Oxygen transfer, mixing time and gas holdup characterization in a hybrid bioreactor

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Abstract

Global oxygen transfer coefficient (k_La), gas holdup (ε) and mixing time were characterized in a hybrid bioreactor (mechanically agitated airlift). For the measurements water and culture medium were used under different agitation and aeration conditions. As a result, oxygen transfer and mixing performance was enhanced due to the use of mechanically agitation; however, oxygen transfer efficiency did not increase for all the agitation and aeration conditions. Correlations for gas holdup and k_La were achieved in the bubbly and coalesced bubbled flow, getting predictions within ±15% of maximum deviation. The designed reactor is shown as a novel alternative for aerobic fermentations in order to get high cell density cultures and for operations where is require to get biomass for further assays without an environment exposure. This bioreactor can be considered as a prototype for the design of larger bioreactors.

Keywords: Bioreactor, Airlift reactor, Mechanically agitated.

1.Introduction

The use of airlift reactors and mechanically stirred tanks in several bioprocesses for industrial applications is increasing (Chisti, 1989; Mohanty *et al*., 2006). Airlift reactors came out as a modified bubble column with a simple structure. They have advantages over bubble columns thanks to a higher liquid overall velocity and turbulence intensity that enhance heat and mass transfer, as well as good mixing properties at low energy consumption and lower shear stress than in mechanically stirred reactors and bubble columns (Mohanty, 2006). Airlift reactors have biochemical applications in processes which use Newtonian fluids and need moderate agitation with low shear and low oxygen transfer cost. However, concentration axial gradients of nutrients and oxygen in tall airlift reactors affect performance adversely (Chisti, 1989; Pollard *et al*., 1998). On the other hand, stirred tanks used in fermentations processes have a wide range of applications and are broadly used by their low costs, but they execute poorly in highly viscous non-Newtonian media, because of possible impeller flooding they cannot be aerated at high gas velocities and have weakly defined mixing pattern compared to airlift reactors (Chisti and Jaúregui-Haza, 2002).

The hydrodynamic reactor behavior knowledgement is necessary to understand the transport phenomena when they are being operated, in this sense many empirical and semi-empirical correlations have been proposed to estimate hydrodynamics parameters like mixing time, liquid circulation velocity, gas holdup and mass transfer (k_La) (Chisti, 1989); but not enough information is available for the reactor performance analysis and design, as well as in most of the cases only applicable to airlift reactors. General equations to evaluate these parameters are limited by the presence of variables like superficial gas velocity, gas or liquid physical properties, reactor geometry and phase separation conditions that affect the reactor performance (Mohanty, 2006).

In order to overcome some limitations of airlift and mechanically agitated reactors, hybrid reactors (mechanically agitated airlift) have been designed and characterized, showing that the oxygen transfer and mixing performance was enhanced due to mechanical agitation; also this kind of reactors can achieve higher rates of fluid circulation than conventional airlift reactors because they have a highly directional flow pattern (Chisti and Jaúregui-Haza, 2002).

Mechanically agitated airlift reactors have been studied by several authors getting good results on their performance. Chisti and Jaúregui-Haza (2002) reported the gas-liquid oxygen transfer, mixing, gas holdup and liquid circulation in a draft-tube airlift bioreactor (diameter and height were 0.755 and 3.21 m, respectively), with two Prochem® Maxflo T hydrofoil impellers and sparged in the annular zone through a perforated pipe ring; two pH probes and a dissolved oxygen electrode were located in the downcomer. Characterization was carried out using water and cellulose fiber slurries (2-4%) in sodium chloride (0.15 M); achieving an improvement in mixing and oxygen transfer capacity due to mechanical agitation. However, this agitation reduced the oxygen transfer efficiency. Pollard *et al*. (1997) studied the liquid circulation time, gas holdup and oxygen transfer performance in a concentric airlift reactor with 0.25 m^3 as total volume (internal diameter of 0.371 m and the ratio of downcomer – riser cross sectional area A_d / A_r was 0.83), agitated with a marine propeller and sparged in the annular zone through a perforated pipe, getting an increment in gas holdup and liquid circulation velocity, that is associated with the increase in oxygen transfer efficiency; unlike to the results gotten by Chisti and Jaúregui-Haza (2002).

2.Materials and methods

2.1.Reactor Size and operating conditions

The hydrodynamic characterization was made in a mechanically agitated airlift. The draft-tube was conformed by a filtration module in stainless steel with pore size of 20 μm and an irregular geometry in order to increase the filtration surface (Fig.1). The area-volume ratio was 0.036 $m^2\,L^{-1}$. Agitation was made with two Rushton turbines, the 6-bladed turbines, 0.075 m in diameter (d_i) were placed at the centerline of the reactor vessel. The vertical distance between the impellers was 0.050 m (s_c) and the lower impeller was located 0.085 m (C) from the bottom of the tank. The bioreactor vessel was rounded bottom, 0.16 m (D) in diameter and its overall height was 0.32 m (H). The draft-tube, 0.09 m in internal equivalent diameter and 0.015 m tall (h_m), was located 0.05 m (C_l) above the bottom of the tank (Fig. 2). The vessel was sparged in the concentric zone through a perforated pipe ring sparger (4 holes of 0.0015 m in diameter located on one concentric sparger) and the vertical distance between the lower turbine and the sparger was 0.035 m (C_{ts}). The ratio riser–downcomer cross sectional area (A_r/A_d) was 0.83. The static liquid height was 0.22 m (h_L) in all experiments.

Measurements were made at 30±1 ºC, using water and culture medium as test fluids. The agitation and aeration conditions for water were: stirring rates 0, 50, 100, 200, 300 and 450 rpm and superficial gas velocities referred to the riser area (U_{Gr}) 0, 0.003, 0.004, 0.006, 0.008, 0.010 and 0.012 m s^{-1}; whereas for the culture media the stirring rates were the same, the superficial gas velocities were 0, 0.004, 0.008, 0.010 and 0.012 m s^{-1}. Air was supplied at 20 psi from a compressor through a filter, pressure controller, flow control valve and rotameter. Each test was made by triplicate.

Fig. 1 Upper **(a)** and lower **(b)** cross view and axial view **(c)** of the design filtration module.

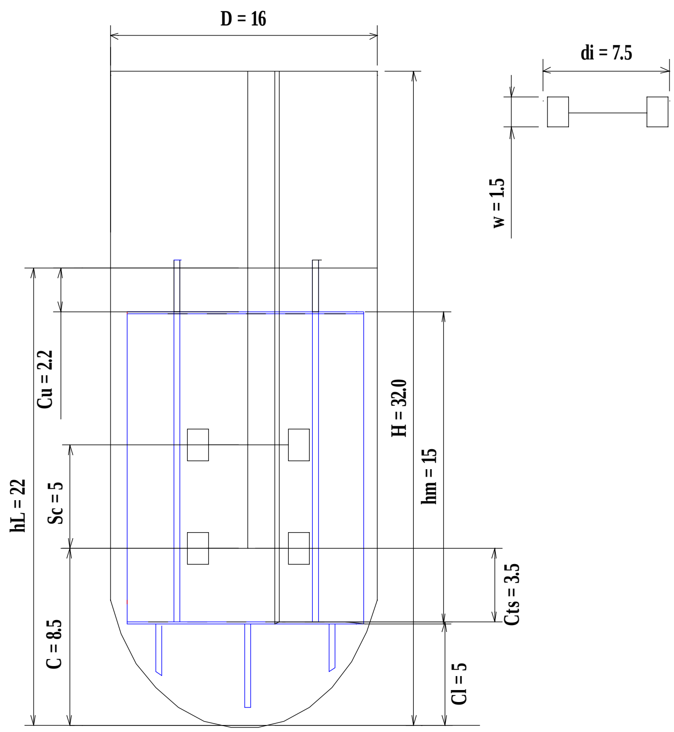

Fig. 2 Reactor dimensions (in centimeters).

2.2.Hydrodynamic characterization

2.2.1. Mixing time (t_m), gas holdup (ε), gas residence time (t_R) and average bubble rising velocity (U_b)

The mixing time was determined by the acid tracer technique (Chisti, 1989; Sánchez Mirón *et al*., 2004), as the time needed for the tracer concentration to reach 95% of its final steady-state value from the instance of tracer input. Once full the reactor with the corresponding fluid, HCl (35 % $^{w}/_{v}$) was added until pH 2 and later air was bubbled by 20 minutes (U_{Gr} = 0.0040 m s^{-1}) to remove any carbonate; later with NaOH (6 M) pH was established in 5.50 ± 0.05. The acid tracer (1 mL of HCl 35 % $^{w}/_{v}$) was added in the reactor through a port in the downcomer section. Adimentional acid concentration $[H^+]$ was estimated with Eq. 1:

$$C_T = [H^+] = \frac{[H^+]_{\text{at instant } t} - [H^+]_{\text{initial}}}{[H^+]_{\text{final}} - [H^+]_{\text{initial}}} \qquad (1)$$

Gas holdup (ε) was measured by the volume expansion method. Eq. 2. (Chisti, 1989; Chisti and Jáuregui-Haza, 2002).

$$\varepsilon = \frac{h_D - h_L}{h_D} \tag{2}$$

Where h_D is the height of the gas-liquid dispersion and h_L is the height of not gasified liquid. The residence time of the gas in dispersion (t_R), and average bubble rising velocity (U_b) in the riser, were calculated using Eqs. 3 and 4, respectively (Chisti, 1989):

$$t_R = \frac{h_L \ \varepsilon}{(1-\varepsilon)\ U_{Gr}} \tag{3}$$

$$U_b = \frac{U_{Gr}}{\varepsilon} \tag{4}$$

2.2.2.Overall gas-liquid volumetric mass transfer coefficient (k_La)

The overall gas–liquid volumetric mass transfer coefficient (k_La) was measured with the dynamic gassing-in method (Chisti, 1989; Chisti and Jáuregui-Haza, 2002). Measurements were made using a dissolved oxygen electrode located in the downcomer to 0.14 m from the top of the surface of the fluid. For each test the fluid was purged by bubbling nitrogen until reaching a dissolved oxygen concentration lower than 5% of air saturation. Later nitrogen flow was suspended, the exit of its bubbles was allowed and the air flow was established to the required condition. The increase in dissolved oxygen concentration was followed with time until the fluid became nearly saturated with oxygen (>90%). The k_La was calculated as the slope of the linear equation:

$$-\ln(1-E) = k_L a\ (t - t_0) \tag{5}$$

In Eq. 5, E is the fractional approach to equilibrium (Chisti, 1989) and can be estimated by Eq. 6:

$$E = \frac{C - C_0}{C^* - C} \tag{6}$$

Where C^* is dissolved oxygen saturation concentration, C_0 is dissolved oxygen initial concentration at time t_0 when a hydrodynamic steady-state has been reestablished (≤ 60 s) upon the beginning of aeration and C is dissolved oxygen concentration at any time t.

2.2.3.Pneumatic power (P_G) and mechanical power (P_M)

The pneumatic power due to aeration was calculated using Eq. 7 (Chisti, 1989; Chisti and Jáuregui-Haza, 2002):

$$\frac{P_G}{V_L} = \frac{\rho_L \; g \; U_{Gr}}{1 + {A_d}/{A_r}} \tag{7}$$

Where P_G is power input due to aeration, V_L culture volume, g gravitational acceleration, U_{Gr} is superficial gas velocity based on the riser cross-sectional area, A_d and A_r are the cross-sectional areas of the downcomer and riser zone, respectively.

The power input due to mechanical agitation (Eq. 8) was calculated using the power number (P_O) versus the impeller Reynolds number curves (Eq. 9) for the Rushton turbines (Casas López *et al*., 2005; Chisti and Jáuregui-Haza, 2002; Hudcova *et al*., 1989; Chisti, 1989)

$$P_M = P_O \; N^3 \; d_i^2 \; \rho_L \tag{8}$$

$$\mathrm{Re}_i = \frac{d_i^2 \; N \; \rho_L}{\mu} \tag{9}$$

Where P_M is mechanical power, d_i impeller diameter, N impeller rotational speed, ρ_L and μ are the fluid density and viscosity, respectively. The total specific power input in the fluid (P) and oxygen mass transfer efficiency (E_M), were estimated with Eqs. 10 and 11 (Chisti, 1989; Chisti and Jáuregui-Haza, 2002):

$$\frac{P}{V_L} = \frac{P_G + P_M}{V_L} \tag{10}$$

$$E_M = \frac{k_L a}{P \,/\, V_L} \tag{11}$$

3.Results and discussions

3.1.Mixing time (t_m) and residence time (t_R)

Fig. 3 shows the variation of mixing time with superficial gas velocity and agitation speeds for water and culture medium. A diminution of mixing time is observed with the increase of superficial gas velocity, however significant improvement are not appreciated at $U_{Gr} \geq 0.01$ m s^{-1}, showing an agitation rate independence. Although, at 50 r.p.m. there is an increase in the mixing time, for impeller speeds until 300 r.p.m. it decreases and further speeds do not make any significant improvement. For all U_{Gr} and agitation rates $\geq$ 300 r.p.m. the mixing time is lower than the gotten without mechanical agitation, while for agitation rates between 50 and 300 r.p.m. the same behavior is observed with $U_{Gr} \geq 0.01$ m s^{-1}.

Mixing time for water and culture medium at the same aeration and agitation conditions has not significant difference (< 5s). Consequently presence of ions and suspended solids (~1.4% $^{W}/_{V}$) in the culture medium do not have an important effect on mixing time. The variance coefficient of mixing time measurements was less than 9 %.

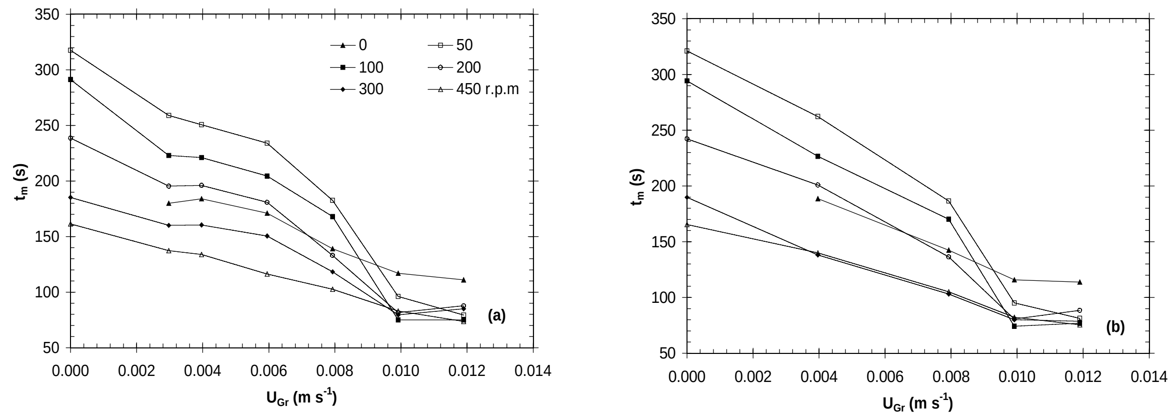

Figure 3. Superficial gas velocity and impeller speed effect on mixing time ((**a**) water, (**b**) culture medium)

Gas residence time (t_R) was calculated considering the reactor as a bubble column, which generates an approach to the phenomenon in the riser. Fig. 4 shows a residence time increase for impeller speeds ≤ 300 r.p.m. while the superficial gas velocity is kept constant; however, greater agitation speeds have no appreciable effect over the residence time. Perhaps, this phenomenon is due to the radial streamlines generated by the Rushton impellers; whereas at constant impeller rates the increment of superficial gas velocity generates a residence time reduction. A slope change is observed in coalesced bubble flow regime (U_{Gr} > 0.01 m s^{-1}), increasing the residence time falling off rate; this, since the bubble rising velocity is increase due to the coalescence effect.

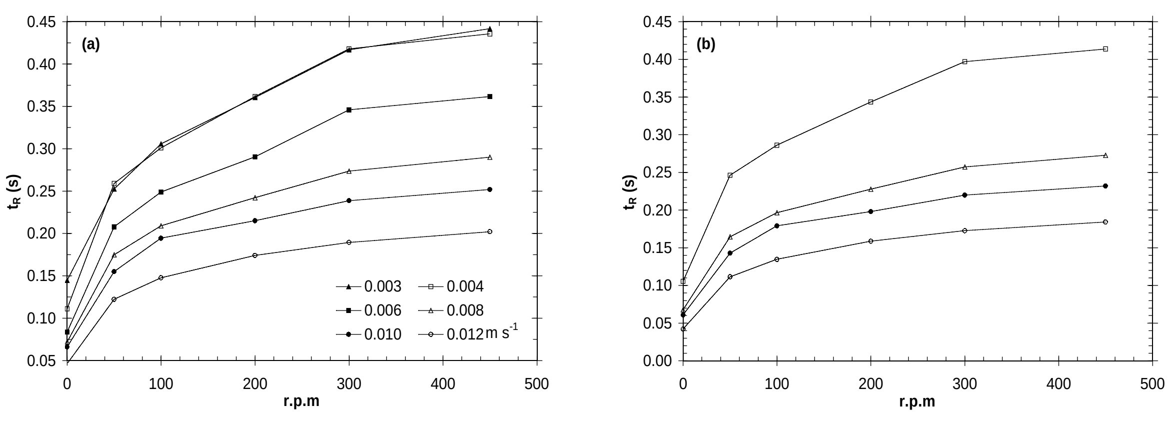

Figure 4. Superficial gas velocity and impeller speed effect on residence time ((**a**) water, (**b**) culture medium)

3.2.Gas holdup (ε)

Superficial gas velocity and agitation speed effect over gas holdup is presented in Fig. 5. For all cases, increasing the superficial gas velocity is observed a fast gas holdup rise until the coalesced bubble flow regime is reached, where is generated a reduction in the holdup increment.

Bubbly flow regime, which interactions among rising bubbles are few, persisted until a $U_{Gr} \leq$ 0.008 m s^{-1}; while at U_{Gr} 0.010 m s^{-1} an increase in the gas holdup rising rate is detected, showing this condition as a transition velocity between the bubbly and coalesced bubbled flow regime (or churn turbulent flow). Since, it is followed by a diminution of gas holdup due to higher bubble coalescence with a superficial gas velocity of 0.012 m s^{-1}. However, an increment of impeller agitation speed over 300 r.p.m. seems not to have a significant effect on gas holdup for all superficial gas velocities.

The gas holdup gotten with culture medium shows a diminution of ~ 9 % compared with the gotten with water (mainly at impeller agitation speeds > 100 r.p.m. and U_{Gr} > 0.008 m s^{-1}), because of turbulence damping effect of solids in culture medium. The variance coefficient of gas holdup measurements was less than 5 %. Furthermore, at U_{Gr} > 0.008 m s^{-1} was observed an enhancement of bubbles retained in the downcomer, effect caused by the increment of the fluid flow in this region that avoids bubbles to rise easily.

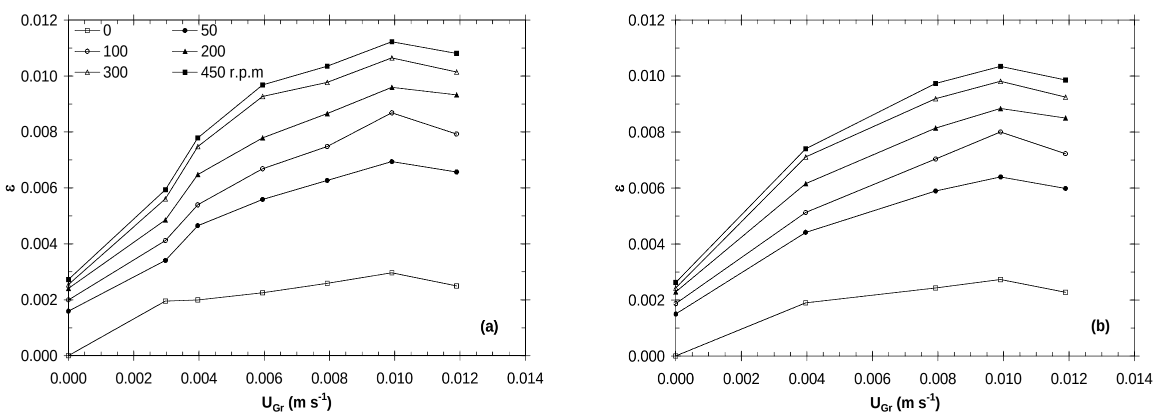

Figure 5. Superficial gas velocity and impeller speed effect on gas holdup ((**a**) water, (**b**) culture medium)

The effect of power input on gas holdup and bubble rising velocity is shown in Fig. 6; where an increase of pneumatic power input (associated with U_{Gr}) cause an increment on ε and U_b. Nevertheless, the presence of mechanical agitation drastically reduces bubble rising velocity, phenomenon agree with its effect on the residence time.

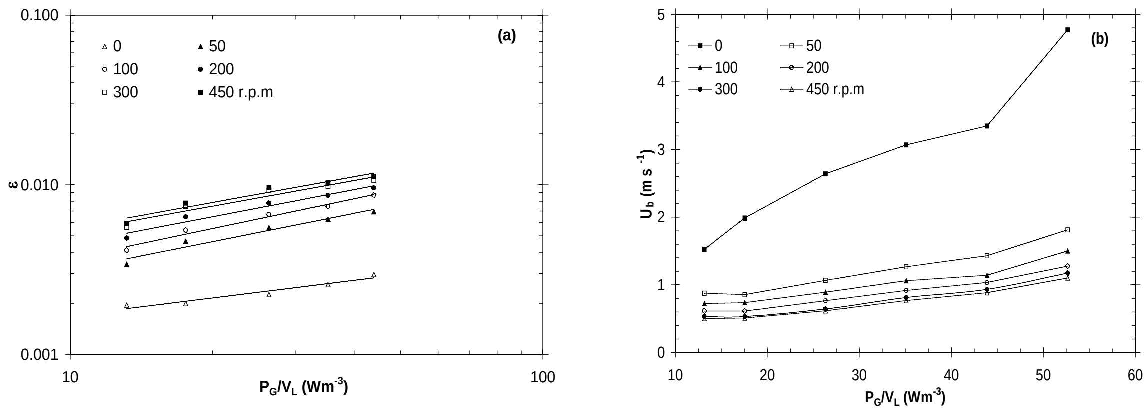

Figure 6. Input power and impeller speed effect on (**a**) gas holdup and (**b**) bubble rising velocity for the air-water system.

Eq. 12 is the correlation that better describes the behavior of gas holdup in airlift reactors:

$$\varepsilon = \alpha \left(\frac{P_G}{V_L} \right)^{\beta} \tag{12}$$

Where α and β are parameters related with the fluid reological nature and flow regime respectively (Chisti, 1989). α and β were evaluated for the reactor without mechanical agitation and superficial gas velocity in bubbly flow regime ($U_{Gr} < 0.010$ m s^{-1}), getting Eq. 13 and Eq. 14 for air-water and air-culture medium systems, respectively. The correlation coefficients were greater than 94 %.

$$\varepsilon = 8 \times 10^{-4} \left(\frac{P_G}{V_L} \right)^{0.3469} \tag{13}$$

$$\varepsilon = 6 \times 10^{-4} \left(\frac{P_G}{V_L} \right)^{0.3873} \tag{14}$$

The difference between the values of α and β for both fluids are of the order of $2x10^{-4}$ W^{-1} m^3 and 0.04 units, respectively. These values present a similar tendency to the presented by Chisti (1989), which β increases with the presence of solids in suspension.

The following nonlinear correlations were gotten for gas holdup in the hybrid reactor using Statgraphics ® plus v. 5.0. The proposed correlations keep a similar structure to the reported by Chisti (1989) for airlift reactors; in which α, that represent the culture medium reological nature is modified by a parameter that is function of the culture medium solid concentration (C_S, % $^W/_V$, for $C_S \leq 1.4$ %); and β, represents the flow condition, is adjusted by a factor that depends on the agitation speed (N, s^{-1}). Fig. 7 presents a suitable representation of experimental gas holdup, with

±10% and ±5% of maximum deviation for bubbly (Eq. 15) and coalesced bubble flow (Eq. 16) regime, respectively.

$$\varepsilon = \left[0.00155208 - 5.72964 \times 10^{-5}\ C_S\right] \left(\frac{P_G}{V_L}\right)^{0.189013 + 0.215991\ N^{0.255007}} \tag{15}$$

$$\varepsilon = \left[0.0288281 - 0.001700792\ C_S\right] \left(\frac{P_G}{V_L}\right)^{-0.132772 - 0.227202\ N^{-0.334161}} \tag{16}$$

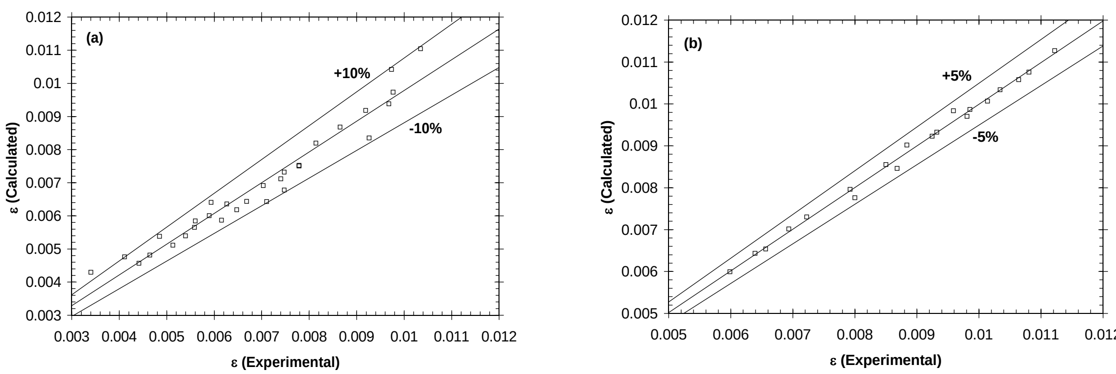

Figure 7. Predicted (Eqs. 15–16) vs. measure gas holdup for bubbly (**a**) and coalesced bubble flow (b).

3.3.Global oxygen transfer coefficient ($k_L a$)

Global oxygen transfer coefficient (k_La) dependence with the superficial gas velocity and agitation speed can be seen in Fig. 8. This is consistent with the observed gas holdup behavior (Fig. 5); now that gas holdup is the main factor that influences the gas-liquid interfacial area. At U_{Gr} = 0.008 m s^{-1} is presented the highest k_La for different agitation speeds; nevertheless, at U_{Gr} > 0.008 m s^{-1} (transition regime between bubbly and coalesced bubbled flow) is observed that k_La decrease and the reduction rate rises with the agitation speed increment. An exceptional case happens at 50 r.p.m. for all the U_{Gr}, where k_La is smaller than in agitation absence. Agitation speeds > 100 r.p.m. support the k_La enhancement; however the highest k_La is gotten at 300 r.p.m. with any U_{Gr}, without a significant difference at higher impeller speeds. The almost constant k_La value at U_{Gr} > 0.008 m s^{-1} and agitation speeds > 300 r.p.m., can be explained since the dissolved oxygen electrode was located in the downcomer and for these conditions a high air bubble retention appears due to the increase of liquid circulation speed in the loop. On the other hand, at agitation speeds ≥ 300 r.p.m. a significant k_La fall comes out as a consequence of higher gas coalescence. A maximum k_La reduction of 20 % was observed with the culture medium at the same reactor conditions with water, possibly as a result of the enhancement in gas coalescence and the interface resistance generated by the ions and suspended solids. The variance coefficient in k_La measurement was < 7 %.

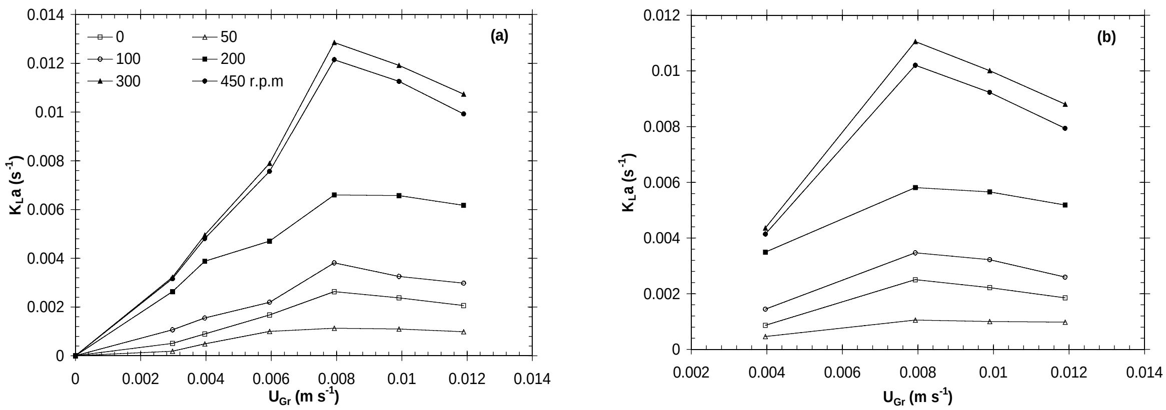

Figure 8. Superficial gas velocity and impeller speed effect on global oxygen mass transfer coefficient ((**1**) water, (**2**) culture medium)

Theoretical consideration of a k_La plot against gas holdup ratio ε/(1−ε) is expected to be linear in any sparged reactor, irrespective of the fluid used and prevailing flow regime. This has been demonstrated for airlift reactors and bubble columns getting an average slope of 1.187 with a coefficient of variance of 15% (Chisty, 1989). In the designed hybrid reactor, the dependence between k_La and gas holdup ratio with the tested aeration and stirring rates showed to be higher than the unit (from 1.3 to 2.6) for both fluids (Table 1, Fig. 9), which demonstrates that the global mass transfer coefficient and the fraction of gas holdup are higher in the designed hybrid reactor, possibly by a radial oxygen transfer through the filtration module.

Table 1. Slope and correlation coefficient for the curves of k_La vs. ε/(1 - ε)

Fluid	**r.p.m.**	**0**	**50**	**100**	**200**	**300**	**450**
Water	**Slope**	3.4847	2.6476	1.7022	1.4050	2.1856	2.4206
	r^2	90.16	94.49	91.37	96.60	93.04	90.05
Culture Medium	**Slope**	1.3391	1.0591	1.2369	1.8541	2.1534	2.1228
	r^2	99.91	90.61	96.06	98.16	95.15	97.12

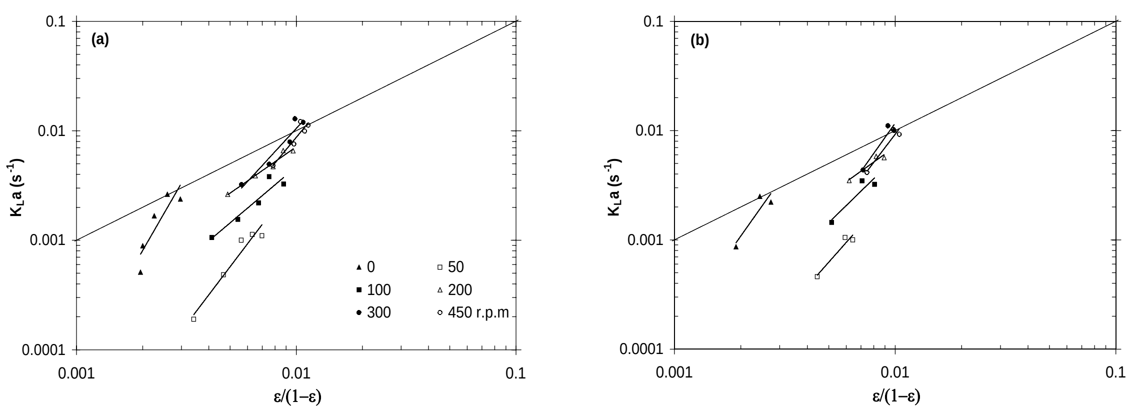

Figure 9. k_La vs. ε/(1 - ε) at different impeller speeds (**a** water, **b** culture medium)

About the oxygen mass transfer efficiency, for all the cases the maximum efficiency was gotten at U_{Gr} 0.008 m s^{-1} (Fig. 10). Based on the impeller speed is observed that E_M for 50 and 450 r.p.m. is lower than the gotten without mechanical agitation. However, for water and culture medium at agitation conditions of 100 – 300 r.p.m. with $U_{Gr} \geq 0.008$ m s^{-1} higher oxygen mass transfer efficiencies are achieved compared to the values of E_M in the reactor without agitation.

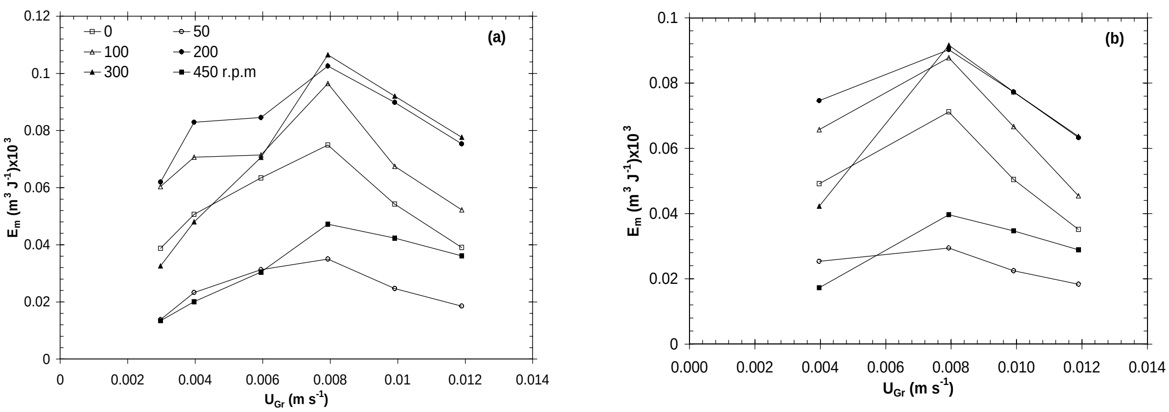

Figure 10. Impeller speed and superficial gas velocity effect on oxygen mass transfer efficiency E_M (**a** water, **b** culture medium)

k_La dependence with agitation speed (N, s^{-1}), superficial gas velocity (U_{Gr}, m s^{-1}) and specific power input due to the mechanical agitation (P_M / V_L, W m^{-3}) was correlated for both air-water and air-culture medium systems, using the software Statgraphics® plus v. 5.0. The nonlinear correlation gotten for bubbly flow ($U_{Gr} \leq 0.008$ m s^{-1}) and coalesced bubble flow regimes (U_{Gr} > 0.008 m s^{-1}) are Eqs. 17 and 18, respectively:

$$k_L a = \left[0.0847102 + 70.979 {\frac{P_M}{V_L}}^{1.12561} \right] U_{Gr}^{\,1.12561 (N + 5.73013)^{0.756744}} \qquad r^2 = 96.84 \qquad (17)$$

$$k_L a = \left[0.0558266 - 0.0640933 {\frac{P_M}{V_L}}^{0.10102} \right] U_{Gr}^{\,0.722712 - 0.170813 N} \qquad r^2 = 95.29 \qquad (18)$$

Predictions from Eqs. 17 and 18 present a ±15% maximum deviation with regard to the measure data (Fig. 11). The proposed correlations agree with the reported by Chisti and Jáuregui-Haza (2002), but there is a noticeable difference with the reported coefficients, possibly due to the difference in the dimensions and geometric configuration of the reactor (~1.5 m^3, aerated in annular zone and with a ratio of $A_r/A_d = 1.27$)

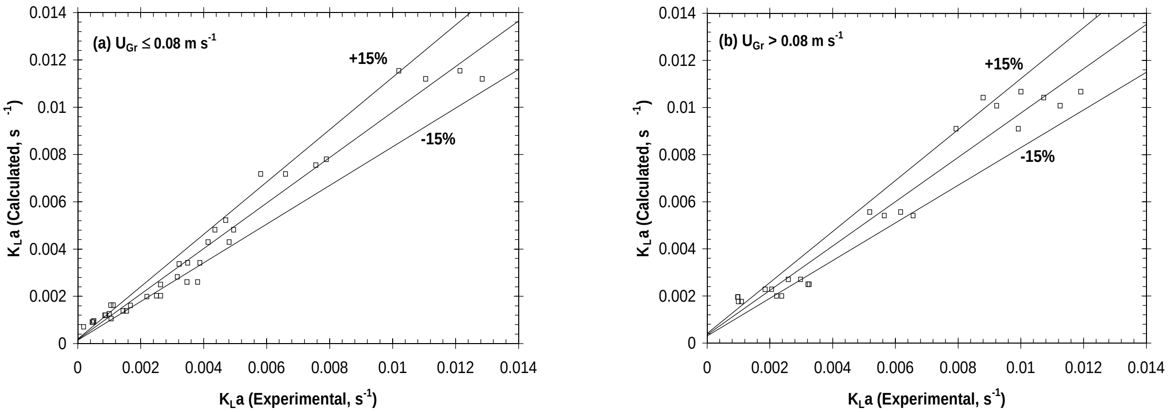

Figure 11. Predicted (Eqs. 17 – 18) vs. measure k_La for air-water system in bubbly (**a**) and coalesced bubbled flow (**b**)

Considering the effect of suspended solids of the culture medium, correlations for k_La as function of agitation speed (N, s^{-1}), concentration of solids (C_S, % $^P/_V$), superficial gas velocity (U_{Gr}, m s^{-1}) and specific mechanic power (P_M / V_L, W m^{-3}) were gotten for the bubbly (Eq.19) and coalesced bubbled flow regime (Eq. 20):

$$k_La = \left[-0.966334 + e^{-0.0012331\,(C_S - 49.4487)} + 61.7874 \left(\frac{P_M}{V_L} \right)^{1.10928} \right] U_{Gr}^{\;0.798673 + 0.0844283\,N} \tag{19}$$

$$k_La = \left[0.000144644 + -2.2671 \times 10^{-5}\, C_S + 0.0876241 \left(\frac{P_M}{V_L} \right)^{1.12264} \right] U_{Gr}^{\;-0.600065 + 0.09531\,N} \tag{20}$$

Hydrodynamic characterization of the designed reactor (airlift with filtration module in the concentric zone and mechanically agitated) with air-water and air-culture medium systems, is in agreement with the results reported by Chisti and Jaúregui-Haza (2002) and Pollard *et al*. (1997) considering the differences in operation volumes, geometry and draft-tube configuration used in each case.

Likewise Chisti (1989), Molina *et al*. (1999) and Chisti and Jaúregui-Haza (2002) describe a transition region between bubbly and coalesced bubble flow regime; which gas holdup and global mass transfer coefficient do not present a significant change and in contrast they decrease due to aeration and agitation conditions in churn turbulent flow. This phenomenon was identified in the designed bioreactor, which transition regime was detected at U_{Gr} > 0.008 m s^{-1}.

4.Conclusions

The designed reactor showed a hybrid behavior of an airlift reactor with mechanical agitation, presenting better global oxygen mass transfer coefficients than airlift reactor without mechanical agitation. The highest efficiency of oxygen transfer for the air-water and air-culture medium systems was presented at a stirring rate of 300 r.p.m. and U_{Gr} 0.008 m s^{-1}.

A transition regime between bubbly and coalesced bubbled flow was identified for the modified reactor at a superficial gas velocity of 0.010 m s^{-1}, condition in which a reduction of global mass transfer coefficient appears.

Nonlinear correlations for the prediction of gas holdup and global oxygen mass transfer coefficient were gotten for bubbly and coalesced bubbled flow, with correlation coefficients higher than 94%.

The hydrodynamic characterization made using water and culture medium displays the same tendency for the measured parameters under different agitation and aeration conditions. A considerable difference between the measured values for both fluids is observed in oxygen mass transfer efficiency, k_La and gas holdup, whereas mixing time and residence time show very similar results.

The linear dependence between the global oxygen mass transfer coefficient and the gas holdup ratio for the designed hybrid bioreactor does not conserve the expected relation for bubble columns and airlift reactor (slope = 1), achieving average values of 2.30 and 1.62 for air-water and air-culture medium systems, respectively.

Nomenclature

A_d	cross-sectional area of downcomer, m^2
A_r	cross-sectional area of riser, m^2
C	instantaneous concentration of dissolved oxygen, kmol m^{-3}
C_S	concentration of suspended solids, kg m^{-3}
C_0	initial concentration of dissolved oxygen, kmol m^{-3}
C^*	saturation concentration of dissolved oxygen, kmol m^{-3}
d_i	impeller diameter, m
E	fractional approach to equilibrium defined by Eq. (6), dimensionless
E_M	mass transfer efficiency defined by Eq. (11), m^3 J^{-1}
g	gravitational acceleration, m s^{-2}
h_D	height of the gas-liquid dispersion, m
h_L	height of not gasified liquid, m
k_La	overall gas-liquid mass transfer coefficient, s^{-1}
N	rotational speed of impeller, s^{-1}
P	total power input, W
P_G	power input due to aeration, W
P_M	power input due to agitation, W
P_0	impeller power number, dimensionless
Re_i	impeller Reynolds number, dimensionless
t	time, s
t_m	mixing time, s
t_R	gas residence time, s
t_0	initial time, s

U_b average bubble rising velocity, m s^{-1}
U_{Gr} superficial gas velocity, m s^{-1}
V_L volume of culture medium, m^{-3}

Greek letters
α exponent in Eq. (12) that represent the reological nature of the fluid, dimensionless
β exponent in Eq. (12) that represent the condition of flow, dimensionless
ε gas holdup, dimensionless
ρ_L liquid density, kg m^{-3}
μ liquid viscosity, Pa s

References

Casas López J.L., Sánchez Pérez, J.A., Fernández Sevilla, J.M., Rodríguez Porcel, E.M., Chisti, Y. (2005). Pellet morphology, culture rheology and lovastatin production in cultures of *Aspergillus terreus. Journal of Biotechnology*. 116, 16 – 77.

Chisti, Y. (1989). Airlift bioreactor. (Elsevier, New York) p. 132 – 190, 230 – 257

Chisti, Y. and Jáuregui-Haza U.J. (2002). Oxygen transfer and mixing in mechanically agitated airlift bioreactor. *Biochemical Engineering Journal*. 10, 143 – 153.

Hudcova, V., Machon, V.and Nienow, A.W. (1989). Gas liquid dispersion with dual Rushton turbine impellers. *Biotechnology and Bioengineering.* 34, 617 – 628.

Mohanty, K., Das, D., Biswas, M. N. (2006). Hydrodynamics of a novel multi-stage external loop airlift reactor. *Chemical Engineering Science.* 61, 4617 – 4624.

Molina, E., Contreras, A. and Chisti, Y. (1999). Gas holdup, liquid circulation and mixing behavior of viscous newtonian media in a split-cylinder airlift bioreactor. *Trans IChemE, Part C*. 77, 27 – 32.

Pollard, D.J., Ison, A.P., Ayazi, Shamlou, P. and Lilly, M.D. (1997). Influence of propeller on *Saccharomyces cerevisae* fermentation in a pilot scale airlift bioreactor. *Bioprocess Engineering.* 16, 273 – 281.

Pollard, D.J., Ison, A.P., Ayazi, Shamlou, P. and Lilly, M.D. (1998). Reactor heterogeneity with *Saccharomyces erythraea* airlift fermentations. *Biotechnology and Bioengineering* 58, 453 – 463.

Sánchez Mirón, A., Cerón García, M.C., García Camacho, F., Molina Grima, E. and Chisti, Y. (2004). Mixing in bubble column and airlift reactors. *Trans IChemE,Part A.* 82 (A10), 1367 – 1374.