

DETERMINATION OF AERATION CAPACITY OF BUBBLED COLUMN TYPE FERMENTORS WITH OXYGEN ELECTRODE

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A method for the determination of the coefficient of oxygen transfer, $k_L a$, in bubbled column type fermentors was derived on the basis of a two-phase dispersion model. This value can be calculated from the dynamic behaviour of the apparatus followed by an oxygen electrode. Verification of the method on a semi-pilot scale showed that the underlying assumptions are correct and the experimental results are in accord with literature data.

The rate of biosynthesis during aerobic cultivation of microorganisms is conditioned by the rate of oxygen transfer between the cultivation medium and bubbling air. The bubbled column as one of new concepts of the fermentor is constructionally simpler than the classical mixed fermentors and is hence preferable especially in building plants for the preparation of microbial proteins in underdeveloped countries¹. Greenshields^{1,2} reported the use of bubbled columns as fermentors in the manufacture of beer, vinegar, fodder yeast, and citric acid. One of possible applications is also biological purification of waste water³. Oels^{4,5} used a 3.9 m long semi-pilot plant column for the cultivation of yeast on glucose, ethanol, and methanol.

For the determination of oxygen transfer in bubble columns, a number of methods has been proposed⁶ based on chemical determination of the interphase transfer, however they are very laborious. The so-called dynamic method is preferable for the determination of aeration capacity of mixed fermentors; its use for testing of industrial apparatuses has been investigated in this laboratory for a long time⁷⁻⁹. Linek¹⁰ used this method for the determination of the transfer coefficient in a counter-current filled column and an absorber with a wetted wall. The method of determination is based on the assumption of a plug flow of both the liquid and gas, which is fulfilled only in the case of high velocities. During practical fermentation in a flow-through system, however, the linear velocity of the liquid is very small, and with respect to an intense back mixing^{6,11,12} the assumption of plug flow is not justified. The Bodenstein number values are in technologically interesting cases in the range 10^{-5} –0.1. Hence, the bubble column type fermentor can be considered from the point of view of stirring rather as an ideal stirrer than a reactor with a plug flow.

Akita¹³ found a relatively complicated correlation for the calculation of $k_L a$ in bubble columns. In the case of a cultivation medium (mineral salts, starch, oil, lard, flour hydrolysate, *etc.*) the viscosity, surface tension, *etc.* cannot be reliably estimated, but it is possible to carry out an experiment on a small model and to apply the results in practice.

THEORETICAL

The back mixing in a bubble column can be described by a two-phase dispersion model, proposed first by Sleicher¹⁴. Basic mass balances derived from it are in principle used nowadays in modelling both chemical reactors¹⁵⁻¹⁸ and fermentors¹⁹. In our case, the two-phase model was used in analysis of the transient characteristic of the batch arrangement for the liquid and flow-through one for the gas. If we place the sensor for the measurement of the dissolved oxygen in one half of the height of the gas-liquid dispersion, we measure with respect to the linear profile of the partial pressure of oxygen along the column¹⁵ the rate of transfer at a mean value of the driving force of absorption. The oxygen balance in the liquid can be accordingly written in the form

$$\frac{\partial C}{\partial t} = D_e \frac{\partial^2 C}{\partial x^2} + k_L a (P y / H - C). \quad (1)$$

The initial and boundary conditions are

$$C = C_0, \quad x \in (0, L), \quad t = 0 \quad (2)$$

$$\partial C / \partial x = 0, \quad x = 0; \quad D_e \partial C / \partial x = k_L (P y / H - C), \quad x = L. \quad (3)$$

On introducing the mean driving force for $x = L/2$ and the dimensionless variables

$$Y = (C_\infty - C) / (C_\infty - C_0), \quad z = x / L \quad (4)$$

Eqs (1)–(3) can be rewritten in the form

$$\frac{\partial Y}{\partial t} = \frac{D_e}{L^2} \frac{\partial^2 Y}{\partial z^2} - k_L a Y, \quad (5)$$

$$Y = 1, \quad z \in (0, 1), \quad t = 0 \quad (6)$$

$$\frac{\partial Y}{\partial z} = 0, \quad z = 0; \quad \frac{\partial Y}{\partial z} = -\frac{k_L L}{D_e} Y, \quad z = 1. \quad (7)$$

These equations are formally the same as the mathematical formulation of internal diffusion in a catalyst with a first-order reaction²⁰. Analytical solution can be found by using the Laplace transformation, but it leads to complicated expressions. The method of linearization^{21,22} of original Eq. (5) using differential operator

$$-\lambda^2 \approx \partial^2 / \partial z^2 \quad (8)$$

enables to transform this partial differential equation into an ordinary one:

$$\frac{dY}{dt} = -(k_L a + D_e \lambda^2 / L^2) Y \quad (9)$$

with $Y(0) = 1$. This equation is analogous to a balance of an ideal mixer and so enables to make use of the results obtained in evaluating data from mixed fermentors⁷⁻⁹. The linear operator λ which allows to correct the aeration capacity is calculated from the transcendental equation

$$\lambda \operatorname{tg} \lambda = k_L L / D_e, \quad (10)$$

whose right side can be easily evaluated with the use of the modified Frösling equation^{23,24}

$$d_B k_L / D_m = 2 + \operatorname{Re}_B^{1/2} \operatorname{Sc}^{1/3}. \quad (11)$$

Since the rate of a bubble in the foam at the upper level of the gas-liquid dispersion is zero, we can estimate the value of the right side of Eq. (10), which lies in technologically important cases in the range $10^{-5} - 10^{-2}$. For sufficiently small λ values Eq. (10) gives

$$\lambda^2 \approx 2 D_m L / D_e d_B. \quad (12)$$

The concentration of dissolved oxygen in the half of the height of the bubble column is then obtained by integration of Eq. (9):

$$Y(t) = \exp [-(k_L a + 2 D_m / d_B L) t]. \quad (13)$$

To measure the oxygen concentration, an oxygen electrode covered with a membrane is used⁷⁻⁹, whose signal is delayed with respect to the actual course of the concentration. A simple transient characteristic of the first order is usually sufficient to describe this delay. Since the electrode is placed in a two-phase system, its interaction with the gas phase plays a role especially at higher gas hold-up. This effect

can be described quantitatively in terms of the local gas hold-up defined as the ratio of the time of exposition of the electrode, t_B , in the gas phase to the total time of measurement, t (ref.^{2,5}):

$$\varepsilon = t_B/t. \quad (14)$$

Since the time of contact of the bubbles is very short in comparison with the time of equilibration of the dissolved oxygen concentration in the column, the mean measured concentration can be approximated by a weighted sum of the dissolved oxygen concentrations in the bulk and at the interface. The weight is equal to the local gas hold-up defined as

$$\bar{Y} = (1 - \varepsilon) Y + \varepsilon Y(\infty). \quad (15)$$

The electrode signal is retarded, so the record of the transient process will be described by the equation

$$dY_p/dt = (\bar{Y} - Y_p)/K_p, \quad (16)$$

where Y_p denotes normalized response and K_p electrode constant, *i.e.* time necessary to attain 63.2% value of response at a step change of the concentration. From Eqs (14)–(16) we obtain a relation enabling to evaluate the aeration capacity, $k_L a$:

$$K_p dY_p/dt + Y_p = (1 - \varepsilon) \exp [-(k_L a + 2D_m/d_B L) t]. \quad (17)$$

For a rapid calculation, a linearized form of Eq. (17) is preferable. If we plot $\ln \cdot (K_p dY_p/dt + Y_p)$ as function of time t , the slope of the straight line led through the experimental points is equal to the value of $k_L a + 2D_m/d_B L$.

EXPERIMENTAL

Measurements were performed in a glass column of 0.20 m inner diameter and a bed height of 0.245, 0.755, 1.795, or 3.24 m. It was filled with a fermentation mineral medium for the continuous cultivation of yeast on ethanol of the following composition (g/l): $(\text{NH}_4)_2\text{SO}_4$ 1.0, KH_2PO_4 1.6, $\text{MgSO}_4 \cdot 7 \text{H}_2\text{O}$ 0.5, $\text{ZnSO}_4 \cdot 6 \text{H}_2\text{O}$ 0.01 (dissolved in tap water). The air entered the column through a dispersator from a fritted glass bottom and its rate of flow was $1.95 \cdot 10^{-4}$ to $1.97 \cdot 10^{-3} \text{ m}^3/\text{s}$. The temperature of measurement was 22°C. The contents of the column was freed from dissolved oxygen by nitrogen. The air flow rate was then controlled by a rotameter and the response of the column was followed by an oxygen electrode, which was a standard polarographic Pt—Ag electrode covered with a polypropylene membrane of a thickness of 15 μm . Its time lag of response was 1.7 s. The electrode signal was recorded by a transistorized recorder.

RESULTS AND DISCUSSION

For the treatment of experimental data, a program enabling to compute the value of $k_L a$ from the linearized form of Eq. (17) was compiled. In our experimental arrangement, the value of $2D_m/d_B L$ was approximately by two orders of magnitude lower than $k_L a$. The time derivative of Y_P was calculated by using the cubic spline function²⁶.

In the graphical verification of Eq. (17) it turned out that the points measured in the interval of Y_P from 1 to 0.85 do not fit the assumed dependence. This phenomenon was observed also with mixed absorbers⁷ and is the result of a slow attainment of the coalescence-redispersion equilibrium in the gas-liquid emulsion. It is hence preferable not to include the points from this region into the evaluation, as has been recommended also in the case of mechanically mixed fermentors⁷.

The gas hold-up was determined from the difference between the volumes of the liquid and gas-liquid dispersion. The calculated values of $k_L a$ were plotted as function of the gas hold-up (Fig. 1) as recommended in a review by Sherwood⁶ on the basis of a work of Akita¹³. The obvious agreement between the evaluated data and the correlation measured by an independent method¹³ is in favour of the method proposed by us.

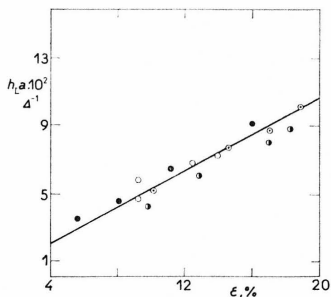


FIG. 1

Calculated Values of $k_L a$ (s^{-1}) for Various Bed Heights

○ 0.245 m, ● 0.755 m, ● 1.795 m, ⊙ 3.24 m. The line denotes correlation according to Akita and Yoshida¹³.

LIST OF SYMBOLS

- a specific interfacial surface
- C concentration of dissolved oxygen
- D_m molecular diffusion coefficient of oxygen in water
- D_e diffusion coefficient
- d_B mean bubble diameter
- H Henry's constant
- K_p electrode constant
- k_L oxygen transfer coefficient
- L height of gas-liquid dispersion in apparatus
- P total pressure

Re_B	Reynolds number referred to a bubble
Sc	Schmidt number
t	time
t_B	time of contact of gas bubble with electrode
x	length coordinate
Y	normalized oxygen concentration in liquid
\bar{Y}	normalized mean oxygen concentration
Y_p	normalized electrode signal
y	oxygen concentration in output gas
z	transformed coordinate (Eq. (4))
λ	linear operator (Eq. (8))
ε	gas hold-up

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