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Research Article

## Oxygen Mass Transfer Rate in a Pressurized Lab-Scale Stirred Bioreactor

Oxygen mass transfer from air to the liquid phase in bioreactors with aerobic cultures has long been a serious impairment to the productivity of various bioprocesses. An increase of the oxygen mass transfer rate (OTR) can be the key to overcome oxygen limitation. The influence of higher air pressure on OTR was measured and a significantly enhanced OTR could be obtained. The oxygen volumetric mass transfer coefficient ( $k_L a$ ) was described by a function of the air pressure in a stirred lab-scale pressurized bioreactor. The correlation obtained proved that  $k_L a$  slightly decreased with higher air pressure, following a power function.

**Keywords:** Air pressure, Oxygen mass transfer, Oxygen transfer coefficient, Pressurized bioreactor

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### 1 Introduction

In aerobic bioprocesses, oxygen is a key substrate. Due to its low solubility in aqueous solutions, it is important to ensure an adequate delivery of oxygen from a gas stream to the culture broths. The oxygen mass transfer rate (OTR) and volumetric oxygen mass transfer coefficient ( $k_L a$ ) must be known, and if possible predicted to achieve an optimum design operation and scale-up of bioreactors. OTR and  $k_L a$  are influenced by a numerous parameters like physical properties of gas and liquid, operating conditions, and geometrical parameters of the bioreactor, and also by the presence of biomass, i.e., the consumption of oxygen by the cells [1,2]. Both parameters can be related and stated mathematically as:

$$\text{OTR} = k_L a (C^* - C) \quad (1)$$

where  $C^*$  is the solubility of oxygen in the liquid,  $C$  is the dissolved oxygen concentration in the liquid, and  $k_L a$  is made up of the mass transfer coefficient ( $k_L$ ) and the interfacial area ( $a$ ). In this equation, the term  $(C^* - C)$  is considered to be the driving force which causes oxygen to transfer from the gas phase to the liquid phase [3].

Several empirical correlations have been proposed to estimate the  $k_L a$  in mechanically agitated bioreactors (STR), with the most well-known function:

$$k_L a = a \left( \frac{P_g}{V} \right)^\delta (v_s)^\gamma \quad (2)$$

where  $P_g$  represents the power input to the aerated bioreactor,  $V$  is the bioreactor working volume, and  $v_s$  is the superficial gas velocity. The parameters  $a$ ,  $\delta$ , and  $\gamma$  are dimensionless constants.

To calculate the power input to the aerated system ( $P_g$ ), the Reynolds number ( $Re$ ) is determined by Eq. (3) and the power number ( $N_p$ ) by Eq. (4).

$$Re = \frac{D_i^2 N \rho}{\nu} \quad (3)$$

$$N_p = \frac{P_g}{\rho N^3 D_i^5} \quad (4)$$

where  $\rho$  represents the liquid density,  $N$  the agitation rate,  $\nu$  the liquid viscosity, and  $D_i$  the impeller diameter.

According to Cheremisinoff and Gupta [4], if the flow regime inside the system is turbulent ( $19\,070 < Re < 38\,141$ ),  $N_p$  is not a function of  $Re$  when the vessel is fully baffled. Consequently,  $P_g$  without aeration ( $P'_g$ ) can be determined by Eq. (5).

$$P'_g = K_T D_i^5 N^3 \rho \quad (5)$$

where  $K_T$  is a constant dependent on the impeller used.

Finally, to determine  $P_g$  in an aerated system, Eq. (6) can be used.

$$P_g = c \left( \frac{P'_g N D_i^3}{F_g^{0.56}} \right)^{0.45} \quad (6)$$

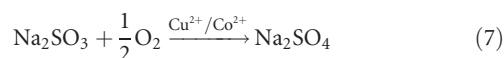
where  $c$  is a constant dependent on the impeller and  $F_g$  is the volumetric gas flow rate.

In order to overcome the oxygen limitation in aerobic microbial cultures, selection of adequate, normally high OTR

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values is crucial. Special aeration systems, e.g., aeration using oxygen-enriched air and increased reactor pressure are techniques applied to increase oxygen availability [5–8]. Also, the use of in situ production of oxygen [9] or the use of a second liquid phase of various organic compounds such as perfluorodecalin [10] or *n*-hexadecane [11] in the culture medium can increase the availability of oxygen to the microorganisms.

A number of methods have been developed to determine the oxygen transfer rate in bioreactors. The techniques vary according to the accuracy required and have advantages and disadvantages depending on the availability of the necessary analytical instruments and material and labor costs [12]. In the absence of microbial cells, the OTR can be estimated by the oxygen absorption rate of a sodium sulfite solution [13]. This technique is based on the reaction of sodium sulfite, a reducing agent, with the dissolved oxygen to produce sodium sulfate in the presence of a catalyst (usually a divalent cation of  $\text{Cu}^{2+}$  or  $\text{Co}^{2+}$ ). The reaction can be expressed as:



The reaction rate is much faster than the oxygen transfer rate and there is a concentration range of sodium sulfite (0.04–1 N) for which the oxygen concentration can be assumed as zero. Therefore, the oxidation rate is controlled by the rate of mass transfer and measures the overall rate. According to the reaction stoichiometry (Eq. (7)), the OTR is estimated as half of the rate of sulfite consumption. Thus, knowing OTR and oxygen solubility, the volumetric oxygen mass transfer can be determined by Eq. (8):

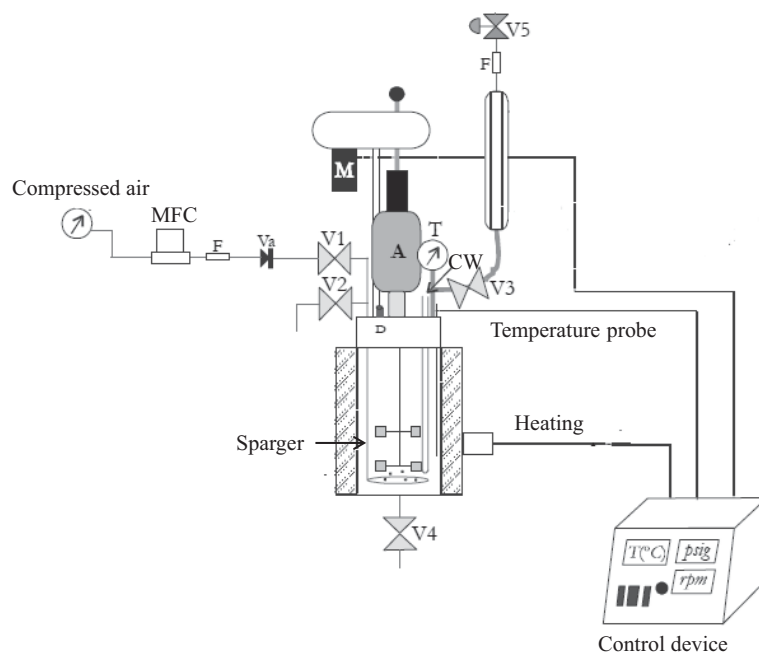
$$\text{OTR} = k_L a C^* \quad (8)$$

Some authors have demonstrated the applicability of pressurized bioreactors in microbial cultures, with enhancements in biomass and product yields [14–16]. Since these improvements could be related to an improvement in oxygen mass transfer due to the increase of oxygen solubility with pressure, it seems important to describe OTR and  $k_L a$  in such bioreactors. Thus, experimental values of OTR were obtained in a lab-scale pressurized bioreactor, by varying the air pressure, the aeration, and the stirring rates. Based on Eq. (2), data fitting to an empirical correlation for the prediction of the  $k_L a$  as a function of air pressure, power input of the aerated bioreactor, and superficial gas velocity was attempted.

## 2 Materials and Methods

### 2.1 Experimental Procedure

A 600-mL stainless steel stirred-tank bioreactor (PARR 4563, Parr Instruments, USA) with 400 mL of operating volume was used (Fig. 1). The bioreactor is a cylinder of 0.063 m diameter (total area 0.00312 m<sup>2</sup> and the ratio  $H/D$  is 3) equipped with



**Figure 1.** Schematic diagram of the stainless steel stirred-tank bioreactor (PARR 4563, Parr Instruments, USA): A – agitator; CW – cooling water; D – security disk; F – air filter; M – motor; MFC – mass flow controller; T – pressure transducer; Va – check valve; V1, V2, V3, V4 – valves; V5 – regulatory valve.

an impeller with two turbines of four pitched blades (0.035 m of diameter), a temperature probe, and a sparger tube for aeration. The gas flow rate was measured with a calibrated mass flow controller (Alicat scientific, Model MC-5SLPM-D). The parameters studied were stirring rate (200, 400, 600 rpm), aeration rate (0.5, 1, 2 vvm, measured under standard temperature and pressure conditions), and total air pressure inside the bioreactor (1–5 bar). The operating pressure was set by manipulation of the pressure of the inlet compressed air and the regulatory valve position in the exit gas line. The reactor was equipped with a pressure transducer (PARR 4842, Parr Instruments, USA) to monitor the total internal pressure.

### 2.2 Oxygen Transfer Rate (OTR)

OTR in bioreactors operating under different conditions was estimated in blank assays by the sulfite oxidation method [13] at 30 °C. A sodium sulfite solution (0.2 mol L<sup>-1</sup>) is oxidized to sodium sulfate in the presence of a catalyst ( $\text{CuCl}_2$ , 0.001 mol L<sup>-1</sup>). At regular times, samples of known volumes were collected and mixed with an excess of iodine solution (0.05 mol L<sup>-1</sup>). The amount of iodine that not reacted with sulfite ion was determined by spectrophotometric absorbance at 595 nm and converted to molar concentration using a previous calibration. The amount of residual sulfite can be estimated by:

$$[\text{SO}_3^{2-}] = \frac{0.05V_I - (V_I + V_s)I_2}{V_s} \quad (9)$$

where  $V_s$  is the sample volume,  $V_1$  is the iodine solution volume, and  $I_2$  is the molar concentration of the iodine solution.

As this method measures the rate of  $O_2$  absorption by a  $Na_2SO_3$  solution, it enables to predict the effect of pressure increase on the oxygen mass transfer capacity of the system.

### 2.3 $k_L a$ Modeling

A correction of Eq. (2) was made in order to predict the effect of absolute pressure,  $P$ , in  $k_L a$ :

$$k_L a = a \left( \frac{P_g}{V} \right)^\delta (v_s)^\gamma P^\beta \quad (10)$$

$k_L a$  values were obtained by dividing the experimental OTR data by the oxygen solubility at 30 °C.

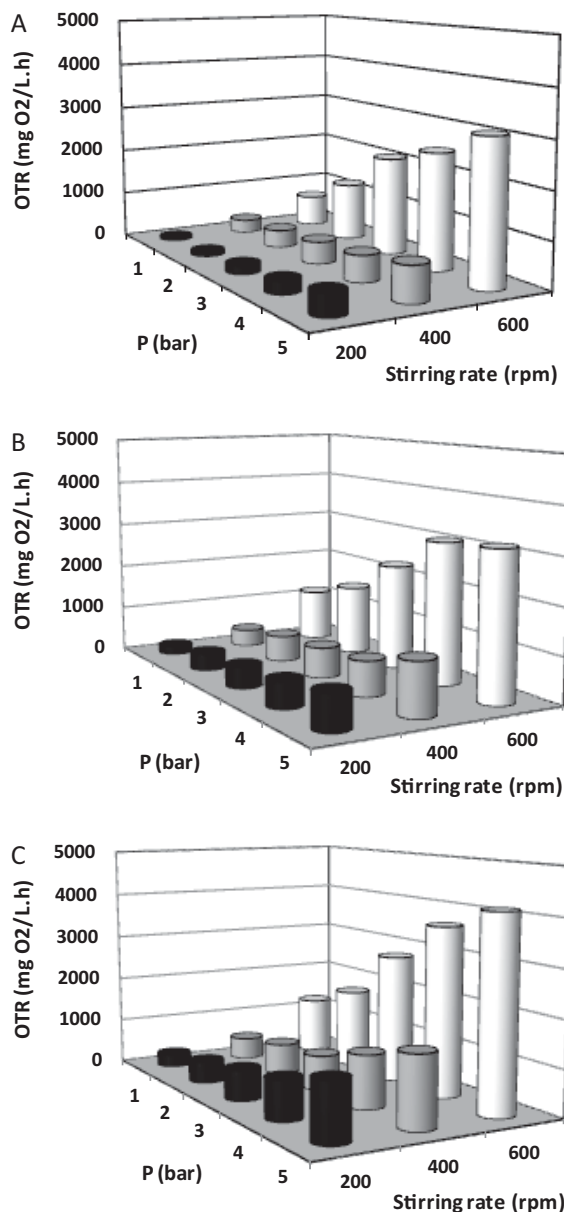
$P_g$  and  $v_s$  in Eq. (10) were calculated with the help of the equations presented in the introduction, converting the aeration rate to real  $F_g$ . The air flow rate inside the reactor was corrected from the measured one under standard conditions to the values of temperature and pressure of the assay using the ideal state gas equation.

In order to predict the effect of air pressure on the oxygen solubility, the Krichevsky-Kasarnovsky equation [17] was applied. The result proved that Henry's law was still valid for the air pressure range used in this work (up to 5 bar). The dimensionless parameters  $a$ ,  $\delta$ ,  $\beta$ , and  $\gamma$  were estimated by minimizing the sum of least squares of the difference between the experimental and modeled value of  $k_L a$ , using the Solver tool of Microsoft Excel 2007 software.

## 3 Results and Discussion

### 3.1 OTR Measurement

To evaluate the effect of total air pressure on OTR values, several experiments were carried out, i.e., changing stirring and aeration rates under increased air pressure up to 5 bar (Fig. 2). Separate analyses of each factor indicated that the increment of aeration and stirring rates and total air pressure inside the bioreactor led to an enhancement of the OTR value. At higher gas flow rates, the gas holdup in the bioreactor increases, leading to a higher surface area of bubbles which in turn increases the  $k_L a$  values. The change in gas flow rate affects the fractional gas holdup, and hence, the  $a$  and consequently  $k_L a$  values [18]. Thus, according to Eq. (8), the OTR value also increases. The OTR improvement was more pronounced with the increment of the stirring rate or the rise of air pressure than with the increase of aeration rate. At 1 bar of air pressure, a 4.3-fold enhancement was achieved with the increase of air flow rate up to 2 vvm. In turn, varying the stirring rate from 200 to 600 rpm, at 0.5 vvm of aeration rate, the OTR value augments by ten times. This lower improvement in OTR with higher aeration rate as compared to the agitation rate was similar for all values of pressure. For all the conditions studied, the increase in agitation proved to be more efficient for OTR enhancement than the increase in aeration. This behavior is in



**Figure 2.** Experimental OTR values under various experimental conditions and aeration rates of (A) 0.5 vvm, (B) 1 vvm, and (C) 2 vvm.

agreement with the results of Amaral et al. [10], Gomes et al. [19], Gómez-Díaz and Navaza [20], Juárez and Orejas [21], and Chen et al. [22] who showed that  $k_L a$  depends more strongly on agitation than on aeration rates for STR.

Impeller speed is the major factor that affects  $k_L a$  values of a stirred-tank bioreactor as it determines the overall power dissipation for any specific impeller design. This effect was attributed to the rapid breakage of the gas bubbles into smaller sizes with higher impeller speed, and consequently enhancement in the gas-liquid interfacial area for mass transfer [1]. Raising the stirring rate from 200 to 600 rpm led to an improvement of the OTR value more significantly at 1 bar air pressure and

0.5 vvm aeration. A 10-fold improvement in OTR was observed by changing the stirring rate from 200 to 600 rpm, at 1 bar and 0.5 vvm. Comparatively, only 4- and 3-fold improvements were found at 1 bar and 2 vvm and with 5 bar and 2 vvm, respectively. It seems that the stirring rate effect is more significant at low  $k_La$  values.

Independently of the stirring and aeration rates tested, the increased air pressure from 1 to 5 bar led to improved OTR values. This effect was even more pronounced at lower stirring and aeration rates. A 7.1-fold improvement of the OTR value at 0.5 vvm and 200 rpm was achieved when air pressure varied from 1 to 5 bar, whereas at 2 vvm and 600 rpm the OTR at 1 bar was four times higher than at 5 bar. In the experiment conducted at 2 vvm aeration rate, the same improvement in oxygen mass transfer (4.3-fold) was achieved with the rise of air pressure from 1 to 5 bar and with the increase of the stirring rate from 200 to 600 rpm. This result proves that the increased air pressure is an alternative to the stirring rate increase which is particularly important for high-cell-density cultures and when the cells are sensitive to shear stress that limits the increase of stirring. Belo and Mota [23] observed that *Escherichia coli* TB1 cells proved to be more sensitive to high shear stress caused by stirring than to air total pressures up to 4 bar.

The oxygen transfer rate raise promoted by the increased air pressure inside the bioreactor is based on the fact that the equilibrium oxygen solubility in the nutrient broth increases linearly with the total air pressure according to Henry's law. Although oxygen-enriched air can achieve the same result, it is costly and requires special handling [24].

Other authors have reported the enhancement in OTR values due to the increased air pressure inside the bioreactor. Yang and Wang [24] found a 2.5-fold increase in OTR by applying air pressures from 1.06 to 2.72 bar. Knoll et al. [6] reported that the oxygen transfer capacity, energy efficiency, and cost efficiency of oxygen transfer can be greatly enhanced by employing elevated reactor pressures of up to 11 bar.

### 3.2 $k_La$ Modeling

The determination of  $k_La$  in bioreactors is essential to establish aeration efficiency and to quantify the effects of the operating variables on the provision of oxygen. In order to predict bioreactor performance when using models that account for the effect of the increased air pressure, an empirical correlation (Eq. (10)) for the  $k_La$  in a pressurized bioreactor was proposed. Using the experimental data obtained in experiments with increased air pressure up to 5 bar, aeration rates ranging from 0.5 to 2 vvm, and stirring rates from 200 to 600 rpm, the values of  $\alpha$ ,  $\delta$ ,  $\beta$ , and  $\gamma$  coefficients from Eq. (10) were estimated as indicated in Eq. (11):

$$k_La = 535 \left( \frac{P_g}{V} \right)^{0.70} (v_s)^{0.48} P^{-0.13} \quad (11)$$

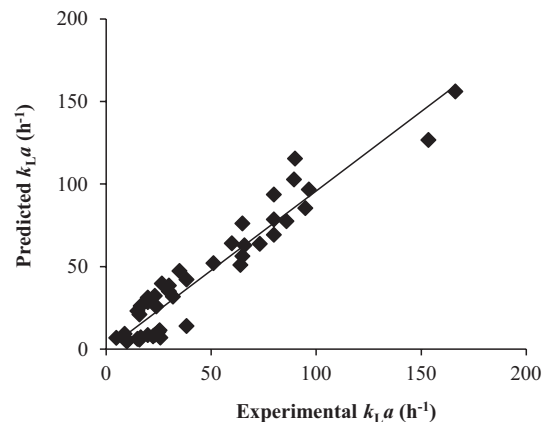
From Eq. (11) it can be concluded that the measured  $k_La$  increases according to the specific power input,  $\left( \frac{P_g}{V} \right)$ , to the power of 0.70, to the superficial gas velocity,  $v_s$ , to the power of 0.48, and decreases with the air pressure,  $P$ , to the power of (-0.13).

Knoll et al. [6] found similar values for coefficients of  $\left( \frac{P_g}{V} \right)$  and  $v_s$ , respectively, of 0.74 and 0.42. However, the authors did not take into account the air pressure in the  $k_La$  mathematical correlation. Zhang et al. [25] proposed a correlation to predict the volumetric mass transfer coefficient in a rotating-drum bioreactor, and the coefficient of volumetric aeration rate ( $Q_G$ ) was 0.6 and for the specific power consumption the coefficient is a function of  $Q_G$ .

The values of coefficients in Eq. (11) indicate that the  $k_La$  dependence was higher on the specific power input than the superficial gas velocity, once the coefficient of  $v_s$  was lower than the coefficient of  $\left( \frac{P_g}{V} \right)$ . The rise of total air pressure had a small negative effect on  $k_La$ , as demonstrated by the coefficient of  $P$ . This means that the increase of air pressure slightly decreases the volumetric mass transfer coefficient. Belo et al. [26] also reported that raising oxygen solubility through the increase in total air pressure enhanced OTR in the pressurized bioreactor and decreased  $k_La$ . This effect can be attributed to the use of a constant gas flow rate (measured under standard conditions) which in fact led to a decrease of the true gas flow rate inside the bioreactor with pressure. This causes a reduction in the number of air bubbles and consequently in the gas holdup [27]. Thus, the total surface area for mass transfer decreases with pressure, in spite of the specific interfacial area increasing with pressure due to the air bubbles' compression. The global effect of pressure in  $k_La$  obtained in this work was more influenced by the negative effect on gas holdup than by the positive effect of the air bubble compression. Moreover, it is accepted that the liquid mass transfer coefficient  $k_L$  is independent of the headspace pressure  $P$ , and the observed pressure effects have to be attributed to changes in the interfacial area  $a$  [28].

Yang and Wang [24] observed that the bioreactor pressurization up to 2.72 bar had little effect on  $k_La$ . Maier et al. [29] reported that the  $k_La$  values in a stirred-tank reactor remain constant irrespective of the reactor pressure if the superficial gas velocity is kept constant.

In Fig. 3, predicted versus experimental  $k_La$  values are plotted with a deviation of 4% of a unitary slope, e.g.,  $k_{L,a}^{\text{predicted}} =$



**Figure 3.** Correlation between the experimental and predicted  $k_La$  values using Eq. (11) with estimated parameters for increased air pressure up to 5 bar, stirring rates from 200 to 600 rpm, and aeration rates from 0.5 to 2 vvm.

0.96  $k_L a_{\text{experimental}}$  which indicates a good approximation between real  $k_L a$  values and the values calculated by the correlation, despite the dispersion of the values ( $R^2 = 0.911$ ), particularly for low  $k_L a$  values as obtained in the experiments conducted at 200 rpm stirring rate.

## 4 Conclusions

Mass transfer between gas and liquid phases in stirred-tank reactors is a key process in the chemical and biochemical industry. Thus, optimization of the bioreactor performance concerning the oxygen mass transfer requirement is a crucial task in industrial bioprocesses. Herein, the effects of increased air pressure on the oxygen transfer rate were investigated. The application of increased air pressure up to 5 bar proved to be a successful means to improve OTR, i.e., this pressure can serve as an alternative to avoid shear stress caused by the increased stirring rates that might be harmful to cells.

An empirical correlation to predict the  $k_L a$  value as a function of pressure, power input, and superficial gas velocity was established. The  $k_L a$  increase turned out to be higher with the rise of specific power input than with the superficial gas velocity whereas a higher total air pressure had only a small negative effect on  $k_L a$ . The proposed correlation for  $k_L a$  prediction could be valuable for further work on the development of strategies for optimization and scale-up of processes where oxygen transfer is a limiting factor.

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## Symbols used

$a$	[ $\text{m}^{-1}$ ]	specific interfacial area
$c$	[–]	constant dependent on the impeller used
$C$	[ $\text{mg}_{\text{O}_2} \text{L}^{-1}$ ]	dissolved oxygen concentration in the liquid
$C^*$	[ $\text{mg}_{\text{O}_2} \text{L}^{-1}$ ]	solubility of oxygen in the liquid
$D$	[m]	bioreactor vessel diameter
$D_i$	[m]	impeller diameter
$F_g$	[ $\text{m}^3 \text{s}^{-1}$ ]	volumetric gas flow rate
$H$	[m]	bioreactor vessel height
$k_L$	[ $\text{m s}^{-1}$ ]	liquid side mass transfer coefficient
$k_L a$	[ $\text{h}^{-1}$ ]	oxygen volumetric mass transfer coefficient
$K_T$	[–]	constant dependent on the impeller used
$N$	[rpm]	agitation rate
$N_p$	[–]	power number
OTR	[ $\text{mg}_{\text{O}_2} \text{L}^{-1} \text{h}^{-1}$ ]	oxygen mass transfer rate

$P$	[bar]	absolute pressure
$P_g$	[W]	power input to the aerated system
$P'_g$	[W]	power input to the non-aerated system
$Q_G$	[vvm]	volumetric aeration rate
$Re$	[–]	Reynolds number
$V$	[ $\text{m}^3$ ]	working volume
$V_I$	[mL]	iodine solution volume
$V_s$	[mL]	sample volume

## Greek letters

$\alpha, \beta, \delta, \gamma$	[–]	non-linear models parameters to be fitted
$\nu$	[ $\text{kg m}^{-1} \text{s}^{-1}$ ]	liquid viscosity
$\nu_s$	[ $\text{m s}^{-1}$ ]	superficial gas velocity
$\rho$	[ $\text{kg m}^{-3}$ ]	liquid density

## References

- [1] S. Suresh, V. C. Srivastava, I. M. Mishra, *J. Chem. Technol. Biotechnol.* **2009**, *84*, 1091.
- [2] F. Garcia-Ochoa, E. Gomez, *Biotechnol. Adv.* **2009**, *27*, 153.
- [3] C. G. Sinclair, *Biotechnol. Lett.* **1984**, *6* (2), 65.
- [4] N. P. Cheremisinoff, R. Gupta, in *Handbook of Fluids in Motion* (Eds: N. P. Cheremisinoff, R. Gupta), Ann Arbor Science, Ann Arbor, MI **1983**.
- [5] M. Lopes, N. Gomes, C. Gonçalves, M. A. Z. Coelho, M. Mota, I. Belo, *Letts. Appl. Microbiol.* **2008**, *46*, 255.
- [6] A. Knoll, B. Maie, H. Tscherrig, J. Buchs, *Adv. Biochem. Eng./Biotechnol.* **2005**, *92*, 77.
- [7] M. Maier, M. Losen, J. Buchs, *Biochem. Eng. J.* **2004**, *17*, 155.
- [8] I. Belo, R. Pinheiro, M. Mota, *Biotechnol. Prog.* **2003**, *19*, 665.
- [9] B. Sonnleitner, U. Hahnemann, *J. Biotechnol.* **1994**, *38*, 63.
- [10] P. F. F. Amaral, M. G. Freire, M. H. Rocha-Leão, I. M. Marrucho, J. A. P. Coutinho, M. A. Z. Coelho, *Biotechnol. Bioeng.* **2008**, *99* (3), 588.
- [11] D. R. Nielsen, A. J. Daugulis, P. J. McLellan, *Biotechnol. Bioeng.* **2003**, *83*, 735.
- [12] M. Novak, V. Klekner, *Biotechnol. Tech.* **1988**, *2*, 243.
- [13] C. M. Cooper, G. A. Fernstorm, S. A. Miller, *Ind. Eng. Chem.* **1944**, *36*, 504.
- [14] M. Lopes, N. Gomes, M. Mota, I. Belo, *Appl. Biochem. Biotechnol.* **2009**, *159*, 46.
- [15] A. Knoll, S. Bartsch, B. Husemann, P. Engel, K. Schroer, B. Ribeiro, C. Stöckmann, J. Seletzky, J. Büchs, *J. Biotechnol.* **2007**, *132*, 167.
- [16] R. Pinheiro, I. Belo, M. Mota, *Letts. Appl. Microbiol.* **2003**, *37*, 438.
- [17] J. M. Prausnitz, R. N. Lichtenthaler, E. G. Azevedo, *Molecular Thermodynamics of Fluid-Phase Equilibria*, 2nd ed., Prentice-Hall, Englewood Cliffs, NJ **1986**.
- [18] I. Belo, A. García-Abuín, D. Gómez-Díaz, J. M. Navaza, I. Vidal-Tato, *Chem. Eng. Technol.* **2011**, *34* (11), 1790.
- [19] N. Gomes, M. Aguedo, J. Teixeira, I. Belo, *Biochem. Eng. J.* **2007**, *35*, 380.
- [20] D. Gómez-Díaz, J. M. Navaza, *Chem. Eng. Technol.* **2003**, *26* (10), 1068.

- [21] P. Juárez, J. Orejas, *Lat. Am. Appl. Res.* **2001**, *31*, 433.
- [22] J.-Y. Chen, C.-M. Wen, T.-L. Chen, *Biotechnol. Bioeng.* **1999**, *62* (3), 311.
- [23] I. Belo, M. Mota, *Bioprocess. Eng.* **1998**, *18*, 451.
- [24] J.-D. Yang, N. S. Wang, *Biotechnol. Prog.* **1992**, *8*, 244.
- [25] Q. Zhang, Z. Whang, S. Wen, G. Liu, X. Wu, W. Cong, *Chem. Eng. Technol.* **2012**, *35* (10), 1842.
- [26] I. Belo, R. Pinheiro, M. Mota, *Appl. Microbiol. Biotechnol.* **2000**, *53*, 517.
- [27] D. Gómez-Díaz, N. Gomes, J. A. Teixeira, I. Belo, *Chem. Eng. J.* **2009**, *152*, 354.
- [28] M. Oyevaar, R. Bos, R. Westerterp, *Chem. Eng. Sci.* **1991**, *46* (5/6), 1217.
- [29] B. Maier, C. Dietrich, J. Büchs, *Trans. Inst. Chem. Eng.* **2001**, *79* (C), 107.