OXYGEN TRANSFER DURING BATCH CULTIVATION IN AN AIRLIFT TOWER FERMENTOR

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Oxygen transfer was studied in an airlift tower fermentor with motionless mixers (Kenics type). The intensity of oxygen transfer was characterized by its volumetric mass transfer coefficient, k_La, which was determined by the balance method. Experimental data of k_La were described by correlation equations and compared with results obtained namely for the airlift fermentor without motionless mixers with respect to energetic consumption. In addition, growth characteristics of the yeast culture *Torulopsis ethanolitolerans* cultivated on ethanol were also investigated.

Oxygen dissolved in the nutrient solutions is one of essential components vital to the microorganism growth. With respect to the low rate of oxygen transfer from the gas to the liquid phase, aeration and agitation of nutrient solutions rank among the necessary conditions of fermentation processes.

Besides the conventional mechanically agitated fermentors, the bubble column fermentors are used widely for submersion aerobic cultivations of microorganism as well. Advantages of these nonmechanically agitated fermentors are known to include namely efficient use of energy for aeration, simple design and construction. The construction of the bubble column reactors leads to attaining the most extensive transport rate of oxygen into the liquid phase at the low possible energetic consumption. The application of motionless mixers built in the reactors of the column type is one of the possibilities. Fan et al.[1], Hsu et al.[2,3], and Wang and Fan[4] studied the oxygen transfer in a bubble column packed with Koch motionless mixers. They achieved higher mass transfer coefficients when the motionless mixers had been inserted in the column. Oxygen transfer in countercurrent multistage columns was studied for aqueous solutions with low viscosity by Voigt and Schügerl[5] and for aqueous solutions with high viscosity by Voigt et al.[6]. Gas holdups and volumetric mass transfer coefficients were studied in a bubble column with 50 different gas–liquid systems comprising pure and mixed organic liquids and various gases by Öztürk et al.[7]. Shah et al.[8] discussed also mass transfer in bubble column reactors and showed the review of correlations for volumetric mass transfer coefficient. Studies in tower reactors with viscous liquids on gas–liquid mass transfer were reviewed by Schumpe and Deckwer[9] who also derived dimensionless correlations on basis of their own and literature data to prediction of k_La in fermentation broths. As some batches of fermentors show the non-Newtonian behaviour, Godbole et al.[10] investigated also oxygen transfer in bubble columns for non-Newtonian liquids.

In this work, an airlift fermentor is investigated for growth characteristics of microorganism and oxygen transfer under conditions of batch cultivation. The effect of the presence of motionless mixers (Kenics type) on oxygen transfer was verified.

As the direct measurement of the interfacial area during the fermentation is not possible, the rate of interphase oxygen transfer is characterized by the volumetric liquid side mass transfer coefficient, k_La, which can be determined, for example, by the oxygen balance method.

EXPERIMENTAL

The batch cultivations were carried out in an airlift tower fermentor which belongs among the bubble column reactors. The flow of the gas–liquid dispersion in the inner tube and in the space between the inner and the outside tube is due to the different density of the gas–liquid mixture in these spaces. The jacket of the airlift fermentor consisted of a 105 mm diameter glass pipe 800 mm in length. Three inner tubes were located over the gas distributor and their diameter was 33·5 mm and their length was 400 mm. The stainless steel perforated plate with 12 holes was used as the gas distributor. The diameter of every hole was 1 mm. The ratio of the hole area to the outside pipe cross-sectional area was 0·11%. The complete apparatus is shown schematically in Fig. 1. Seven elements of the static mixer Kenics-180° were placed in each of the inner tubes. The elements of the motionless mixer (Fig. 2) were made of plexiglass and their width was equal to the diameter of the inner tube, while their length was 1·5 times their width.

The salt solution contained: 85% H_3PO_4, 35 ml; KOH, 30 g; $MgSO_4 . 7 H_2O$, 32 g; $ZnSO_4 . 7 H_2O$, 0·5 g; NaCl, 12·2 g; $CuSO_4.5 H_2O$, 0·02 g; and $MnCl_2.4 H_2O$, 0·05 g per litre. The solution of Ca and Fe was prepared separately and contained 165 g of $CaCl_2.2 H_2O$, and 10·6 g of $FeCl_3.6 H_2O$ per litre. The fermentor was filled with the medium in which the salt solution and the solution of Ca and Fe were added according to the expected biomass production, i.e. 1 ml of the salt solution and 0·1 ml of the solution of Ca and Fe on 1 g biomass addition[11]. At beginning of each cultivation, 5 g of $(NH_4)_2SO_4$ were also added in the fermentor. The bulk of the liquid batch was 5 l. During cultivation the mixture of ethanol and ammonia liquor mixed in the volumetric ratio 3·5 : 1 was dosed continuously to be pH 4. Defoaming agent (oleic acid) was added to the nutrient solution during all the cultivation runs. The temperature of the liquid batch was kept at 30°C. All cultivations were performed with the yeast culture *Torulopsis ethanolitolerans* from the collection of the Research Institute of Fodder Industry, Prague. The initial dry biomass concentration was always less than 2 g l^{-1}. The cultivation studies were performed within the range of the air flow rates from 5 to 30 l min^{-1}. Corresponding values of the superficial air velocities were from 0·96 to 5·77 cm s^{-1}. The concentration of oxygen dissolved in the fermentation broth was measured by means of an electrode Pt–Ag–AgCl and registered by the linear recorder. The concentrations of oxygen, nitrogen, and carbon dioxide in the exit gas as well as in the gas entering the fermentor were measured by means of the chromatograph Chrom 5.

The dry weight of the yeast in the cultivation medium was determined gravimetrically. After centrifugation of the samples of the culture broth, the solids were dried for 3 h at 105°C.

RESULTS AND DISCUSSION

Growth Characteristics

Typical biomass propagation data for the airlift fermentor with motionless mixers are shown in Fig. 3. The course of these dependences depends on the achievement of the oxygen limitation state. At the lowest air flow rate of $5 \, l \, min^{-1}$, the oxygen limitation state was set approximately in a quarter of an hour after beginning of the run. It is clear that the time of the oxygen limitation decreased with increasing air flow rate. In an oxygen limitation state, the biomass productivity was raised from 0·15 to $1{\cdot}6 \, g \, l^{-1} \, h^{-1}$ with increasing aeration. During cultivation runs, i.e. after 5·5 h, the concentration of dry biomass raised nearly twice at an air flow rate of $5 \, l \, min^{-1}$, while the concentration of dry biomass raised approximately seven times

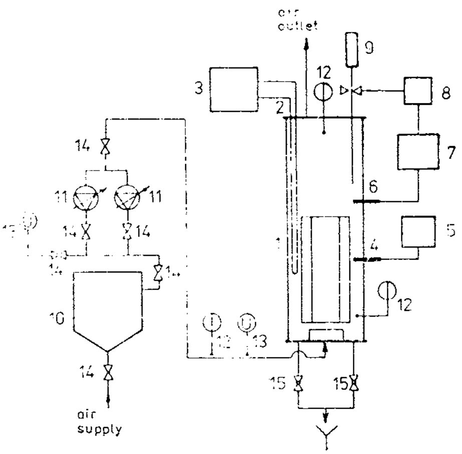

FIG. 1

Flowsheet of the experimental setup. 1 Airlift fermentor, 2 heat exchanger, 3 thermostat, 4 oxygen electrode, 5 linear recorder, 6 pH electrode, 7 automatic titrimeter, 8 electromagnetic valve, 9 measuring burette, 10 pressure leveling vessel, 11 air rotameter, 12 thermometer, 13 U-tube pressure gauge, 14 valve, 15 coke

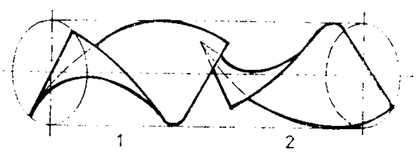

FIG. 2

Arrangement of two elements of the motionless mixer (Kenics type). 1 Right-hand element, 2 left-hand element at an air flow rate of 30 l min^{-1}. The maximum specific growth rates of 0·28 to 0·45 h^{-1} achieved at given air flow rates agree with published values for yeast[12].

Consequently, the values of the criterion, P_G/p, which characterizes the energy consumption related to the biomass production, were evaluated as well. The values of P_G/p rose from 0·513 to 1·25 W h g^{-1} with increasing air flow rate and are slightly greater than those for the airlift fermentor without motionless mixers[13]. For cultivations of yeast *Candida utilis* in a tower fermentor with ejector-type gas distributor, the value of P_G/p = 1·64 W h g^{-1} was achieved for maximum productivity by Zahradník et al.[14].

Oxygen Transfer

In order to determine the volumetric liquid side mass transfer coefficient, k_La, the oxygen balance method was applied for the system in a steady state. It was assumed that the liquid phase in the fermentor was complete mixed. The calculation of k_La was described earlier[15,16]. The measured values of k_La were plotted against the superficial air velocity in Fig. 4. In a comparison with an airlift fermentor without motionless mixers[13], it showed that the greater values of k_La were attained for the airlift fermentor with static mixers in the region of the air flow rates within 15 and 30 l . . min^{-1}. At air flow rates 5, and 10 l min^{-1}, the influence of the presence of the motionless mixers on the value of k_La was not obvious. It must be noted that k_La values raised from 0·014 to 0·11 s^{-1} for the fermentor without static mixers. The results obtained can be compared with those published for cultivations of yeast

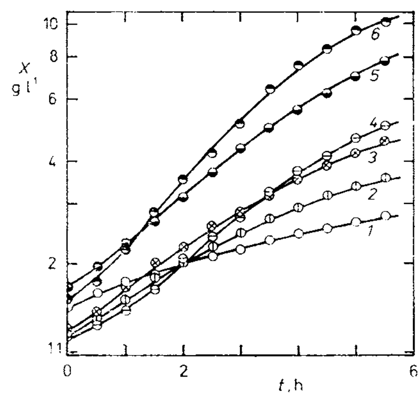

FIG. 3
Time profiles of yeast growth. Air flow rate (l min^{-1}): *1* 5; *2* 10; *3* 15; *4* 20; *5* 25; *6* 30

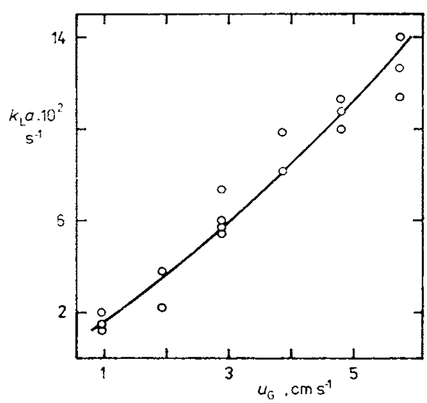

FIG. 4
Volumetric coefficient of oxygen transfer as a function of the superficial air velocity

culture *Candida utilis*. In a laboratory mechanically agitated fermentor, Páca et al.[17] achieved values of k_La approximately in the range 0·025–0·05 s^{-1} in calcium and natrium bisulphite liquors, while for fermentors with considerable different working volume Votruba et al.[18] calculated k_La values ranging from 0·181 to 0·321 s^{-1}. In a tower fermentor with ejector-type gas distributor, Zahradník et al.[14] achieved k_La values from 0·127 to 0·212 s^{-1}. Owing to the different character of cultivated microorganisms and fermentor construction used, a comparison of these data with experimental results obtained can, however, serve only to a qualitative estimation of efficiency of the airlift fermentor studied in this work. It is not also possible to compare absolute values of k_La with results obtained by the authors[1–4] because they worked with a continuous-flow system within a range of substantially higher superficial gas velocities in contrast to the batch system used in the present work.

The dependences of the volumetric coefficient of oxygen transfer on either the superficial air velocity[19] or the gas power input[20] are often described. By the least squares method, the correlation between the superficial air velocity in m s^{-1} and the volumetric mass transfer coefficient in s^{-1} was determined in the form

$$k_La = 4{\cdot}21u_G^{1.21} \tag{1}$$

(correlation coefficient 0·98), and correlation between the gas power input in W m^{-3} and the volumetric mass transfer coefficient in s^{-1}, as

$$k_La = 2{\cdot}94 \,.\, 10^{-4}P_G^{0.86} \tag{2}$$

(correlation coefficient 0·94). In addition, the correlation between dimensionless groups was also derived. Values of k_La obtained were included in the Sherwood

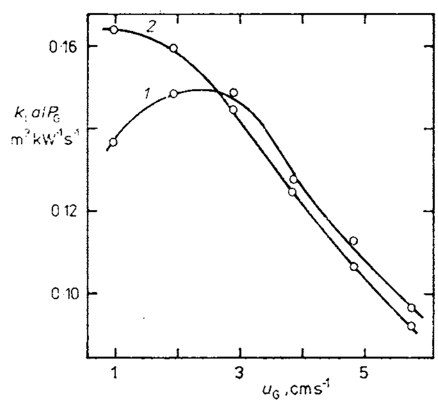

FIG. 5

Parameter k_La/P_G as a function of the superficial air velocity. *1* Fermentor with motionless mixers, *2* fermentor without motionless mixers number and they expressed their dependence on Froude number in the form

$$Sh = 1{\cdot}77 \,.\, 10^7 Fr^{1{\cdot}20} \tag{3}$$

(correlation coefficient 0·98). Eq. (*3*) was derived for Sherwood numbers ranging from $5{\cdot}63 \,.\, 10^4$ to $6{\cdot}06 \,.\, 10^5$ and Froude numbers ranging from $9{\cdot}5 \,.\, 10^{-3}$ to $5{\cdot}7 \,.\, 10^{-2}$. The Froude number exponent is approximately the same as exponent in Eq. (*1*) because the values of the Froude number were proportionate to those of the superficial gas velocity. Öztürk et al.[7] and Schumpe and Deckwer[9] reported lower influence of the Froude number ($Sh \sim Fr^{0.68}$, $Sh \sim Fr^{0.49}$, respectively), their correlations, however, include also the Bond, Schmidt, and Gallilei numbers. As the intrinsic properties (density, viscosity, surface tension) of broth changed very little during cultivations at all runs, the further dimensionless groups are not included in Eq. (*3*).

Furthermore, results obtained in this work were compared with respect to an energetic consumption as well. For given air flow rates, the parameter k_La/P_G which characterizes the efficiency of utilization of the power input supplied into to the volume unit of the batch of the liquid was evaluated. The values of k_La/P_G determined for both fermentor with motionless mixers and without motionless mixers[13] are plotted with respect to the superficial air velocity to demonstrate the influence of the motionless mixers on the oxygen transfer (Fig. 5). Results obtained can be compared with Zahradník et al.[14] who found k_La/P_G of 0·021 and 0·034 $m^3 \,.\, kW^{-1} \, s^{-1}$.

In conclusion, it must be noted that the differences were observed for a fermentor with and without motionless mixers. A very fine dispersion of gas recirculated into the space between the inner and the outside tube is formed in the presence of the motionless mixers. In the system with motionless mixers the stay time of air bubbles in the space of the inner tube is increased, but gas holdup is decreased in the space between the inner and outside tube due to lower liquid recirculation velocity. The overall gas holdup decreases accordingly, but in the region of greater air flow rates the volumetric coefficient of oxygen transfer increases with insertion of the motionless mixers which, however, cause a larger pressure drop in the inner tubes. Finally, results of cultivation experiments obtained in this work were derived from experimental data observed in a small-scale fermentor and their application, however, would not be likely suitable in the large-scale equipments.

SYMBOLS

D	oxygen diffusivity in water, $m^2 \, s^{-1}$
d	diameter of fermentor, m
Fr	$= u_G/(gd)^{0.5}$ Froude number

g gravity acceleration, $m\,s^{-2}$
k_La volumetric coefficient of oxygen transfer (aeration capacity), s^{-1}
P_G specific gas power input related to a unit of liquid phase volume, $W\,m^{-3}$
p overall productivity of the fermentation process, $g\,l^{-1}\,h^{-1}$
Sh $= k_Lad^2/D$ Sherwood number
t time, h
u_G superficial gas velocity, $m\,s^{-1}$
X dry biomass concentration, $g\,l^{-1}$

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