 g/L achieved in less than four days.

5. Acknowledgement
We express our gratitude to the University of Indonesia, which funded this research through the scheme of Penelitian Unggul Perguruan Tinggi No 2769/UN2.R3.1/HKP.05.00/2017.

6. References
[1] Borowitzka M 1996 Algae as food Microbiology of Fermented Foods ed B J B Wood (London: Blackie Academic & Professional) pp 585-602
[2] Seo I, Lee I, Huang H, Hong S, Bitog J, Kwon K, Lee C, Kim Z, Cuello J 2012 Biosyst. Eng. 113 (13) 229
[3] Pegallapati A K, Nirmalakhandan N 2012 Bioresource Technol. 124 137
[4] Sforza E, Enzo M, Bertucco A 2014 Chem. Eng. Res. Des. 92(6) 1153
[5] Ketheesan B, Nirmalakhandan N 2013 Bioresour. Technol. 136 689
[6] Nauha E K, Alopaeus V 2013 Chem. Eng. J. 229 550
[7] Zhang X, Gong X, Chen Y 2015 J. Ind. Microbiol. Biotechnol. 23(1) 691
[8] Muharam Y, Dianursanti, Pramadana A B, Wirya A S 2017 Chem. Engineer. Trans. 56 1555
[9] Joshi J B 1980 Trans. Inst. Chem. Engs. 58 228
[10] Krishna R, Urseanu R, van Baten J M, Ellnenberger J 1999 Chem. Eng. Sci. 54 4903
[11] Grima E M, Perez J A S, Camacho F I G, Medina A R 1993 J. Chem. Technol. Biotechnol. 56(4) 329