Evaluation of Heat Distribution And Aeration On Xylanase Production From Oil Palm Empty Fruit Bunches Using Tray Bioreactor

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Abstract. Indonesia imports ±99% of its industrial enzyme supply, mostly from China, India, Japan, and Europe. Xylanase is one of the most widely used enzymes in industries, mainly pulp and paper, animal feed, and bakery. Previous research has shown that Aspergillus fumigatus ITBCCCL170 can produce xylanase enzyme using the Solid State Fermentation (SSF) process in a tray bioreactor using OPEFB. The tray bioreactor model for solid-state fermentation was developed and further used to simulate the effects of aeration and heat distribution on biomass growth and enzyme production. The results showed that solid-state fermentation in a tray bioreactor is indeed far from ideal as the model simulated. Yet, parallel aeration can increase cell and enzyme production, which needed further validation.

1. Introduction
The essential parameters in the SSF system are the distribution of heat and mass transfer [1,2]. Several things influence the heat and air distribution in the tray bioreactor, namely, the water content in the bed (Wb) and the bioreactor (Wi). Furthermore, the bioreactor's oxygen content (Oks) influenced the number of cells that grow (X). The heat of metabolism is generated by microorganisms and the heat of evaporation that occurs during incubation. Therefore we need a model that can describe the changes in the tray bioreactor system.

Microorganisms release some metabolism heat (Qmet) along with their growth. This metabolic heat can increase the bed temperature (Tb) and further reduce water in the bed. Both may harm the growth of microorganisms, which can only occur at a limited range of temperature and above a minimum level of humidity. To overcome this, the air supplied into the system (the aeration) can lower bed temperature. The airflow pattern inside the bioreactor contributes to the heat distribution as it may create a proper environment for the microbes to grow. However, the introduction of air into the system should not cause the evaporation of free water in the bed. The effect of airflow, humidity, and mixing on the growth of microorganisms in the two-tray bioreactor modelling reported that the use of saturated air of 0.1 LPM of air flow rate and mixing once a day gave the highest xylanase activity of 236.3 U/g [3].

Rajagopalan and Modak [4] introduced simultaneous modelling of heat and mass balance in the growth of Aspergillus niger in a tray bioreactor. The pan's height influenced the profile of oxygen
concentration in the tray bioreactor [5]. Montero [6] has reported the modelling of forced aeration on a tray bioreactor. However, since the model build was for a single tray, he could not simulate the effect of aeration distribution. Another model for the tray bioreactor is the growth profile in the aeration variation of Aspergillus oryzae and Aspergillus awamori in a circular tray bioreactor [7].

This research studied the effects of aeration and heat distribution on cell and enzyme production during a solid-state fermentation for xylanase production using Aspergillus fumigatus on Oil Palm Empty Fruit Bunches (OPEFB) as the primary substrate.

2. Materials and Method

The model was built as an ideal system due to the airflow pattern. The specific growth rate as a function of bed temperature was fitted empirically to approximate the experimental cells' values and enzymes. It was assumed that the system was homogenous, and the bed was initially saturated with Oxygen. The water vapour transferred into the bioreactor headspace was assumed to occur only from the surface's thin layer. The mixture of OPEFB and water in the bed was considered a pseudohomogenous system and will only be represented by OPEFB. The system was set as an isothermal system because the bioreactor chamber temperature was controlled at 37°C. The saturated air was supplied from the bottom of the bed at a constant flow rate of 0.006 m³/h and 28°C. As for the bioreactor, it is assumed that the distance between the tray does not affect heat transfer.

The model is divided into three parts: the basic model in which the heat and aeration are assumed to be homogeneously mixed inside the bioreactor (one chamber bioreactor). In this model, the influence of the presence and absence of air is simulated. The model was built as combinations of smaller homogeneously mixed chambers in series or parallel: the two-chamber and four-chamber bioreactors.

2.1. Building a model

The model built consists of a mass balance and a heat balance to obtain a complete picture of the process. Water mass balance is in the form of water content in the bed due to the water transfer from the bed to the bioreactor. While the bioreactor water content was due to mass transfer from the saturated air into the bioreactor. The heat balance is in the form of sensible heat generation, metabolic heat, and the release of evaporative heat from the bed to the bioreactor. Besides, the growth rate of microorganisms and product formation is the main focus of modelling.

2.2. Basic model

The one-chamber model means placing 1000 grams of EFB in one pan with a dimension of 26x26x7 cm, and the bioreactor consists of one tray in a 0.12 m³ bioreactor room. Figure 1. shows the mass and heat transfer process in a one-chamber bioreactor.
Figure 1 shows that the inlet air will absorb some heat generated in the bed ($Q_{\text{met}}$) and evaporation heat ($Q_{\text{evap}}$). The incubation took place at a constant $T_{\text{opt}}$ with $R\text{H}_0$ 85%. Some metabolic water ($W_{\text{met}}$) is produced, and some water in bed evaporated ($W_{\text{evap}}$), which moves into the flowing air. Because there is a temperature gradient between the bed and the bioreactor, a significant amount of sensible heat is released ($Q_{\text{sen}}, t$ and $Q_{\text{sen,b}}$). Changes in bed temperature in subsystem two (2) will affect subsystem one (1).

2.2.1. Growth kinetic
Several model equations describe the growth of microorganisms in solid-phase bioreactors, as presented in Table 1. [8]

| Linear      | $r_X = \frac{dX}{dt} = K$ (1) |
|-------------|-------------------------------|
| Exponential | $r_X = \frac{dX}{dt} = \mu X$ (2) |
| Logistic    | $r_X = \frac{dX}{dt} = \mu X \left(1 - \frac{X}{X_{\text{max}}}\right)$ (3) |

The model used an exponential growth kinetic equation, a first-order kinetic equation, with $\mu$ value was empirically constructed following polynomial regression as a function of bed temperature in a range Aspergillus fumigatus growth temperature. The specific growth rate was used as a constraint in the iteration of the model's equations simultaneously.

2.2.2. Product formation kinetic
Similar difficulties in modelling products' formation, such as the rate of nutrient utilization, make the model equations more complex [9]. The empirical equations can present a proportional relationship between cell mass and product [10]. This study assumed the $m_p$ value to 0, which had been reported before [11].

$$r_p = \frac{dP}{dt} = Y_{p/X} \frac{dX}{dt} + m_p X$$ (4)

$$r_{Ox} = O\text{UR} = Y_{O/X} r_X + m_O X$$ (7)

2.2.3. Oxygen balance
The Oxygen balance in the tray bioreactor describes the change in the mole fraction of Oxygen at any time due to the intake of airflow, the outflow of air, and microorganisms' Oxygen consumption. The equation which states the relationship between the flow of Oxygen is,

$$\frac{dO_{Ks}}{dt} = F_g y_{O_{2,in}} - F_g y_{O_{2,out}} + O\text{UR}$$ (6)

Whereas the rate of oxygen consumption (OUR) correlates with the growth of microorganisms [9].

$$r_{Ox} = O\text{UR} = Y_{O/X} r_X + m_O X$$ (7)

2.2.4. Heat balance in the bioreactor
Heat transfer in the tray bioreactor occurs in a thin layer of the bed surface due to airflow at the top and bottom of the bed. Besides, there is a certain amount of heat generated from the metabolism of
microorganisms. The heat transferred to the top of the bed is evaporation heat, which causes some of the bed water to evaporate. As microbes grow, a temperature gradient has formed accordingly. This temperature gradient produces sensible heat transferred from the bed into the air at the top of the bed. In general, the equation is written,

$$m_b C_p b \frac{dT_b}{dt} = Q_{met} - Q_{sen} - Q_{evap}$$   \hspace{1cm} (8)$$

The metabolic heat produced because of microorganism growth was defined and written as,

$$r_Q = Q_{met} = Y_{Q/O} r_X + m_o X$$   \hspace{1cm} (9)$$

$$Q_{met} = \left( \frac{460.000}{Y_{Q/O}} \right) r_X$$   \hspace{1cm} (10)$$

The formation of metabolic heat can increase the bed temperature; in the tempeh production process, it has been reported that an increase in temperature reaches 3°C/cm with a bed height of 5.1 cm [12]. Even the centre's bed temperature can reach 47°C to 49.6°C under saturated air entering at a rate of 0.47 Lpm. Thus the metabolic heat has a significant impact on the increase in bed temperature. One way to reduce the bed's temperature is to provide saturated air with relative humidity above 95%. If the saturated air temperature is lower than the bed's temperature, there is a sensible heat transfer. The sensible heat transfer is expressed in the equation;

$$Q_{sen} = -U_t A_b (T_b - T_{in}) - U_b A_b (T_b - T_{in})$$   \hspace{1cm} (11)$$

Because the conductivity of aluminium (pan material) is much greater than that of the bed, then $U_t = U_b$ so that

$$Q_{sen} = -2 U_t A_b (T_b - T_{in})$$   \hspace{1cm} (12)$$

As long as the bed temperature is higher than the bioreactor temperature, the bed's cooling by the release of sensible heat is still quite significant. However, if the bed temperature is the same as the bioreactor temperature, the sensible heat release is no longer effective. Thus, cooling the bed needs to be assisted by introducing saturated air so that the bed's temperature reduction is more effective [9].

In the evaporation process, a diffusion process requires a mass transfer coefficient in the model's equation. Because this study was lack of experiment data, the model uses a correlation equation for the mass transfer coefficient [13]. The humidity gradient between the bed and air becomes the driving force of the evaporation process so that the heat equation for evaporation becomes,

$$Q_{evap} = \lambda k_m A_b (\mathcal{H}_b - \mathcal{H}_{in})$$   \hspace{1cm} (13)$$

2.2.5. Water balance
The system needs humidity to dissipate heat metabolites because water's heat capacity is greater than air, making it easier to absorb heat. If the heat can be controlled, water evaporation from the bed can be prevented to maintain productivity. The water can be free water available outside of microorganisms, metabolized water evaporated into the macro environment, and water becomes a nutrient solvent. The water balance is divided into two: the water balance in the bed and the bioreactor.

Because the bed is a closed system, it only involves changes that occur in the bed, namely the formation of water metabolites and loss of evaporated water.
\[ r_{W_b} = r_{W_{met}} r_{W_{evap}} \]  

(14)

Metabolite water is water produced when microorganisms grow, so the water metabolite equation is analogous to the enzyme production equation, namely,

\[ r_{W_{met}} = Y_{W/X} r_X \]  

(15)

Evaporation water is water that is evaporated from the bed because there is a humidity gradient or water activity between the bed and the bioreactor [9]. The rate of evaporation was shown in equation (16) and needed adjustments so that each component of the equation has the same unit,

\[ r_{W_{evap}} = k_m A_b (\mathcal{H}_b - \mathcal{H}_{in}) \]  

(16)

There were some adjustments made on the model to maintain each term had the same unit, as \( k_m \) was in m/h while \( r_{W_{evap}} \) was in g water/g OPEFB-h.

2.3. Simulation

The model simulation was performed using MATLAB and ode45 solver according to the model that has been built, one-chamber bioreactor, serial airflow in two-chamber, and four-chamber bioreactor and parallel airflow in the four-chamber bioreactor. The coefficients used are tabulated in Table 2.

The experiments that have been performed by using 1 kg of OPEFB in 2 trays [3] compared to the model built.

| SYMBOL | NAME | VALUE | UNIT (SOURCE) |
|--------|------|-------|---------------|
| \( A_i \) | Cross-sectional area of bioreactor | 0.2 | m² (calculated) |
| \( A_b \) | Bed surface area | 0.2704 | m² (calculated) |
| \( C_{p,b} \) | Specific heat capacity of OPEFB | 1.4827 | J/g-K [14] |
| \( E \) | Enzyme produced | Simulated | gram |
| \( F_{g} \) | The volumetric flow rate of air | 0.006 | m³/hr |
| \( \mathcal{H}_b \) | Bioreactor humidity | Simulated | g water/g dry air |
| \( \mathcal{H}_{in} \) | Air inlet humidity | Calculated | g water/g dry air |
| \( h_c \) | Convective heat transfer coefficient | 61.5 | W/m²-K [15] |
| \( k_m \) | Water mass transfer coefficient | Correlation | [13] |
| \( L \) | Pan length | 0.52 | m |
| \( m_b \) | OPEFB mass | 1000 | gram |
| \( O_k \) | O₂ mole fraction | Simulated | - |
| \( P_{sat} \) | The total pressure of the system | 1 | atm |
| \( p_{sat} \) | Saturated pressure | Calculated | atm |
| \( p_a \) | The partial pressure of water vapour | Calculated | atm |
| \( R \) | Ideal gas constant | 0.000082 | m³-atm/mol-K |
| \( T_b \) | Bed temperature | Simulated | K |
| \( T_{in} \) | Inlet air temperature | 301,15/310,15 | K |
| \( V_i \) | Bioreactor volume | 0.12 | m³ (calculated) |
| \( V_b \) | Bed volume | 0.0189 | m³ (calculated) |
| \( y_{O_2,in} \) | O₂ mole fraction of inlet air | Simulated | - |
| \( y_{O_2,out} \) | O₂ mole fraction of outlet air | Simulated | - |
| \( y_{air,in} \) | Water mole fraction of inlet air | Simulated | - |
| \( y_{air,out} \) | Water mole fraction of outlet air | Simulated | - |
| \( Y_{E/X} \) | Yield coefficient of enzyme produced per grown cell | 1.4-gram/gram cell | Data |
| \( Y_{Q/O} \) | Yield coefficient of heat produced per O₂ consumed | 460 | kJ/mol [16] |
Table 3. Coefficients and parameters used in the model (continued).

| SYMBOL | NAME                        | VALUE  | UNIT (SOURCE)               |
|--------|-----------------------------|--------|-----------------------------|
| Y<sub>O/X</sub> | Yield coefficient of consumed O<sub>2</sub> per grown cell | 0.7    | [17]                        |
| U<sub>T</sub>   | Global heat transfer coefficient | 4524.589 | J/jam-m<sup>2</sup>-K (calculated) |
| λ       | Heat of evaporation         | 2502   | J/g water [18]              |
| X       | Cell mass                   |        | Simulated gram              |
| μ       | The specific growth rate coefficient |        | Calculated 1/hr             |

3. Results

The simulation for studying the effect of aeration is presented in figure 2. The simulation was performed on one-chamber bioreactor model.

3.1. Effect of absence and presence of aeration on one-chamber bioreactor

Figure 2a showed that aeration caused a sudden decrease in the bed temperature (solid line), while a non-aerated process can maintain the bed temperature at ±310.15K for about 45 hours (dashed-line). The sudden decrease in bed temperature affected cell growth, as can be seen in figure 2b. As cell growth was decreased, the Oxygen was consumed at a lower rate, and it did not show any changes up to 120 hours of fermentation time. While the water balance showed that there were only slight changes in water in bed, and bioreactor humidity did not show any changes. These changes will not occur in the experiment because as microorganisms grow, there will be enough metabolic heat produced and increase bed temperature [12]. The expected decrease in bed temperature was too fast, presumably due to relatively low growth, while the air rate was too large. For this reason, changes in the rate of air are also simulated at a value of 0.003 m<sup>3</sup> / hour and 0.012 m<sup>3</sup> / hour. The results are presented in Table 3.

As shown in Table 3, the higher the aeration flow rate, the bed temperature decreases started from early simulation time and becomes lower than the optimal temperature (310.15K) at the end of simulation time. The difference between those three is the rate of the bed temperature increase. The model explains the effect of the airflow rate of air on changes in bed temperature. However, the decrease in bed temperature due to aeration appears extreme because it occurs as soon as the bed comes into contact with air.

The model simulation showed different results compared to real experimental data. This difference occurs because the air and heat were distributed homogeneously in the model. The real condition indicated that the bioreactor was not homogeneously mixed. Furthermore, the experiment can not validate the model because the bioreactor was not as ideal as the model, and the comparisons can be seen in Table 4.
Figure 2. The aeration effect on one-chamber bioreactor, solid line for no aeration and dashed line for aeration at 0.1 LM$^{-1}$.

### Table 4. Aeration flow rate effect on bed temperature.

| Airflow rate, m$^3$/jam | $T_b$ at 24 h (K) | $T_b$ at 100 h (K) | $T_b$ at 168 h (K) |
|------------------------|-----------------|-------------------|-------------------|
| 0                      | 310.2           | 310.9             | 315.9             |
| 0.003                  | 300.6           | 301               | 306               |
| 0.006                  | 299             | 300               | 305               |
| 0.012                  | 299             | 299               | 302               |

### Table 5. Comparison between experimental data and model.

| Aeration | Experiment | Model |
|----------|------------|-------|
|          | Cell, mg/g OPEFB | The enzyme mg/g OPEFB | Cell, mg/g OPEFB | The enzyme mg/g OPEFB |
| Without  | 0.495      | 0.632 | 0.755 | 1.054 |
| With     | 0.796      | 0.946 | 0.516 | 0.719 |

3.2. The effect of the division of the bioreactor chamber in serial airflow pattern
As has been reported, increasing the tray bioreactor's production capacity due to an increase in the bioreactor space in the form of the number of trays [19]. The simulation was tabulated in table 5.
From this simulation, it can be concluded temporarily that the serial flow pattern in two and four-chamber causes the system to be not homogeneous. It can be seen from the gap in the number of cell products and enzymes in each bioreactor chamber. The division of the bioreactor is identical to the division of the mass of the media. In contrast, the division of the flow pattern in series causes Oxygen's supply to the second chamber, third and fourth to decrease because it has been consumed. Therefore, with the same airflow rate of 0.006 m³ / hour, the four chambers oxygen intake is higher than the two-chamber. This condition does not cause a decrease the bioreactor's total productivity because the cell and enzyme production is accumulated. Following what was stated by Mitchell et al. (2006), increasing the chambers used will increase fermentation production in a tray bioreactor. Although each chamber's production is different, the total production of cells and enzymes increases with increasing bioreactor chambers.

**Table 6. Resume of simulation results.**

| Chamber | Cell, mg cell/g OPEFB | The enzyme, mg xylanase/g OPEFB |
|---------|-----------------------|---------------------------------|
|         | One-Chamber | Two-Chamber | Four-Chamber | One-Chamber | Two-Chamber | Four-Chamber |
| Chamber 1 | 0.52        | 0.85        | 1.90        | 0.98        | 1.60        | 3.60        |
| Chamber 2 | 0.03        | 0.03        | 0.02        | 0.05        | 0.03        |
| Chamber 3 | 0.02        | 0.02        | 0.02        | 0.02        |
| Chamber 4 | 0.02        | 0.02        | 0.02        |
| Total    | 0.44        | 0.50        | 0.98        | 0.82        | 0.92        |

**3.3. The effect of parallel airflow pattern**

Airflow patterns simulations were carried out by dividing the compressor's airflow, fixed at 0.006 m³ / hr, into four bioreactor chambers. The results obtained are presented in table 6.

**Table 7. Resume of simulation.**

| Chamber | Cell, mg/g OPEFB | The enzyme mg/g OPEFB |
|---------|------------------|----------------------|
|         | One-Chamber | 4-Paralel | 4-Serial | One-Chamber | 4-Paralel | 4-Serial |
| Chamber 1 | 0.52        | 2.43       | 1.90     | 0.98        | 4.60       | 3.60     |
| Chamber 2 | 2.43        | 0.03       | 4.60     | 0.03        |
| Chamber 3 | 2.43        | 0.02       | 4.0      | 0.02        |
| Chamber 4 | 2.43        | 0.02       | 4.60     | 0.03        |
| Total    | 0.52        | 2.43       | 0.50     | 0.98        | 4.60       | 0.92     |

Based on the tabulation results in Table 6, it can be seen that a parallel bioreactor, resulting in higher total cell and enzyme production when compared to the series flow. There is an increase of approximately five times the yield in the four-chambers bioreactor. The parallel bioreactor ensures the availability of fresh air supply so that cell growth is maximized and uniform even though the first chamber of the bed in the serial bioreactor produces more cells. The first chamber's serial bioreactor produces more cells and enzymes than the first chamber of a parallel bioreactor. This condition was due to the more intense heat transfer that occurs in the one-chamber bioreactor. Even though the total intake air flow rate is the same, the air's linear velocity in the bioreactor is different. The linear velocity of air in the one-chamber bioreactor is higher than in the four-chamber parallel bioreactor. This condition allows for better heat transfer to occur in parallel bioreactors. Heat transfer takes place more slowly so that the decrease in layer temperature is slighter and can even return to the optimal temperature for microbial growth. As a result, microbes' oxygen consumption occurs slower in a one-chamber bioreactor because the bed temperature is lower than the optimal temperature. The bed water in the one-chamber bioreactor appears to increase faster than in the parallel bioreactor. All of these changes result in better productivity of the parallel bioreactor than the one-chamber bioreactor.

This result follows Mitchell et al. [19], who stated that to increase production capacity using a tray bioreactor, namely by increasing the number of trays or trays. Thus it is important to apply parallel bioreactor arrangement so that homogeneous conditions are maintained, and the tray bioreactor can
increase productivity. The use of four lamps in the experiment was an attempt to distribute heat. Thus in the model, four lamps will be modelled as a variation of the four-chamber bioreactor. The results of the comparison are presented in Table 7.

Model simulation results and experimental results show a similar phenomenon; cell production increases when the process occurs in more chambers. However, there was an anomaly in the production of enzymes experimentally. But it was believed this was due to the inaccurate protein analysis process because it cannot distinguish between cells and fungal spores. The enzyme production simulated by the series bioreactor arrangement (0.92 mg / g OPEFB) was close to experimental results (0.78 mg / g OPEFB). However, in the experiment using four lamps, data was not obtained for each pan, so it was impossible to determine the system conditions. Therefore, model validation is needed to ensure the correctness of the model.

Table 8. Comparison of four-chamber in Experiment and Simulation.

| Chamber          | Experiment | Simulation |
|------------------|------------|------------|
|                  | Cell, mg/g OPEFB | The enzyme mg/g OPEFB | Cell, mg/g OPEFB | The enzyme mg/g OPEFB |
| One-Chamber      | 1.18       | 0.82       | 0.52         | 0.98         |
| Four-Chamber     | 1.49       | 0.78       |              |              |
| Four-Chamber serial | -         | -          | 0.50         | 0.92         |
| Four-Chamber parallel | 2.43     | 4.60       |              |              |

4. Conclusion

A model on tray bioreactor for solid-state fermentation has been developed. An ideal, homogeneously mixed heat and aeration (one chamber) can not explain the phenomena: aeration improve growth and enzyme productivity. This phenomenon showed, in reality, the fermentation is far from a homogeneously mixed system. The phenomenon of the presence or absence of aeration was explained well in 2 or 4 bioreactor chambers. What happens to the bioreactor can be predicted with a partially mixed system. The airflow pattern in the bioreactor tray used can be changed to parallel to increase productivity.

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