COMPUTER MODEL FOR BAKERS' YEAST PRODUCTION

By

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ABSTRACT

This paper concerns modelling of fermentation for bakers' yeast (Saccharomyces Cerevisiae) production using sugar cane blackstrap molasses as a feed. The problem constitutes of expressing physical phenomena, mathematically, as function of operating conditions.

The program developed is easy to run and does not cost much in terms of preparations. With this program different operating conditions were tested and results show that available equations are not exhaustive. Thus more research is recommended.

1. INTRODUCTION

Bakers' yeast growth requires carbon source and oxygen as major components. Carbon sources in use are petroleum products e.g. ethylene and agricultural products. Supply of agricultural products is difficult to maintain for they fluctuate with environmental conditions. However, agricultural products are more easily metabolized by microorganisms than the petroleum products. Thus, agricultural products, here sugar cane blackstrap molasses, has been chosen for modelling.

Oxygen is sparsely soluble in fermentation broth and its mass transfer rate is small. The mass transfer coefficient can be improved by using multiple impellers, changing reactors geometry, supplying pure oxygen and increasing agitation speed.

2. Development of the Model

Models of varying degree of complexity do exist. In this paper focus is made on lumped parameter model. This model expresses the specific growth rate as a function of the abiotic medium.

This paper covers only stirred tank reactor which is represented in Fig. 1 and can be run as batch, fed-batch or continuous reactor. The reaction which takes place can be represented, generally as:

\[ CH_\alpha O_\delta + \alpha_1 NH_\beta \Delta [ + essential \text{ Nutrients}] \rightarrow \gamma_1 CH_{\epsilon \alpha} O_\delta N_\epsilon ash + u_4 CO_\beta + \gamma_4 H_\delta O \]

where \( a, b, c, d, e \) are constants.
Mass balance of all models of stirred tank reactor i.e. batch, semi-batch and continuous are as follows:

**Cell mass**

\[ \frac{dX}{dt} = (\mu - a_2)X - D_2X + D_1X_0 \]  \hspace{1cm} (1)

**Substrate:**

\[ \frac{dS}{dt} = D_1S_0 - D_2S + \frac{M_sX - \mu X}{Y_s} \]  \hspace{1cm} (2)

where

\[ D_1 = \frac{F_1}{V} \land D_2 = \frac{F_2}{V} \]

for batch operation \( D_1 \), \( D_2 = 0 \), fed-batch \( D_1 > 0 \), \( D_2 = 0 \) and for continuous system \( D_1 = D_2 > 0 \).

Specific growth rate, can be expressed as function of the limiting substrate:

\[ \mu = \frac{\mu_{\text{max}} \cdot C_o}{K_o + C_o} \]  \hspace{1cm} (3)
Fig. 1: Stirred Tank Reactor
In aerobic fermentation air stream flows into the reactor to provide the necessary oxygen for growth. The dissolved oxygen is given by the equation below:

$$\frac{dC_o}{dt} = k_i a (C^* - C_o) - Q_{o2}X - D_i C_o$$  \hspace{1cm} (4)$$

where

$$Q_{o2} = \frac{\mu}{Y_o} + m_o$$  \hspace{1cm} (5)$$

$$Y_o = 3 \frac{\theta_i}{2 \alpha_i r_i (1 - \theta_b)}$$  \hspace{1cm} (6)$$

$$\theta_b = \frac{a_i r_b}{a_i r_s} Y_s$$  \hspace{1cm} (7)$$

and

$$Y_s = y_b \frac{a_i}{a_b}$$  \hspace{1cm} (8)$$

It has been observed that the value of $Y_s$ varies as

$$0 \leq Y_s \leq \frac{a_i r_s}{a_b r_b}$$  \hspace{1cm} (9)$$

but it never exceeds 0.7.

2.1 Oxygen Mass Transfer Coefficient

In agitated aerobic reactors the rate of oxygen transfer varies with power supplied for agitation of the broth. Power requirement depends on the liquid rheology. Most fermentation media involving yeast, exhibit Newtonian fluid behaviour. And an empirical correlation developed by Michel and Miller is:

$$p_s = 0.0312 P.Fr^{-0.044} (Q/ND_i^3)^{-0.28} (H/D_i)^{0.6}$$  \hspace{1cm} (10)$$

where

$$p = N_e D_i^5 N^2$$  \hspace{1cm} (11)$$

and

$$N_e = 5 - 6 (Oosterhuis et AL. (1983))$$

A number of correlations for expressing oxygen transfer have been developed. Among them Richards' correlation is adopted for laboratory scale fermenters (6 - 27L) whereas Fukuda - Richards correlation is adopted for pilot scales (100 - 42,000L)

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\[ k_i \alpha = K \left( \frac{P_i}{V} \right)^{0.4} N^{0.7} V^{0.5} - \text{Richards}(1961) \text{correlation} \] (12)

\[ k_i \alpha = (2.0 + 2.8 N_i) \left( \frac{P_i}{V} \right)^{0.56} N^{0.7} V^{0.4} \text{Fukuda/Richards correlation.} \] (13)

2.2 Solubility of Oxygen

Fermentation broth contains a mixture of organic and inorganic solutes. Equilibrium oxygen concentration in such situation can be obtained by log - additive approach (Schumpe (1985), Popovic et al. (1979)). However, since it is very difficult to measure oxygen concentration analytically, activity coefficient is measure instead. Assuming ideal gas conditions the solubility is expressed as:

\[ C_o = \frac{M_o P_o}{RT} \] (14)

With the basic equations an interactive program for single stage reactor to facilitate the investigation on the most suitable conditions was developed.

The program was tested for the following data combinations:

Table 1: Test Data Combination

<table>
<thead>
<tr>
<th>Substrate Conc. (g/l)</th>
<th>Agitation speed (rpm)</th>
<th>No. of agitators</th>
<th>Pressure (Bar)</th>
<th>Air flow (VVM)</th>
</tr>
</thead>
<tbody>
<tr>
<td>5</td>
<td>200</td>
<td>1</td>
<td>1.1</td>
<td>0.2</td>
</tr>
<tr>
<td>10</td>
<td>300</td>
<td>2</td>
<td>2.0</td>
<td>0.5</td>
</tr>
<tr>
<td>20</td>
<td>400</td>
<td>3</td>
<td>3.0</td>
<td></td>
</tr>
<tr>
<td>40</td>
<td>500</td>
<td></td>
<td>4.0</td>
<td>0.7</td>
</tr>
<tr>
<td>60</td>
<td></td>
<td></td>
<td>5.0</td>
<td>1.0</td>
</tr>
</tbody>
</table>

3. MODELLING RESULTS AND DISCUSSION

The results are shown in figures 2 - 5. In Fig. 2 it can be seen that the mass transfer coefficient increases with number of impellers for Fukuda/Richards correlation whereas Richards correlation shows no effect. Number of Impellers however, is limited by liquid height. And an economical maximum in use is 3 impellers.
Fig. 3 shows that increase in agitation speed favours mass transfer as this increase turbulence. Effectiveness of mixing, however, varies with type of agitator. Thus further investigations and establishment of equations for various types of agitators are called for.

Increase in inlet air flowrate favours mass transfer coefficient as shown in Fig. 4. Nevertheless, restriction must be made to avoid flooding.

Availability of oxygen for bioprocess is highly affected by the pressure of the inlet air. Increase in the air pressure increase the oxygen depletion time and biomass yield as shown in Fig. 5. Since fermenters are operated at ambient pressure then increase in inlet air pressure will diminish residence time of the gas in the reactor thus decreasing oxygen mass transfer. Alternatively, pure oxygen has to be used. And it might be worth investigating the feasibility of fermentation in pressure vessels.

CONCLUSION

With above remarks it can be concluded that:
- Oxygen transfer is very small especially if air is supplied. Growth stops the moment oxygen concentration becomes lower than biomass demand.
- Existing mass transfer equations are not exhaustive. As such different types of agitators for example, cannot be tested for suitability. Hence more research on such lines is required.
- Specific death rates as expressed by Arrhenius equation, does not take into account the effect of limiting nutrient concentration. Hence it looks unrealistic.
Reference

1. Oosterhuis, N.M.G. and N.W.F. Kossen
   Oxygen Transfer in a Production Scale Bioreactor

2. Popovic, M. and H. Niels
   Oxygen Solubilities in Fermentation Fluids

3. Schumpe, A.
   Gas solubility in Biomedia
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   Modelling of Fermenter for Bakers Yeast Production
NOMENCLATURE

\( a_b \) weight fraction of carbon in organic substrate
\( a_s \) weight fraction of carbon in biomass
\( C_0 \) Oxygen Concentration \( \text{kg/m}^3 \)
\( D_{h, d} \) Dilution rate \( \text{h}^{-1} \)
\( D_i \) Diameter of Impeller \( m \)
\( F_{1, f} \) Volumetric Flowrate of liquid phase \( m^3\text{h}^{-1} \)
\( F_r \) Froude number
\( H \) Height of liquid \( m \)
\( k_{la} \) overall mass transfer Coefficient \( \text{kg m}^{-3}\text{h}^{-1}\text{at m}^{-1} \)
\( K \) Constant
\( K_0 \) Saturation Constant for Monod kinetic model \( \text{kg m}^{-3} \)
\( m^o \) Maintenance Coefficient based on oxygen \( \text{g mol O}_2\text{(g biomass.h)}^{-1} \)
\( m_s \) Maintenance Coefficient based on substrate \( \text{g substrate(g biomass.h)}^{-1} \)
\( M_g \) Molecular weight of gas \( \text{g mol} \)
\( N \) Impeller speed \( \text{rpm} \)
\( N_i \) Number of Impeller
\( N_o \) Power Number
\( P \) Non-gassed Power Supply \( W \)
\( P_g \) Gassed power \( W \)
\( P_o \) Partial Pressure of Oxygen \( P_a \)
\( Q \) Volumetric gas Flowrate \( m^3\text{h}^{-1} \)
\( Q_{o2} \) Specific Rate of Oxygen Consumption \( \text{g mol O}_2\text{(g biomass.h)}^{-1} \)
\( r_b \) Reduction potential of organic substrate
\( r_s \) Reduction potential of biomass
\( R \) Universal Gas Constant \( kJ(kmol.K)^{-1} \)
\( Re \) Reynolds Number
\( S \) Organic substrate concentration \( \text{kg m}^{-3} \)
\( S_0 \) Substrate concentration in feed medium \( \text{kg m}^{-3} \)
\( T \) Time h; temperature \( ^\circ C \)
\( T \) Temperature \( ^\circ K \)
\( V \) Fermenter Volume \( m^3 \)
\( X \) Biomass concentration \( \text{kg m}^{-3} \)
\( X_o \) Biomass concentration in the feed medium \( \text{kg m}^{-3} \)
\( \gamma_b \) Biomass stoichiometric constant
\( Y_o \) Biomass yield based on oxygen \( \text{g biomass(g mol O}_2\text{)}^{-1} \)
\( Y_s \) Biomass yield based on substrate \( \text{g biomass(g substrate)}^{-1} \)
Greek Letters

\( a_1 \)  
Stoichiometric coefficient  
mol

\( a_2 \)  
Specific growth rate  
h^{-1}

\( \mu \)  
Specific growth rate h^{-1}; Viscosity  
cP

\( \theta_b \)  
Constant  
-

\( \beta \)  
Stoichiometric coefficient  
mol

Superscripts

*Equilibrium concentration
Fig 2 Effect of Number of Impellers on the Mass Transfer Coefficient.

Fig 3 Effect of Agitation Speed on Mass Transfer Coefficient.
Fig. 4 Effect of Air Flowrate on Mass Transfer Coefficient.