Mass transfer coefficient in stirred tank reactors for xanthan gum solutions

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Abstract

The oxygen transfer rate in Newtonian and non-Newtonian fluids in a stirred tank reactor (STR) of 20 l of operating volume has been studied. The volumetric mass transfer coefficient, \( k_L a \), has been measured by a dynamic technique in solutions of xanthan gum for a wide interval of operational conditions. This coefficient has been determined changing many variables, such as the superficial gas velocity \( (V_g) \), the stirred speed \( (N) \) or the power input by unit of volume \( (P/V) \). Also, the properties of the liquid phase, mainly the rheologic properties of the liquid, as well as the geometry of the stirrer, have been studied considering the effect of the number and type of stirrers (paddle or turbine), the number of blades of the stirrers and the sparger type (ring and disk). The \( k_L a \) values are dramatically affected by the viscosity of the liquid; to take into account this effect, the rheology of the system has been described using both the Ostwald-de Waele and the Casson models. Dimensional correlations —as a function of the operational conditions— have been obtained, and also dimensionless equations —expressing the Sherwood number as a potential function of the Reynolds, Aireation and Weber numbers— have been determined for several stirrer types.

Keywords: Oxygen transfer; Non-Newtonian liquids; Stirred tank reactor

1. Introduction

Gas–liquid contact is a matter of decisive importance for describing systems involving chemical reactions or biological processes. Mass transfer between phases may often become the limiting step of the overall process rate. In such cases, the volumetric oxygen transfer coefficient, \( k_L a \), must be known in order to carry out the design and scale-up of bioreactors. Examples are aerobic fermentation, in special cultures with highly viscous media, (e.g. production of antibiotics, polysaccharides and waste treatments). In all of these processes, the standard stirred tank reactor (STR) is used almost universally, especially in the fermentation industry.

Therefore, gas–liquid mass transfer in STR has been the subject of many works. Most of them dealing with those cases when the liquid phase is a Newtonian fluid or a low viscosity media [1–6]. When the gas–liquid mass transfer processes is enforced in a STR, additional complications arise if a non-Newtonian liquid is used, usually due to changes in the interfacial gas–liquid area [3].

Several measuring methods for \( k_L a \) have been employed, but only some of them yield indisputably correct data. Chemical methods, based on the measurement of a chemical reaction rate, did not appear suitable since the addition of chemical reagents changes the bubble coalescence behaviour. Physical methods, both those carried out under steady state conditions and those in transient state or dynamic methods, are more used in the last decades. The steady state method is based on measuring the oxygen concentration of the air inflow and air outflow, accurate modelling of gas phase mixing being required for the correct interpretation of the measurement of the oxygen transfer rate [7]. The dynamic method is based on the measurement of the dissolved oxygen concentration during desorption or absorption experiments, eliminating the addition of reactive substances and proving satisfactory reproducibility. The characteristic time for most oxygen electrodes is of the same magnitude than the characteristic time for the oxygen transfer process \( (1/k_L a) \); therefore, most oxygen electrodes have a short response time to changes in dissolved oxygen concentration, which is a feature of fundamental importance to determine \( k_L a \).

The oxygen mass transfer in STR is a function of many variables, such as the physical properties of the liquid (viscosity, surface tension, etc.), the geometry of the vessel and stirrer, the type of sparger and the operational conditions. Unfortunately, the available information in the literature about the effect of these variables on the mass transfer is sometimes confusing.
In the literature, both dimensional and dimensionless equations for the mass transfer volumetric coefficient, have been proposed. The $k_L a$ values have been correlated to the combination of stirrer speed, $N$, superficial gas velocity, $V_s$, and liquid effective viscosity, $\mu_{eff}$, obtaining equations as follows:

$$k_L a = C_1 \cdot V_s^{n} \cdot N^{m} \cdot \mu_{eff}^{p} \quad (1)$$

Nevertheless, other equations have also been proposed including the effect of power input by unit volume ($P/V$):

$$k_L a = C_2 \cdot V_s^{n} \cdot \left( \frac{P}{V} \right)^{m} \cdot \mu_{eff}^{p} \quad (2)$$

where the constants $C_1$ and $C_2$ depend on the geometry of the vessel and the stirrer employed.

Another approach has been the use of another type of equations, using dimensionless numbers, as follows:

$$Sh = C \cdot Re^{n} \cdot Na^{p} \cdot Wi^{q} \quad (3)$$

The exponent values in both types of equations change for different authors, also being influenced by the geometry of the tank.

This work deals with the study for Newtonian and non-Newtonian liquids of the influence of several of these variables on the mass transfer rate in stirred tank reactors, such as the stirrer type and the gas distributor design, also taking into account the operational conditions, such as superficial gas velocity and stirrer speed. In order to check the influence of other properties of the liquid, runs were performed employing not only water but electrolyte solutions and aqueous solutions of xanthan gum for varying the rheological properties of the liquid—described according to Ostwald–de Waele and Casson models. Experimental data have been fitted using both types of equations, dimensional as Eqs. (1) and (2), and dimensionless as Eq. (3).

2. Experimental

This study has been carried out in a baffled stirred tank 20 l in volume (0.30 m in diameter and a diameter to height ratio equal to 1). Four baffles (0.31 m high, 0.03 m in wide and 0.001 m thick) were installed to favour turbulence and to prevent the formation of vortices. Agitation was provided with a multiple-stirrer system consisting of a shaft with identical evenly-spaced and paddle or disk turbine stirrer type. The diameter of these stirrers was always of 10 cm, and they were located at 10 cm from the bottom of the reactor. Each stirrer is made of a number of blades having 1 mm in thickness. The characteristic dimensions of the stirrers are listed in Table 1. Two gas sparger were employed, one disk and another ring type, 10 cm in diameter with orifices of 1 mm and 2 mm in diameter, respectively. The disk sparger type containing four series of 12 uniformly distributed orifices and the ring sparger type containing 24 distributed orifices each 15°. For all experiments, air was used as gas, the superficial gas velocities were varied from $4.0 \times 10^{-4}$ to $1.1 \times 10^{-3}$ m s$^{-1}$ and the stirrer speed from 3.33 to 7.5 s$^{-1}$.

Commercial xanthan gum from Jungbunzlauer (Austria), food grade, was used for changing the rheological behaviour of the liquid. Solutions were prepared by slowly adding the xanthan gum to water, while stirring at 5 s$^{-1}$ for 24 h and maintaining the solution temperature at 28 °C. A constant density of 1000 kg m$^{-3}$, and an oxygen diffusivity of $2.13 \times 10^{-9}$ m$^2$ s$^{-1}$ were assumed for all solutions. The surface tension was measured at 28 °C using a ring tensiometer and were found to be identical to the surface tension of water ($72 \times 10^{-3}$ N m$^{-1}$).

An oxygen electrode, connected to a Rossmount polargraphic oxygen analyzer, was inserted horizontally through the vessel wall at the same level as the impeller. The probe had a 95% dynamic response, always in less than 20 s. Gas flow was determined by a rotameter and the stirrer speed was measured using a stroboscope.

The mass transfer volumetric coefficient, $k_L a$, was determined by a dynamic technique as previously described by other authors [8–10]. First, the liquid placed into the vessel was sparged with air until the saturation dissolved oxygen concentration was reached. Then, nitrogen was introduced downward into the vessel and the dissolved oxygen concentration was recorded as function of time. Hence, the rate of change of concentration in the liquid phase is given by:

$$\frac{dC_l}{dr} = k_L a \cdot (C^* - C_l) \quad (4)$$

In order to take into account the rheological properties of the liquid, the apparent viscosity of several xanthan gum solutions was determined. A Brookfield LVT-Synchro-electric viscosimeter, with a microsample jacketed, and a Brookfield spindle no 18 were used for the rheological measurements. Several studies have dealt with the rheology of xanthan gum solutions and most have found pseudoplastic behaviour. The flow behaviour has been usually described in terms of the Ostwald–de Waele or power-law model [11–18]:

$$\tau = k \cdot \gamma^n \quad (5)$$

Polymer solutions have different degrees of pseudoplasticity. Consistency index, $k$, varied from $7.08 \times 10^{-1}$ to $26 \times 10^{-1}$; the power-law index $n$ was between 0.3 and 0.8.
Table 2
Rheological parameters of the liquid at 28 °C, for both Ostwald-de Waele and Casson models

<table>
<thead>
<tr>
<th>C, (μe x 10⁻⁹)</th>
<th>Ostwald-de Waele Model</th>
<th>Casson Model</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>k x 10⁻¹</td>
<td>n</td>
</tr>
<tr>
<td>0.1</td>
<td>4.92</td>
<td>0.84</td>
</tr>
<tr>
<td>0.2</td>
<td>9.50</td>
<td>0.80</td>
</tr>
<tr>
<td>0.3</td>
<td>14.06</td>
<td>0.75</td>
</tr>
<tr>
<td>0.4</td>
<td>19.73</td>
<td>0.71</td>
</tr>
<tr>
<td>0.5</td>
<td>25.67</td>
<td>0.69</td>
</tr>
<tr>
<td>0.6</td>
<td>33.33</td>
<td>0.58</td>
</tr>
<tr>
<td>0.7</td>
<td>146.32</td>
<td>0.49</td>
</tr>
<tr>
<td>1.0</td>
<td>212.70</td>
<td>0.46</td>
</tr>
<tr>
<td>2.0</td>
<td>707.61</td>
<td>0.32</td>
</tr>
</tbody>
</table>

9.42 x 10⁻³ Pa s and flow behavior index, n, varied from 0.32 to 0.84. Recently a few studies suggested the existence of a yield stress, τo [19,20]. To evaluate the apparent yield stress, the experimental data were fitted to the Casson model:

\[ \tau^{1/2} = k_e \gamma^{1/2} + k_c \gamma^{1/2} \]  
(6)

In Table 2 the values of the parameters for both rheological models are shown. The shear rate in the STR is locally non-uniform. In the literature, several expressions for an effective mean shear rate, γeff, in the STR have been used to calculate the effective viscosity [21,22]. Any correlation of the volumetric mass transfer coefficient will be affected by the rheological model used and the choice of the relationship for the effective shear rate. In this work, the average shear rate was related to the stirrer speed by the equation proposed by Ref. [21], which is the most often used in the literature [1,3,16,23]:

\[ \gamma_{eff} = K \cdot N \]  
(7)

where K has values from 11 to 13, for paddle and turbine stirrers type [21,24]. In this work a value of 11.5 was assumed [1,22] and the effective viscosity, μeff, was then defined as:

\[ \mu_{eff} = \frac{\tau}{\gamma_{eff}} \]  
(8)

In the case of fluids with a rheological behaviour that can be described by the power-law model, the effective viscosity is given by the following equation, indicating that the effective viscosity decreases with increasing stirrer speed:

\[ \mu_{eff} = k \cdot (K \cdot N)^n \]  
(9)

When the rheological properties of the liquid is described by the Casson model, at high shear rates, the effect of τo becomes negligible and the viscosity can approximately be given by Eq. (10); if μe is considered to be equal to k_c then Eq. (11) is obtained:

\[ \tau^{1/2} = k_c \cdot \gamma^{1/2} \]  
(10)

\[ \tau = \mu_c \cdot \gamma \]  
(11)

where the parameter μ_c is equivalent to a Newtonian viscosity, μ_c not dependent on shear rate.

3. Results and discussions

The volumetric mass transfer coefficient, kLa, has been determined in a wide interval of operational conditions. The measurements have been carried out changing the superficial gas velocity, V_m, between 0.4 x 10⁻³ and 1.1 x 10⁻³ m s⁻¹, the stirred speed, N, between 3.3 and 7.5 s⁻¹ (therefore the power input by unit of volume, P/V, was also changed). Also the rheological properties of the liquid have been changed, using xanthan gum solutions at different concentration. Finally the effect of the number and type of stirrers (paddle or turbine), the number of blades of the stirrers and the sparge type (ring and disk) have also been taken into account. In all, a total of 387 measurements of kLa were analyzed in this work (data available on request). The influences of the operational conditions (V_m, N or P/V) and the viscosity of the liquid are going to be considered later, integrating such influences in equations.

3.1. Influence of the geometry and number of stirrers

The stirrer type used in mechanically-agitated vessels plays an important role, producing efficient heat transfer and intensive mixing and aeration. For example, one of the major problems during the production of microbial polysaccharides is to maintain homogeneous mixing of the broth and adequate oxygen transfer to the cells. Also, in the selecting of stirrer type, it is very important the shear effect under certain conditions of fermentations, principally using shear-sensitive biological organisms [25-28]. A large variety of stirrs such as turbines, paddles, propeller and other stirrer geometries (such as Lightning, Maxfol T, Intermig, etc.) has been used. However, oxygen transfer studies in mechanically-agitated vessels with solutions of polymers with rheological properties similar to those of the broth in xanthan fermentations have practically all been made with disk turbines [2,29,4,30,31] and paddle stirrers [3,32,33].

In this work, the effect of numerous stirrers with different dimensions and number of blades has been studied. The values of kLa, for a stirrer speed of 5 s⁻¹ and xanthan gum concentration of 0.1 kg m⁻³ are shown in Figs. 1 and 2, for different turbine and paddle stirrers. As can be seen, a similar influence of operational condition on kLa is found. However, the disk turbine with six and eight curved blade stirrers give larger kLa values than those obtained with flat blade disk turbine and the paddle stirrers, at least at lower viscosity.

Fig. 3 presents the results of kLa obtained in a multiple-stirrer system. As can be seen, a small but significant difference is found at low viscosity when several stirrers are used. But for high viscosity almost the same results are obtained when one, two or three stirrers are employed.
3.2. Effect of sparger type

A relevant improvement on mass transfer rate is achieved by increasing the interfacial area. The gas distributor, as well as mechanical agitation, would appear to be a likely instrument to enlarge the interfacial area, since direct influence can be exercised on the initially formed bubble sizes. Several authors have discussed the influence of sparger type on $k_{La}$. Höcker et al. [34] and Leckie et al. [35,36] found that $k_{La}$ values depend on the sparger type. Rewatkar and Joshi [31] have proposed that sparger location with respect to the impeller is the most important parameter. Nevertheless, other authors [7] have reported that the choice of sparger has little effect on the mass transfer rate.

In this work, two different gas distributor types have been tested with a 0.1% (w/v) xanthan gum solution: ring and disk sparger type. The test was carried out with only one stirrer, a curved blade disk turbine with six curved blades. Fig. 4 shows the results obtained, as can be seen, the sparger type does not appear to exert any significant influence on mass transfer. Only the $k_{La}$ value is smaller when disk sparger of 2 mm hole diameter is used.

3.3. Dimensional correlation of $k_{La}$ values

Many correlations for $k_{La}$ have been proposed in the literature, in both dimensional and dimensionless forms. The
early dimensional correlations for volumetric oxygen transfer coefficient proposed for Newtonian and non-Newtonian fluids—equations similar to Eq. (1)—take only into account the operational conditions \( (N, V_s) \) and the variations of rheological properties \( (\mu_{\text{eff}}) \) \([2,3,6]\). Sometimes \( N \) is omitted, and another variable, the power input per unit liquid volume, is used, equations being formulated similar to Eq. (2) \([3,5,8,37-40]\). Only the correlations proposed by Nishikawa et al. \([23]\) and Roman et al. \([39]\) include both the variables \( N \) and \( P/V \).

In this work, \( k_La \) has been fitted to experimental data by regression analysis to the Eq. (1), which takes into account superficial gas velocity, \( V_s \), stirrer speed, \( N \), and liquid effective viscosity, \( \mu_{\text{eff}} \), for different type and number of stirrers and a wide interval of viscosity of the liquid, those corresponding from 0.1 to 2.0 kg m\(^{-3}\) of polymer concentration in solution.

When the rheology of the liquid is described by a power-law model, the following equation was obtained:

\[
k_La = C_1 \cdot V_s^{1/3} \cdot N^{2/3} \cdot \mu_{\text{eff}}^{-2/3}
\]  

(12)

while if the Casson model is employed, the following equation is obtained:

\[
k_La = C_1' \cdot V_s^{1/3} \cdot N^{2/3} \cdot \mu_{\text{eff}}^{-1.0}
\]  

(13)

If Eq. (2) is employed for the experimental data fitting, that is replacing the stirrer speed for the gassed power input per unit of volume, considering a power-law rheological model, the following relationship is found:

\[
k_La = C_2 \cdot V_s^{1/3} \cdot \left( \frac{P}{V} \right)^{0.6} \cdot \mu_{\text{eff}}^{-2/3}
\]  

(14)

while when the Casson model is employed, the following equation is obtained:

\[
k_La = C_2' \cdot V_s^{1/3} \cdot \left( \frac{P}{V} \right)^{0.6} \cdot \mu_{\text{eff}}^{-1.0}
\]  

(15)

In Tables 4 and 5 the values of the constants \( C_1, C_1', C_2 \) and \( C_2' \) obtained by regression of the experimental data, for different types and number of stirrers, are given. The results show that the effect of the superficial gas velocity has a direct effect upon oxygen transfer rate coefficient, giving a value of 0.67 for the exponent of this variable. Similar values of this exponent on superficial gas velocity have been obtained in previous works, as can be seen in Table 3, scarcely depending on the equipment geometry, for both Newtonian and non-Newtonian fluids \([3,6,41]\).

The analysis of the influence of stirrer speed shows that \( k_La \) increases with \( N \), according to an exponent value of 2. Similar values of that exponent on stirrer speed have been reported in previous works \([1,6,42]\). However, the values are slightly higher than those obtained for non-Newtonian fluids by other authors \([2,3]\).

The effect of polymer concentration, that is the viscosity and rheological behaviour of the liquid, on \( k_La \) has been studied over a wide interval of operational conditions. It was found that the mass transfer rate decreases with increasing effective viscosity of the liquid, due to several effects:

- increased viscosity leads to a decrease in the degree of liquid flow turbulence and hereby to a decrease in mass transport intensity;
- with the smaller degree of turbulence, the dynamic equilibrium bubble size increases.
Table 4
Values calculated for constants $C_1$, $C_2$ and confidence interval for the 95% of probability. Experimental values of $k_L a$ fitted to Eqs. (12) and (13).

<table>
<thead>
<tr>
<th>Stirrer</th>
<th>Ostwald-de Waele model</th>
<th>Casson model</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>$k_L a = C_1 V^{2/3} N^{2/3} \mu_c^{3/5}$</td>
<td>$k_L a = C_2 V^{2/3} N^{2/3} \mu_c^{3/5}$</td>
</tr>
<tr>
<td>4FBT</td>
<td>$5.31 \pm 0.10$</td>
<td>$6.50 \pm 0.15$</td>
</tr>
<tr>
<td>6CBT</td>
<td>$6.22 \pm 0.13$</td>
<td>$7.56 \pm 0.13$</td>
</tr>
<tr>
<td>Two-4FBT</td>
<td>$6.06 \pm 0.14$</td>
<td>$8.11 \pm 0.25$</td>
</tr>
<tr>
<td>Two-6CBT</td>
<td>$7.60 \pm 0.13$</td>
<td>$9.32 \pm 0.18$</td>
</tr>
</tbody>
</table>

Table 5
Values calculated for constants $C_1$, $C_2$ and confidence interval for the 95% of probability. Experimental values of $k_L a$ fitted to Eqs. (14) and (15).

<table>
<thead>
<tr>
<th>Stirrer</th>
<th>Ostwald-de Waele model</th>
<th>Casson model</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>$k_L a = C_1 V^{2/3} \left( \frac{P}{V} \right)^{0.6} \mu_c^{2/3}$</td>
<td>$k_L a = C_2 V^{2/3} \left( \frac{P}{V} \right)^{0.6} \mu_c^{1.0}$</td>
</tr>
<tr>
<td>4FBT</td>
<td>$7.20 \pm 0.14$</td>
<td>$8.75 \pm 0.23$</td>
</tr>
<tr>
<td>6CBT</td>
<td>$6.85 \pm 0.14$</td>
<td>$8.34 \pm 0.15$</td>
</tr>
<tr>
<td>Two-4FBT</td>
<td>$6.23 \pm 0.17$</td>
<td>$7.59 \pm 0.27$</td>
</tr>
<tr>
<td>Two-6CBT</td>
<td>$5.76 \pm 0.11$</td>
<td>$7.06 \pm 0.15$</td>
</tr>
</tbody>
</table>

In this work, the rheology of the liquid has been described both by the power law and by the Casson model. The results show that this influence can be described employing exponent values on the $\mu_{eff}$ and $\mu_c$ of $-2/3$ and $-1$, respectively.

The use of the variable $P/V$ instead of $N$ has the advantage that $k_L a$ is a result of bubble coalescence that decreases rapidly with the ion concentration increasing. This explains the different $k_L a$ values when an electrolyte is added to water as shown in Fig. 6, depending on the grade of coalescence and salt concentration. The $k_L a$ values obtained by Ogut and Hatch [3], using the sulphite measurement method, are much higher than those usually obtained for an electrolyte solution. This is due to the strong ionic state of the sulphite solution and to the chemical reaction enhancement [46]. The values obtained in this work for electrolyte solutions (0.1 M Na$_2$SO$_4$) are found to be in the interval of the values proposed by Van’t Reit [7] and Nocentini et al. [4]. The addition of a surface active agent (e.g. antifoaming) causes a decreases in $k_L a$, a consequence of a change in the specific surface area and the action as a barrier to mass transfer [47].

![Fig. 5. Prediction of several equations of $k_L a$ in water. $V_r = 0.005$ m s$^{-1}$](image_url)

![Fig. 6. Prediction of several equations of $k_L a$ in solutions of electrolytes. $V_r = 0.005$ m s$^{-1}$](image_url)
Values calculated for constants $C$, $C'$ and confidence interval for the 95% of probability. Experimental values of Sherwood number fitted to Eqs. (18) and (19).

<table>
<thead>
<tr>
<th>Stirrer type</th>
<th>$C$</th>
<th>$C'$</th>
</tr>
</thead>
<tbody>
<tr>
<td>4FBT</td>
<td>6.86 ± 0.12</td>
<td>0.20 ± 0.07</td>
</tr>
<tr>
<td>6CBT</td>
<td>8.04 ± 0.16</td>
<td>0.25 ± 0.07</td>
</tr>
<tr>
<td>Two-4FBT</td>
<td>9.77 ± 0.21</td>
<td>0.27 ± 0.07</td>
</tr>
<tr>
<td>Two-6CBT</td>
<td>9.83 ± 0.17</td>
<td>0.31 ± 0.07</td>
</tr>
</tbody>
</table>

3.4. Dimensionless correlation of $k_La$ values

The mass transfer coefficient, $k_La$, for non-Newtonian fluids in sparged and stirrer vessels is considered to be a function of numerous variables, as indicated above. In order to determine the combined effects of such variables, including, operational conditions, liquid and gas properties (diffusivity, density, surface tension and viscosity) and the geometry of the tank, a relationship with a dimensionless number can be used:

$$\frac{k_LaT^2}{D_L} = f_2\left(\frac{\rho NT^2}{\mu_{eff}}, \frac{\rho N^2T^3}{\mu_{eff} \cdot \rho D_L \cdot \sigma \cdot V}, g\right)$$ (16)

This type of equation has been employed by many authors [1, 23, 42, 48], but different dimensionless numbers have been used in the different works.

In this work, after different trials (combinations of different dimensionless numbers), the best results have been obtained employing the following relationship:

$$\frac{k_LaT^2}{D_L} = C:\left(\frac{\rho NT^2}{\mu_{eff}}, \frac{NT}{V}, \frac{\rho N^2T^3}{\sigma}\right)^{1/3}$$ (17)

When the Ostwald-de Waele model is used for the substitution of the effective viscosity in Eq. (17), the experimental data are fitted to the following equation:

$$\frac{k_LaT^2}{D_L} = C:\left(\frac{\rho NT^2 - \rho N^2T^3}{k \cdot \rho \cdot \mu_{eff}}, \frac{NT}{V}, \frac{\rho N^2T^3}{\sigma}\right)^{1/3}$$ (18)

while if the Casson model is employed, then the following equation is obtained:

$$\frac{k_LaT^2}{D_L} = C'\left(\frac{\rho NT^2 - \rho N^2T^3}{\mu_{eff}}, \frac{NT}{V}, \frac{\rho N^2T^3}{\sigma}\right)^{1/3}$$ (19)

In Table 6 the values of the constants obtained by regression analysis of data are given, where the values of the constants $C$ and $C'$ depend on the number and stirrer type and the rheological model used.

Measured values of experimental Sherwood number were plotted against the Sherwood number values calculated by
4. Conclusions

The volumetric mass transfer coefficient has been measured in a stirred tank reactor with different types of paddle and disk turbine stirrers. It has been observed that, for both the paddle and the disk turbine stirrers, an increase in the number of blades produces an increase in the volumetric mass transfer coefficient value. Also, it has been seen that the disk turbine is more effective than the paddle stirrer.

When the viscosity of the liquid is low, an increase in the number of stirrers produces an increase in $k_L a$ value; however, for high viscosity values, the number of stirrers does not influence the mass transfer rate. With respect to the sparger, it has been observed that under the conditions studied the choice has little effect on the mass transfer coefficient.

For xanthan gum solutions, the values of the volumetric oxygen transport coefficient can be related to the operational variables (stirrer speed, superficial gas velocity and effective viscosity of the liquid) according to an exponential function, with exponents values of 2/3, 2.0 and $-2/3$, respectively, when the Ostwald-de Waele model is used to described the rheological behaviour of the liquid. The last exponent changes if the Casson model is employed for rheology description, taking the value $-1.0$.

If $P/V$ is employed instead of $N$, then the exponents take the following values: 2/3, 0.6 and $-2/3$ when power-law model is used, and 2/3, 0.6 and $-1.0$ when the Casson model was employed.

A dimensionless correlation has been found for the prediction of $k_L a$ for Newtonian and non-Newtonian liquids, expressed as the Sherwood number as a function of the Reynolds, the Airation and the Weber numbers, in both cases using Ostwald-de Waele and Casson models. The correlation proposed is able to give the $k_L a$ values with an error around ±10%, always less than 15%.

5. Nomenclature

<table>
<thead>
<tr>
<th>Symbol</th>
<th>Description</th>
</tr>
</thead>
<tbody>
<tr>
<td>$A_s$</td>
<td>Specific interfacial area, m$^{-1}$</td>
</tr>
<tr>
<td>$C, C'$</td>
<td>Constants in Eqs. (3), (17)–(19)</td>
</tr>
<tr>
<td>$C_{1.2}$</td>
<td>Constants in Eqs. (1), (2), (12)–(15)</td>
</tr>
<tr>
<td>$C_L$</td>
<td>Oxygen concentration in the liquid bulk, mol m$^{-3}$</td>
</tr>
<tr>
<td>$C^*$</td>
<td>Oxygen concentration at gas–liquid interface, mol m$^{-3}$</td>
</tr>
<tr>
<td>CMC</td>
<td>Carboxy-methyl-cellulose</td>
</tr>
<tr>
<td>$D$</td>
<td>Vessel diameter, m</td>
</tr>
<tr>
<td>$D_L$</td>
<td>Oxygen diffusivity coefficient of the gas in the liquid, m$^2$s$^{-1}$</td>
</tr>
<tr>
<td>$d$</td>
<td>Thickness of the blade of the stirrer, m</td>
</tr>
<tr>
<td>$g$</td>
<td>Gravitational acceleration, m s$^{-2}$</td>
</tr>
<tr>
<td>$l$</td>
<td>Blade width, m</td>
</tr>
<tr>
<td>$K$</td>
<td>Parameter of the equation of Meztner–Otto</td>
</tr>
<tr>
<td>$k$</td>
<td>Consistency index in a power-law model, Pa s$^{0.6}$</td>
</tr>
<tr>
<td>$k_L$</td>
<td>liquid–phase mass transfer coefficient, m s$^{-1}$</td>
</tr>
<tr>
<td>$k_L a$</td>
<td>Volumetric oxygen transfer in liquid side, m s$^{-1}$</td>
</tr>
</tbody>
</table>
$k_c$ Parameter of the rheological model of Casson, N$^{1/2}$ g$^{1/2}$ m$^{-1}$

$M$ Hub out diameter, m

MSPA Modified sodium poly-acrylate

$N$ Stirrer speed, s$^{-1}$

$N_a$ Aireation number (NT/V$_r$)

$n$ Flow index in a power-law model

$P$ Power input at gassed conditions, W

$r$ Horizontal blade length, m

$Re$ Reynolds number ($\rho N T^2 / \mu _{eff}$)

$T$ Stirrer diameter, m

$t$ Time, s

$V$ Volume of the liquid in the vessel, m$^3$

$V_g$ Superficial gas velocity, m s$^{-1}$

$W$ Width of the disk, m

$We$ Weber number ($\rho N^2 T^3 / \sigma$)

$x_{1,2,3}$ Exponents of the dimensionless numbers in Eqs. (3) and (17)

Greek Letters

$\alpha$ Exponent of the superficial gas velocity

$\beta$ Exponent of the stirrer speed

$\beta^*$ Exponent of the power input per unit volume

$\lambda$ Exponent of the viscosity

$\gamma$ Shear rate, s$^{-1}$

$\delta$ Exponent of the power dissipated by the stirrer

$\mu$ Newtonian viscosity, Pa s

$\mu_a$ Apparent viscosity at 30 rpm, Pa s

$\mu_c$ Viscosity according to the Casson model, Pa s

$\mu _{eff}$ Effective viscosity, Ostwald-de Waele model, Pa s

$\rho$ Liquid density, kg m$^{-3}$

$\sigma$ Interfacial tension, N m$^{-1}$

$\tau$ Shear stress, N m$^{-2}$

$\tau_0$ Yield stress, N m$^{-2}$

Key for stirrer type

CBT Curved blade turbine (4CBT, with 4 blades; 6CBT, with 6 blades; etc.)

FBT Flat blade turbine (4FBT, with 4 blades; 6FBT, with 6 blades; etc.)

FBP Flat blade paddle (4FPB, with 4 blades; 6FBP, with 6 blades; etc.)

PBP Pitched blade paddle (4PBP, with 4 blades; 6PBP, with 6 blades; etc.)

Two-6FBT Two stirrers of the type indicated (6FBT, in the example)

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References


