International Journal of Chemical, Materials and Biomolecular Sciences
ISSN: 2415-6620
Vol:7, No:11, 2013

Prediction of Oxygen Transfer and Gas Hold-Up in Pneumatic Bioreactors Containing Viscous Newtonian Fluids

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***Abstract*—**Pneumatic reactors have been widely employed in various sectors of the chemical industry, especially where are required high heat and mass transfer rates. This study aimed to obtain correlations that allow the prediction of gas hold-up (ε) and volumetric oxygen transfer coefficient (k_La), and compare these values, for three models of pneumatic reactors on two scales utilizing Newtonian fluids. Values of k_La were obtained using the dynamic pressure-step method, while ε was used for a new proposed measure. Comparing the three models of reactors studied, it was observed that the mass transfer was superior to draft-tube airlift, reaching ε of 0.173 and k_La of 0.00904s^{-1}. All correlations showed good fit to the experimental data ($R^2 \geq 94\%$), and comparisons with correlations from the literature demonstrate the need for further similar studies due to shortage of data available, mainly for airlift reactors and high viscosity fluids.

***Keywords*—**Bubble column, internal loop airlift, gas hold-up, k_La.

I. Introduction

The widespread development of biotechnology has impacted diverse sectors of the economy over the last few years. Most affected are the agricultural, fine chemical, food processing, and pharmaceutical industries, where different technologies are used to cultivate cells, tissues, and microorganisms [1].

In aerobic microbial bioprocesses, oxygen is essential for the growth and maintenance of the microorganisms, and for product synthesis. However, due to its low solubility in broths, which are usually viscous aqueous media, oxygen must be provided continuously during the process, as a result of which knowledge of oxygen transfer parameters, such as gas hold-up (ε_G) and volumetric oxygen transfer coefficient (k_La) are needed for bioreactor design and scale-up [2].

Among the existing models of bioreactors, pneumatic reactors are becoming increasingly popular in bioprocesses, due to their advantages of simple mechanical design, low power consumption, low maintenance cost, high homogenization capacity, intense mixing, and relative high mass transfer efficiency compared stirred tank reactors [3]-[6].

For this reason, there has been extensive investigation gas-liquid mass transfer in pneumatic reactors, aimed principally at obtaining empirical correlations to enable prediction of oxygen transfer parameters, including the gas hold-up and k_La [7]-[10]. Such correlations are of great importance in order to be able to expand the industrial applications of these of these bioreactors, due to the complexity of the oxygen transfer phenomena and the use of the correlations in bioreactor scale-up. However, this field is not yet fully evolved, and there are few studies that have significant physical properties influencing mass transfer, such as viscosity, which is the focus of the present work.

A further consideration is that correlations based on dimensional analysis can enable accurate estimation of the parameters for similar systems with different dimensions. However, despite this advantage, few studies have considered this approach [11], [12], [24], [25], [29], [30].

The literature contains many studies in which the mass transfer of bubble column reactors has been evaluated, but there are few reports related to airlift reactors [12]-[16], particularly the split-cylinder airlift design. Comparative studies of different model and scale of bioreactors, as presented in the present work, are practically nonexistent [17], [18]. Moreover, most of the studies reporting correlations that describe such systems used water as the fluid, and therefore did not take into account the effect of viscosity on mass transfer.

The objective of the present study was therefore to conduct an investigation of oxygen transfer in Newtonian fluids with high viscosity, making the comparison between the k_La and gas hold-up values, obtained for three models of pneumatic reactors: bubble column (BC), draft-tube internal loop airlift (DTA) and split-cylinder internal loop airlift (SCA), in two volumes of work: 5 and 10 L, and also obtain general dimensionless correlations for predicting the k_La and gas hold-up for these reactors.

II. Materials and Methods

A. Equipment

The three types of pneumatic reactor of similar geometries were constructed of stainless steel, except for the gas-liquid separator, which was made of glass. The fluid was mixed by air injection through perforated pipe spargers (0.5 mm diameter). The equipment used is illustrated schematically in Fig. 1, and the geometrical relationships for the three models and the two different bioreactor scales are provided in Table I.

B. Fluids

Three glycerol solutions were used as viscous Newtonian

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fluids. The dynamic viscosities of the fluids at 30 ºC were determined with the aid of a digital concentric-cylinder rheometer (Brookfield Engineering Laboratories, Model LV-DVIII+). Other physical properties, such as oxygen diffusivity, density, and surface tension were obtained from the literature. Five specific air flow rates (1 to 5 vvm) were used. In all experiments, the reactor temperature was maintained at 30 ºC. The physical properties of the solutions are given in Table II.

TABLE I
DIMENSIONS OF THE PNEUMATIC REACTORS USED IN THIS STUDY

Dimensions	Bubble column		Airlift			
			Draft-tube		Split-cylinder	
	5 L	10 L	5 L	10 L	5 L	10 L
H_1 (mm)	-	-	45	55	45	55
H_2 (mm)	-	-	55	45	55	45
H_3 (mm)	-	-	350	450	350	450
H_4 (mm)	450	550	450	550	450	550
H_5 (mm)	600	700	600	700	600	700
De_1 (mm)	125	160	125	160	125	160
De_2 (mm)	135	170	135	170	135	170
Di_1 (mm)	-	-	85	105	-	-
Di_2 (mm)	-	-	75	95	-	-
L (mm)	-	-	-	-	124	158
H_4/De_1	3.60	3.44	3.60	3.44	3.60	3.44
$A_D.A_S^{-1}$	-	-	1.78	1.84	1.38	1.38
$Di_2.De_1^{-1}$	-	-	0.60	0.59	-	-

TABLE II
PHYSICAL PROPERTIES OF NEWTONIAN FLUIDS AT 30ºC

Fluid	μ_L (Pa.s)	ρ_L (kg.m^{-3})	$D_L\times10^9$ (m^2.s^{-1})	σ (kg.s^{-2})
G20	0.020	1189.02	0.280	0.0676
G25	0.025	1199.04	0.246	0.0673
G30	0.030	1207.23	0.222	0.0670

C. Gas Hold-Up

According to the methodology described by Chisti (1989), gas hold-up can be determined by measuring the heights of the gas-free liquid and the gas-liquid dispersion, and then applying (1).

However, due to large fluctuations of the fluid at the time of aeration, such measurements are very imprecise. We therefore propose an adaptation of the conventional method of measurement. A transparent glass tube connected to a positive displacement pump was used. The tube was inserted through the top of the reactor until suction of the liquid started. At this moment, the pump was turned off, and the immersion height of the tube in the reactor was measured. This procedure was performed at four different points in the reactor in order to minimize the error due to oscillations of the liquid. The immersion height of the tube was then subtracted from the total height of the reactor, and the average of the four points was defined as the height of the gas-liquid dispersion, h_D. The same procedure was followed to obtain the height of the gas-free liquid, h_L.

$$\varepsilon_G = \frac{h_D - h_L}{h_D} \qquad (1)$$

D. Volumetric Oxygen Transfer Coefficient (k_La)

Values of k_La were determined under different operational conditions using a dynamic pressure-step method [20]. In this procedure, an increase in the oxygen concentration was obtained by means of a step pressure increase of approximately 15 kPa inside the vessel, achieved by partially closing a control valve to restrict the air flow rate. The absolute pressure at the head space was measured using a digital manometer (T&S Equipamentos Eletrônicos, Model SC990).

A sterilizable amperometric electrode (Mettler-Toledo, Model InPro 6800), linked to a silicone membrane (Mettler-Toledo, Model InProT96) and mounted on the top section above the riser, was used to measure the change in oxygen concentration over time.

The oxygen concentration signal was recorded by a data acquisition system at 1 s intervals, until the dissolved oxygen concentration reached saturation. Assuming that the probe had a first-order response and that the liquid phase was perfectly mixed, and taking into account the delay of the electrode, the volumetric oxygen transfer coefficient could be calculated by fitting (2) to the experimental data by nonlinear regression.

$$C_e = C_{e0}.e^{-k_e.(t-t_0)} + C_S.\left(1 - e^{-k_e.(t-t_0)}\right) + \frac{k_e.(C_S - C_{e0})}{k_e - k_La}.\left(e^{-k_e.(t-t_0)} - e^{-k_La.(t-t_0)}\right) \qquad (2)$$

In (2), Ce_0 is the signal of the electrode under the initial conditions, when $t = t_0$; C_s is the saturation dissolved oxygen concentration in the liquid phase, and k_e is the time constant of the oxygen probe, calculated from the inverse of the response time (t_e).

To determine t_e, the oxygen probe was first placed in a nitrogen atmosphere until it reached a value of zero. At this moment, the electrode was removed from the controlled atmosphere and entered into contact with atmospheric air. The response time of the electrode was determined as the time at which the signal reached 63.2% of its maximum value. All the values of k_La were obtained in duplicate.

III. RESULTS AND DISCUSSION

Global gas hold-up and volumetric oxygen transfer coefficient values were obtained for BC, DTA and SCA reactors, and the influence of reactor volume, air flow rate and viscosity on these parameters are evaluated.

Gas hold-up values ranged from 0.029 to 0.173, with the airlift bioreactors being most efficient. The values obtained were 0.037 to 0.173 (DTA) and 0.035 to 0.166 (SCA). For all reactor types the 10 L volume provided higher ε_G values, with the influence of reactor size being lowest for the bubble column reactor.

According to [19], a simple correlation that can predict gas hold-up values, taking into account air flow rate (U_{GR}) and viscosity (μ_L), may be expressed as:

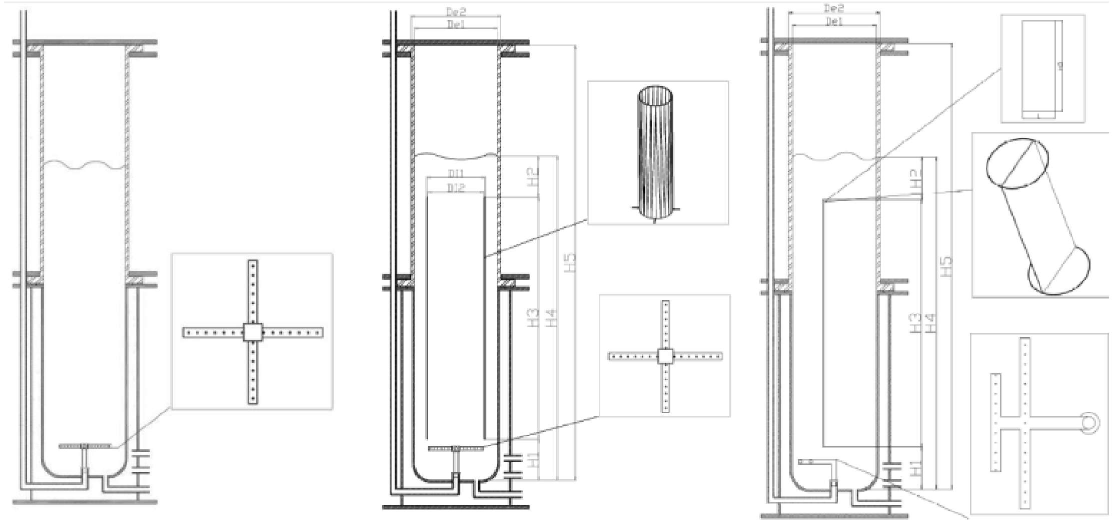

Fig. 1 Geometrical configurations of the bioreactors and spargers

$$\varepsilon_G = \alpha \cdot U_{GR}^{\beta} \cdot \mu_L{}^{\delta} \quad (3)$$

Equation (3) was fitted to the experimental gas hold-up data. The parameter values obtained after fitting are presented in Table III.

For all three bioreactors, ε_G was strongly influenced by the variables studied. There was a positive effect with respect to U_{GR} and a negative effect with respect to μ_L, with the BC reactor being most sensitive to these variables. In all cases, the experimental data showed a good fit to the proposed models, with R^2 values exceeding 95%.

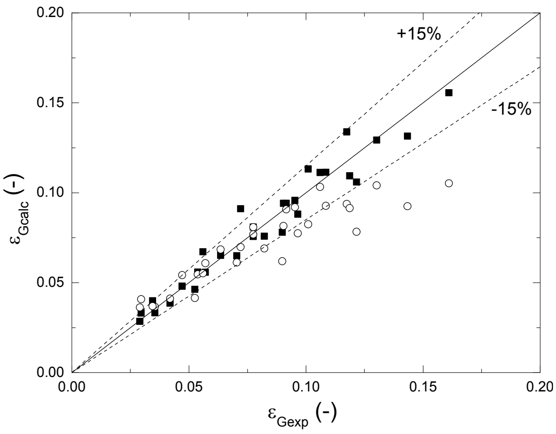

Fig. 2 Comparison between calculated (ε_{Gcalc}) (3) and experimental (ε_{Gexp}) data of gas hold-up for BC reactor: (■) this study, (○) [16]

Fig. 2 shows a comparison between the experimental data of global gas hold-up (ε_{Gexp}) and those predicted by (3) (ε_{Gcalc}) for the BC reactor. Comparison with data obtained from correlations published by other authors (Table VII) also can be seen on Fig. 2. Few of the experimental data points exceeded a deviation of 15%. There was good agreement with the data of [16], which evaluated gas hold-up in a bubble column reactor using Newtonian fluids with viscosities of up to 17 cp. Working conditions that were similar to those employed in the present study may have contributed to the observed similarities.

TABLE III
PARAMETERS OBTAINED FOR GAS HOLD-UP CORRELATION (3) FOR EACH MODEL OF REACTOR

Reactor	Parameters			R^2
	α	β	δ	
BC	0.067±0.029	0.752±0.045	-0.826±0.107	0.95
DTA	0.039±0.011	0.643±0.029	-0.716±0.074	0.97
SCA	0.136±0.044	0.707±0.034	-0.452±0.083	0.96

The experimental gas hold-up data for the DTA reactor showed excellent correlation with the proposed model, with a deviation less than 15% (Fig. 3).

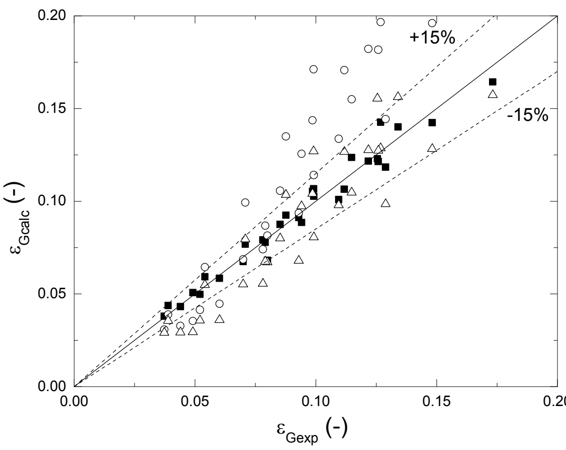

Fig. 3 Comparison between calculated (ε_{Gcalc}) (3) and experimental (ε_{Gexp}) data of gas hold-up for DTA reactor: (■) this study, (○) [21] and (△) [11]

The literature reports few studies concerning prediction of gas hold-up values for DTA reactors. Studies of the relationship between this parameter and physical properties of the fluid are even scarcer. In the work described in [11] were evaluated different bench reactor sizes, as well as many fluid viscosities, and the correlations obtained were similar to those observed here, with few points exceeding a deviation of 15%. In other work [21], good correlation was only obtained for low air flow rates.

Comparisons between calculated and experimental gas hold-up data in SCA can be seen in Fig. 4. A good correlation between the proposed model in this study and the experimental data can be evidenced, since the points did not exceed ±15% deviation.

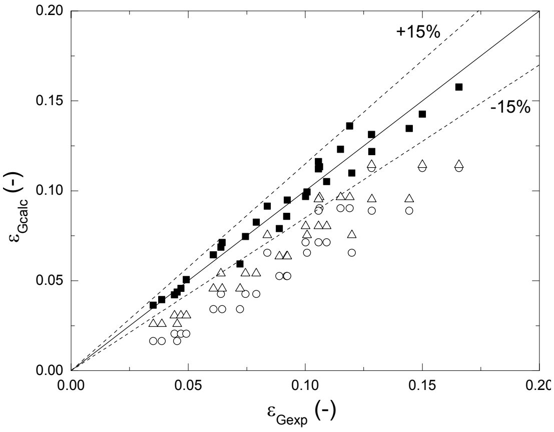

Fig. 4 Comparison between calculated (ε_{Gcalc}) (3) and experimental (ε_{Gexp}) data of gas hold-up for SCA reactor: (■) this study, (○) [27] and (△) [28]

For the SCA reactor, it can be observed from Fig. 4 that the models were generally unable to satisfactorily predict the gas hold-up. Some of the few models in the literature were proposed by [27] and [28]. However, in neither case the influence of viscosity was taken into consideration in the gas hold-up prediction model.

Dimensional analysis of correlations can result in more complex models with better fit to the experimental data, with the advantage of allowing evaluation of parameters independently of the reactor scale. General correlation based on dimensionless groups for the prediction of gas hold-up of the three models of bioreactors were proposed taking into account the influence of the Froude number (Fr), Schmidt number (Sc), Bond number (Bo), and Galilei number (Ga). The general correlation is expressed by (4).

$$\varepsilon_G = \alpha \, . \, Fr^{\beta} . \, Sc^{\delta} . \, Bo^{\theta} . \, Ga^{\lambda} \quad (4)$$

Froude (5), Schmidt (6), Bond (7) and Galilei number (8) are defined by:

$$Fr = \frac{U_{GR}}{\sqrt{g \, . \, D_e}} \quad (5)$$

$$Sc = \frac{\mu_L}{\rho_L \, . \, D_L} \quad (6)$$

$$Bo = \frac{g \, . \, \rho_L \, . \, D_e^2}{\sigma} \quad (7)$$

$$Ga = \frac{g \, . \, \rho_L^2 . \, D_e^3}{\mu_L^2} \quad (8)$$

where, U_{GR} is the superficial gas velocity in the riser (m.s^{-1}), g is the gravitational acceleration (m.s^{-2}), D_e is the equivalent diameter of the system (m), μ_L is the dynamic viscosity (Pa.s), ρ_L is the density of liquid (kg.m^{-3}), D_L is the oxygen diffusivity coefficient (m^2.s^{-1}), and σ is the surface tension of liquid (kg.s^{-2}).

Table IV shows the parameters values of (4) fitted experimental data of the three bioreactor models considering both scales.

TABLE IV
PARAMETERS OBTAINED FOR GAS HOLD-UP CORRELATION GIVEN (4) FOR EACH MODEL OF REACTOR

Reactor	Parameters					R^2
	α (×10⁴)	β	δ	θ	λ	
BC	1.7	0.710	0.184	-0.414	0.573	0.97
DTA	1.3	0.614	0.223	-0.444	0.537	0.99
SCA	1.6	0.692	0.334	-0.484	0.488	0.96

As expected, the model based on dimensionless numbers resulted in a better fit to the experimental data, resulting in an R^2 greater than 96%. The influence of the dimensionless groups was of the same order of magnitude as the gas hold-up values, with parameter values that were nearly identical for the different bioreactor models. It was further noted that the Froude number, which concerns the ratio between the inertial forces and the force of gravity [11], had the greatest influence on the outcome; this number.

The comparisons between experimental and the calculated data for the dimensionless gas hold-up model are shown in Figs. 5, 6, and 7 for BC, DTA and SCA, respectively. For comparison, other similar correlations proposed in the literature (Table VII) were also used to predict gas hold-up in the pneumatic bioreactors.

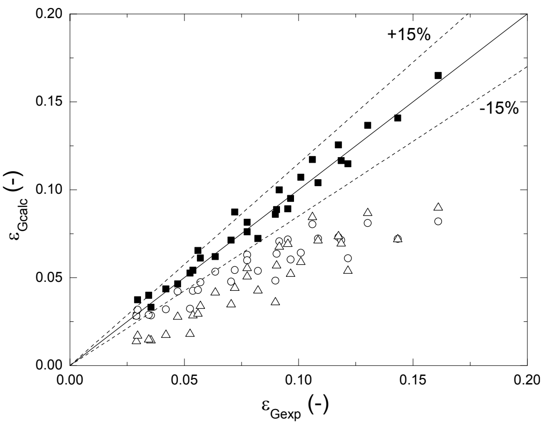

Fig. 5 Comparison between calculated (ε_{Gcalc}) (4) and experimental (ε_{Gexp}) data of gas hold-up for BC reactor: (■) this study, (○) [16] and (△) [24]

Fig. 5 reveals a good correlation between the experimental data and the dimensionless model proposed for the gas hold-up prediction, with few points exceeding ±15% of deviation. However, the same behavior cannot be observed for the correlations found in the literature. For example, the dimensionless correlation proposed by [24], which uses dimensionless groups similar to those used in the present study have an influence of the Froude number (β=1.0) greater than found the present model (β=0.710).

Furthermore, in the present work, there is a negative influence of the Bond number (θ=-0.414). The opposite was obtained by [24] for this dimensionless group, obtaining a low but positive effect of Bond number on the gas hold-up (θ=0.125).

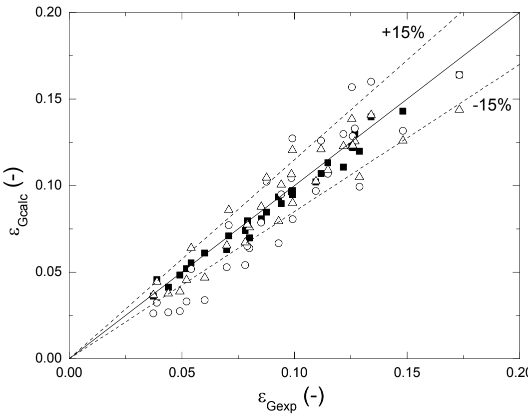

Fig. 6 Comparison between calculated (ε_{Gcalc}) (4) and experimental (ε_{Gexp}) data of gas hold-up for DTA reactor: (■) this study, (○) [11] and (△) [26]

As noted earlier, the proposed model to calculate the gas hold-up in DTA reactor composed of dimensionless numbers showed a good fit to the experimental data, as evidenced in Fig. 6.

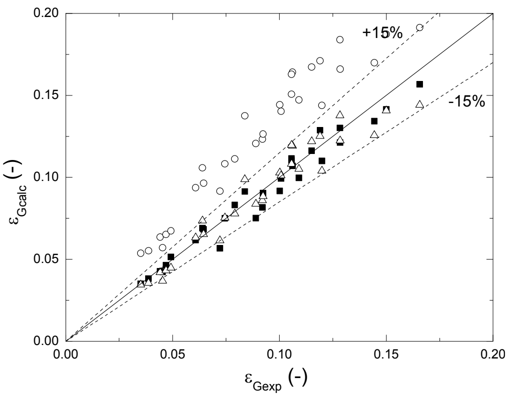

Fig. 7 Comparison between calculated (ε_{Gcalc}) (4) and experimental (ε_{Gexp}) data of gas hold-up for SCA reactor: (■) this study, (○) [29] and (△) [30]

The correlations proposed by [26] and [11] were able to provide satisfactory fits to the experimental data. In the model proposed by [11], there was a similar influence of the dimensionless groups on the gas hold-up values. There was greater influence of the Schmidt number (δ=0.471) and Froude number (β=0.982) in the correlation obtained by these authors, similar to that obtained in the present study.

In the study described in [30], the authors evaluated the mass transfer in a SCA reactor with 21 L working volume and different types and concentrations of alcohols. It was noted that the correlation proposed by the authors was similar to the model obtained in this study, showing excellent agreement with the experimental data of gas hold-up with deviation lower than 15%. For SCA, the Bond number had a negative influence on ε_G in the present work. However, in the literature works of [29] and [30] Bond number presented a positive effect on ε_G, θ=0.18 and θ=0.13, respectively.

Volumetric oxygen transfer coefficient (k_La) values ranged from 0.00177 to 0.00904 s^{-1} and similar behavior of the k_La was observed with respect to ε_G, where the highest values were obtained for the DTA reactor of 10 L working volume among the reactor models studied. Furthermore, analogously to gas hold-up was observed an increase of k_La with increasing of air flow rate, and a reduction of k_La with increasing of viscosity.

Data of k_La were used to allow calculation of the experimental Sherwood number (Sh), which is defined by (9).

$$Sh = \frac{k_La \, . \, D_e}{D_L} \tag{9}$$

where D_e is the equivalent diameter of the system (m) and D_L is the oxygen diffusivity coefficient ($m^2.s^{-1}$).

The Sherwood number may be correlated with experimental values of k_La and therefore with the dimensionless group used to provide a more complete analysis of the mass transfer phenomenon, together with other dimensionless groups such as the Froude (Fr), Schmidt (Sc), Bond (Bo), and Galilei (Ga) number. The correlation given by (10) was fitted to experimental data.

$$Sh = \alpha \, . \, Fr^{\beta} . \, Sc^{\delta} . \, Bo^{\theta} . \, Ga^{\lambda} . \, \varepsilon_G^{\varphi} \tag{10}$$

The parameters obtained by fit of (10) for the three bioreactor models are shown in Table V. Excellent correlations between the experimental and calculated data were obtained for the Sherwood number, with R^2 values exceeding 96%. As can be seen, the Schmidt number was the group with the strongest influence on the Sherwood number, especially in the case of the BC reactor. This result was expected, since these dimensionless groups are closely linked. The Sherwood number represents the correlation between convective and diffusive mass transport, while the Schmidt number concerns the relationship between the momentum and mass transport [15].

TABLE V
PARAMETERS OBTAINED FOR SHERWOOD NUMBER GIVEN (10) FOR EACH MODEL OF REACTOR

Reactor	Parameters						R^2
	α ($\times 10^4$)	β	δ	θ	λ	φ	
BC	0.5	0.255	0.921	0.089	0.756	0.453	0.97
DTA	3.2	0.287	0.786	0.163	0.709	0.463	0.97
SCA	18.1	0.133	0.729	0.318	0.580	0.616	0.97

Table VI presents the ranges of the dimensionless groups of (4) and then to the (10).

TABLE VI
THE RANGES OF THE DIMENSIONLESS GROUPS USED FOR THE FIT OF (4) AND (10) FOR THREE BIOREACTOR MODELS

Reactor	Ranges of the dimensionless groups				
	Sh*	Fr	Sc	Bo	Ga
BC	1.21×10^5 to 7.60×10^5	6.13×10^{-3} to 3.31×10^{-2}	6.01×10^4 to 1.12×10^5	2.70×10^3 to 4.52×10^3	3.10×10^7 to 1.42×10^8
DTA	4.35×10^4 to 2.91×10^5	2.20×10^{-2} to 1.22×10^{-1}	6.01×10^4 to 1.12×10^5	9.70×10^2 to 1.59×10^3	6.70×10^6 to 2.97×10^7
SCA	1.64×10^4 to 1.41×10^5	2.29×10^{-2} to 1.24×10^{-1}	6.01×10^4 to 1.12×10^5	3.97×10^2 to 8.17×10^2	1.76×10^6 to 1.09×10^7

* Sherwood number used only for fit the (10).

Excellent correlation between calculated (10) and experimental Sherwood number data can be evidenced in Figs. 8-10, where for the three bioreactor models the deviations were lower than 15%. These figures highlight the shortage of dimensionless correlations available in the literature using viscous fluids in their experiments, especially for the airlift bioreactors.

Can be seen in Fig. 8 that the correlation proposed by [25] was well adapted to the experimental data of the present work, with few points exceeding ±15% deviation, mainly for the lower air flow rate, while the correlation described by [24] remained moved away of the data for high air flow rates.

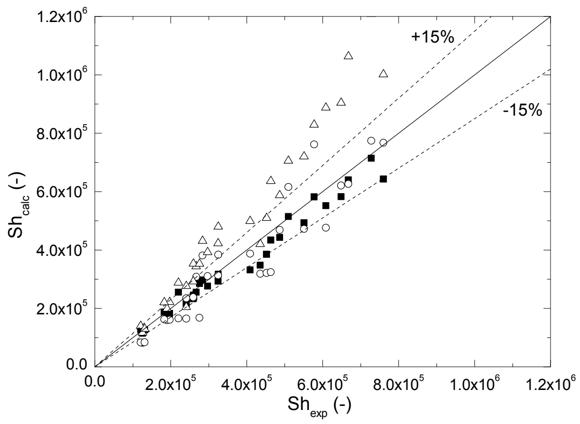

Fig. 8 Comparison between calculated (Sh_{calc}) (10) and experimental (Sh_{exp}) data of Sherwood for BC reactor: (■) this study, (○) [25] and (△) [24]

The correlation proposed by [12] showed great similarity to that obtained in the present study, where both cases, the Sherwood number was strongly influenced by the Schmidt number, with δ=0.779 and δ=0.786, respectively, similar to observed previously for the proposed model for the BC reactor.

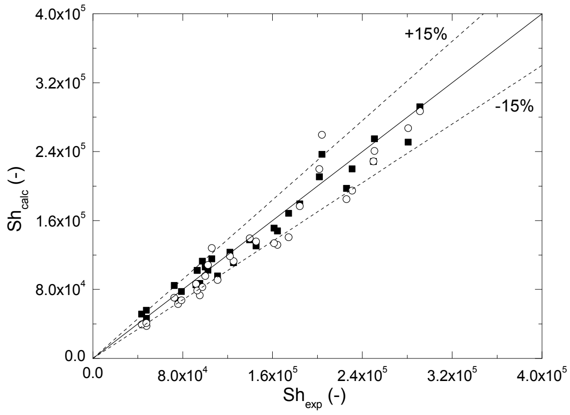

Fig. 9 Comparison between calculated (Sh_{calc}) (10) and experimental (Sh_{exp}) data of Sherwood for DTA reactor: (■) this study, (○) [12]

Finally, Fig. 10 shows a great agreement between experimental and calculated data of the Sherwood number for SCA reactor, a very important result since other models found in the literature (Table VII) was not able to express these data accurately.

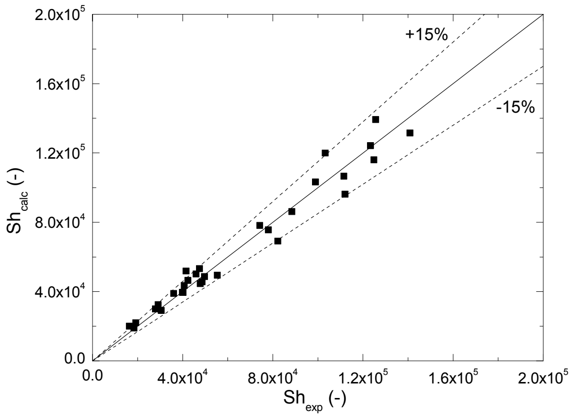

Fig. 10 Comparison between calculated (Sh_{calc}) (10) and experimental (Sh_{exp}) data of Sherwood for SCA reactor: (■) this study

IV. CONCLUSION

The main objective of this work was to investigate the influence of the air flow rate, physical properties (ρ, μ, σ and D_L) on global gas hold-up (ε_G) and volumetric oxygen transfer coefficient (k_La) in two scales of three models of pneumatic reactors widely used in bioprocesses.

Both oxygen transfer parameters were higher for the airlift reactors, in particular to the draft-tube airlift, and the 10 L reactors which exhibited the highest values in all cases.

With experimental data of ε_G and k_La was possible to obtain a simple correlation for gas hold-up taking into account the influence of viscosity and air flow rate. Also, dimensionless correlations for gas hold-up and Sherwood number were proposed, which allowed a better fit to the experimental data. In all cases, the obtained models presented excellent fittings to the experimental data with few points exceeding ±15% deviation.

After comparison of the proposed models with correlations available in the literature, it can be concluded the shortage of available correlations able of provide satisfactorily these mass transfer parameters widely used for project and scale-up of bioreactors, even for similar systems to the present study, which emphasizes the necessity for further studies regarding the transfer phenomenon in pneumatic mass bioreactors, particularly airlift reactors with high-viscosity fluids, widely found in bioprocesses.

ACKNOWLEDGMENT

Authors would like to acknowledge the financial supports by grants 2011/23807-1 and 2012/17756-8, São Paulo Research Foundation (FAPESP).

APPENDIX

TABLE VII
CORRELATIONS AVAILABLE IN THE LITERATURE FOR CALCULATING GAS HOLD-UP AND SHERWOOD NUMBER

Correlation	Ref.
Bubble Column	
$\varepsilon_G = 0.43 \,.\, U_{GR}^{0.87} \,.\, \mu_L^{-0.18}$	[22]
$\varepsilon_G = 0.21 \,.\, U_{GR}^{0.58} \,.\, D_C^{-0.18} \,.\, \mu_L^{-0.12}$	[23]
$\varepsilon_G = 0.672 \,.\, g^{-0.131} \,.\, U_{GR}^{0.578} \,.\, \rho_G^{0.062} \,.\, \rho_L^{0.069} \times \mu_G^{0.107} \,.\, \mu_L^{-0.053} \,.\, \sigma^{-0.185}$	[16]
$\frac{\varepsilon_G}{(1-\varepsilon_G)^4} = 0.2 \,.\, Bo^{0.125} \,.\, Ga^{0.5} \,.\, Fr$	[24]
$\varepsilon_G = 0.672 \,.\, \left(\frac{U_{GR} \cdot \mu_L}{\sigma}\right)^{0.578} \,.\, Mo^{-0.131} \times \left(\frac{\rho_G}{\rho_L}\right)^{0.062} \,.\, \left(\frac{\mu_G}{\mu_L}\right)^{0.107}$, where $Mo = \frac{\mu_L^4 \cdot g}{\rho_L \cdot \sigma^3}$	[16]
$Sh = 0.142 \,.\, Re^{0.875} \,.\, Sc^{0.5} \,.\, Bo^{0.6} \,.\, Fr^{0.075}$	[25]
$Sh = 0.6 \,.\, Sc^{0.5} \,.\, Bo^{0.62} \,.\, Ga^{0.31} \,.\, \varepsilon_G^{1.1}$	[24]
Draft-tube Airlift	
$\varepsilon_G = 0.0492 \,.\, U_{GR}^{1.066} \,.\, \left(\frac{\mu_L}{\rho_L}\right)^{-0.355}$	[21]
$\varepsilon_G = 0.994 \,.\, U_{GR}^{0.916} \,.\, \left(\frac{\mu_L}{\rho_L}\right)^{-0.03}$	[11]
$\frac{\varepsilon_G}{(1-\varepsilon_G)^4} = 0.16 \,.\, \left(\frac{U_{GR} \cdot \mu_L}{\sigma}\right)^{0.964} \,.\, \left(\frac{1}{Mo}\right)^{0.289} \times \left(\frac{D_i}{D_C}\right)^{-0.222} \,.\, \left(\frac{d}{D_C}\right)^{-0.0237}$	[26]
$\varepsilon_G = 1.0E^{-4} \,.\, Fr^{0.982} \,.\, Sc^{0.471} \,.\, Bo^{-0.40} \,.\, Ga^{0.42}$	[11]
$Sh = 4.6E^{-5} \,.\, Fr^{0.642} \,.\, Sc^{0.779} \,.\, Bo^{0.245} \,.\, Ga^{0.673} \,.\, \varepsilon_G^{0.2}$	[12]
Split-Cylinder Airlift	
$\varepsilon_G = 1.277 \,.\, U_{GR}^{1.06}$	[27]
$\varepsilon_G = 0.7426 \,.\, U_{GR}^{0.8167}$	[28]
$\frac{\varepsilon_G}{(1-\varepsilon_G)^4} = 0.42 \,.\, Bo^{0.18} \,.\, Ga^{0.086} \,.\, Fr$	[29]
$\frac{\varepsilon_G}{(1-\varepsilon_G)^4} = 0.18 \,.\, Bo^{0.13} \,.\, Ga^{0.1} \,.\, Fr$	[30]
$Sh = 0.025 \,.\, Re^{0.67} \,.\, Sc^{0.71} \,.\, Bo^{0.31}$	[29]
$Sh = 0.15 \,.\, Re^{0.83} \,.\, Sc^{0.5} \,.\, Bo^{0.22}$	[30]

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