

# MASS TRANSFER COEFFICIENTS IN BATCH AND CONTINUOUS REGIME IN A BUBBLE COLUMN

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## ABSTRACT

The physical absorption has been experimentally investigated in a bubble column for to determination of volumetric mass transfer coefficients referred to the liquid phase,  $k_L a$ , and quantify the effect of the viscosity of the absorbent solution on the same, operating in continuous and batch regime with respect to the liquid phase. The variables studied in the process have been the physical properties of the liquid phase (density and viscosity) and the physical properties of the gas phase (diffusivity and solubility), the gas flow rate and the pore diameter as well as the liquid flow rate. The experimental results have been correlated through two equations, depending on the flow regim. Both equations reproduce the experimental data with an average deviation of 10%.

## INTRODUCTION

Mass transfer process in gas-liquid systems is an elemental field in chemical engineering, both on industrial and laboratory level. In the former case, this process was carried out using specific apparatus, but it is very difficult to predict its behaviour only with theoretical models [1, 2, 3]. Due to it, it is necessary to evaluate the volumetric mass transfer coefficients or the specific interfacial areas at the laboratory level. Many authors had developed different models in order to simulate industrial absorbers, i.e. apparatus with the wall-wetted column [4, 5, 6, 7], the rotator tambour [8], the mobile band absorbed [9], the laminar jet apparatus [10] and the bubble column [11].

The behaviour of an industrial column is similar to the laboratory model when for a same solution and equal operation conditions, the volumetric volumetric mass transfer coefficient ( $k_L a$ ) remains constant. When we perform absorption studies with pure gas, i.e. without resistance to mass transfer in gas phase, it is not possible to determine the liquid side mass transfer coefficient ( $k_L$ ) and the specific interfacial

area (a) independently. Then the volumetric mass transfer coefficient ( $k_L a$ ) is obtained through absorption experiments that have a character purely physical (physical method).

Bubble columns are important among the different gas-liquid contactors because they present some advantages respect other contactors, like minor cost of maintenance, effective interfacial areas and very high mass transfer coefficients, big versatility of operation and high residence times. This type of contactors is widely used in the chemical industry as absorbers, fermenters and anaerobic and aerobic biology reactors. The design parameters of these apparatus [4, 5, 12, 13, 14], are: flow regime, coalescence and bubble size, hold-up, mass transfer coefficient and interfacial gas-liquid area. Further, there are some factors which influence the hydrodynamic of the system, like the physical properties of the contact phases, the operation parameters (gas flow, liquid flow and liquid volume) and geometric parameters (diameter and height of the column, height of the liquid in the column)

Different authors show that the volumetric mass transfer coefficient ( $k_L a$ ) is a parameter related directly with superficial velocity and physical properties of liquid phase. However, the relation between these parameters is distinct from some authors to others (Table 1). This fact can be attributed to the different plates and the different range of variables [12, 17, 18, 19]. Most part of authors reproduce the experimental results using equations with the physical properties of the solutions, with the physical properties of the gases, and other characteristics of the systems. Other authors use correlations with dimensionless numbers [12, 13, 18, 20]. It is possible to classify both types of equations in two big groups: equations that depend on hold-up [12, 14, 17] and equations that are non-dependent on this parameter [20, 21, 22].

Some authors [23, 24] correlated the  $k_L a$  values with the superficial gas velocity. In these equations the influence of the physical properties is reflected in the constant of the correlation. Taking into account the diversity of correlations that exist in the bibliography, we have opted for the experimental determination of the volumetric mass transfer coefficient, as well as the obtainment of the empirical equations that reproduce the behavior of our system.

In this work we will present the results obtained upon accomplishing physical absorption experiments in batch and continuous regime. We have correlated  $k_L a$  with the superficial gas velocity, with the physical properties of the liquid phase, with the liquid flow rate and with the bubble size. We have employed as liquid phase aqueous sucrose solutions (0-135 g/l), and as gas phase pure  $\text{CO}_2$  in the different experiments.

## **MATERIALS AND METHODS**

### **Mass transfer measurements**

Mass transfer measurements were carried out using the apparatus shown in Fig.1. Except for the contact device, this set-up has been described in detail elsewhere [25]. The bubble column used as contact device in this work, 1, was made from two metacrylate cylinders. The internal diameter of the column is 11.3 cm, the external

diameter is 12.1 cm and the height is 108.6 cm. In order to make all absorption experiments to the same temperature, a thermostated liquid pass between the two cylinders of the column. The top plate is flat and has three orifices, a central orifice for a thermometer, 2, two off-centre orifices for inflow of liquid, 3, and outflow of gas, 4.

Table 1. Previous correlations for volumetric mass transfer coefficient in a bubble column

Author	System	Operating Variables	Equation
<b>Akita</b> <sup>[12]</sup> (1973)	H <sub>2</sub> O, glycol, methanol Na <sub>2</sub> SO <sub>4</sub> , CCl <sub>4</sub> , NaCl, O <sub>2</sub> , Air, Helium, CO <sub>2</sub>	$u_G=5.3 \cdot 10^{-3}$ - $4.2 \cdot 10^{-1}$ m/s $d_c=15$ - $60$ cm T= $40$ - $10$ °C Single-orifice sparged	$k_L a = c_2 D_L^{0.50} g_L^{-0.12} \left( \frac{\sigma}{\rho_L} \right)^{-0.62} D^{0.17} g^{0.93} \epsilon_g^{1.10}$ $Sh=c_2 Sc^{0.50} Bo^{0.62} Ga^{0.31} \epsilon_g^{1.10}$
<b>Akita</b> <sup>[18]</sup> (1974)	H <sub>2</sub> O, glycol, methanol, CCl <sub>4</sub> , Sodium sulphite Air, O <sub>2</sub>	$u_G=3 \cdot 10^{-3}$ - $4 \cdot 10^{-1}$ m/s $d_c=7.7$ - $15$ cm T= $5$ - $40$ °C Single-orifice sparged	$k_L=0.5 g^{0.60} D_L^{0.50} \rho_L^{0.37} \sigma^{-0.37} d_{vs}^{0.50}$ $Sh=0.5 Sc^{0.50} Bo^{0.37} Ga^{0.25}$
<b>Hikita</b> <sup>[21]</sup> (1981)	H <sub>2</sub> O, Sucrose, butanol, methanol, Na <sub>2</sub> SO <sub>4</sub> , KCl, K <sub>2</sub> SO <sub>4</sub> , K <sub>3</sub> PO <sub>4</sub> , KNO <sub>3</sub> , AlCl <sub>3</sub> , CaCl <sub>2</sub> Air, O <sub>2</sub> , H <sub>2</sub> , CH <sub>4</sub> , CO <sub>2</sub>	$u_G=4.2 \cdot 10^{-2}$ - $3.8 \cdot 10^{-1}$ m/s $d_c=10$ - $150$ cm, T= $10$ - $30$ °C Contin.-semicont Reg. Single-orifice sparged	$k_L a = 14.9 D_L^{0.60} \sigma^{-1.02} \mu_L^{-0.08} \mu_g^{0.24} \rho_L^{0.85} g^{0.75} u_G^{0.76}$
<b>Hughmark</b> <sup>[17]</sup> (1967)	Water, glycerine Air, O <sub>2</sub>	$u_G=4 \cdot 10^{-2}$ - $4.5 \cdot 10^{-1}$ m/s Multiple-orifice sparged $d_c > 10$ cm	$k_L = 0.65 \frac{D_L \rho_L^{0.2}}{\sigma^{0.6}} + \frac{0.051 D_L^{0.04} \rho_L^{1.68}}{\sigma^{0.06} \epsilon_g^{0.78} U_L^{0.779}}$ $\frac{U_g^{0.31} g^{0.04} \mu_L^{0.88}}{\sigma^{0.06} \epsilon_g^{0.78} U_L^{0.779}}$
<b>Kawase</b> <sup>[13]</sup> (1987)	Water, carbopol 1, 2, Sodium carboxymethylcellulose, CO <sub>2</sub>	$u_G=1 \cdot 10^{-2}$ - $1 \cdot 10^{-1}$ m/s Multiple-orifice sparged $d_c=14$ - $762$ cm	$k_L a = 8.35 \cdot 10^{-4} u_G^{0.44} \mu_a^{-1.01}$ $Sh = 0.646 Sc^{0.5} Bo^{0.62} Fr^{0.057} Re^{0.62}$
<b>Kawase</b> <sup>[20]</sup> (1996)	Water, glycerine, Sodium carboxymethylcellulose Air	$u_G=1 \cdot 10^{-2}$ - $1 \cdot 10^{-1}$ m/s Multiple-orifice sparged $d_c=23$ cm	$k_L a = 2.14 \cdot 10^{-3} u_G^{0.52} \mu_a^{-0.89}$ $Sh = 0.142 Sc^{0.50} Bo^{0.60} Fr^{0.07} Re^{0.87}$
<b>Nakanoh</b> <sup>[19]</sup> (1980)	Water, Sucrose, Sodium carboxymethylcellulose, O <sub>2</sub>	$u_G < 1 \cdot 10^{-1}$ m/s Multiple-orifice sparged $d_c=14.55$ cm, T= $30$ °C	$Sh = 0.09 Sc^{0.5} Bo^{0.75} Fr^{1.0} Ga^{0.39}$
<b>Sada</b> <sup>[24]</sup> (1985)	NaCl, NaOH, Ca(OH) <sub>2</sub> CO <sub>2</sub> , O <sub>2</sub>	$u_G=1 \cdot 10^{-2}$ - $1 \cdot 10^{-1}$ m/s Multiple-orifice sparged $d_c=5$ cm, T= $35$ °C	$k_L a = c u_G^{0.86}$ <b>c = f (solution)</b>
<b>Sotelo</b> <sup>[14]</sup> (1988)	Sucrose Glycerine, Water, Ethanol Air, CO <sub>2</sub>	Multiple-orifice sparged T= $25$ °C $u_G=6.4 \cdot 10^{-4}$ - $4.9 \cdot 10^{-3}$ m/s	$k_L a = b u_G^n$ <b>b, n = f (solution)</b>
<b>Zhao</b> <sup>[22]</sup> (1994)	Sucrose, CMC, oil, water, SAE CO <sub>2</sub>	Multiple-orifice sparged $d_c=14$ cm $u_G=0.8 \cdot 10^{-2}$ - $6.5 \cdot 10^{-3}$ m/s	$k_L a = b u_G^{0.80} \mu^{-0.37} H^{-0.50}$

The baseplate has three orifices, a central orifice, 5, for inflow of gas through a porous plate 4 cm in diameter, 10, (plate 1: equivalent pore diameter  $150\text{-}200 \cdot 10^{-6} \text{ m}$ ; plate 2: equivalent pore diameter  $90\text{-}150 \cdot 10^{-6} \text{ m}$ ; plate 3: equivalent pore diameter  $40\text{-}90 \cdot 10^{-6} \text{ m}$ ) and two off-centre orifices for a thermometer, 6, and outflow of the liquid phase, 7.

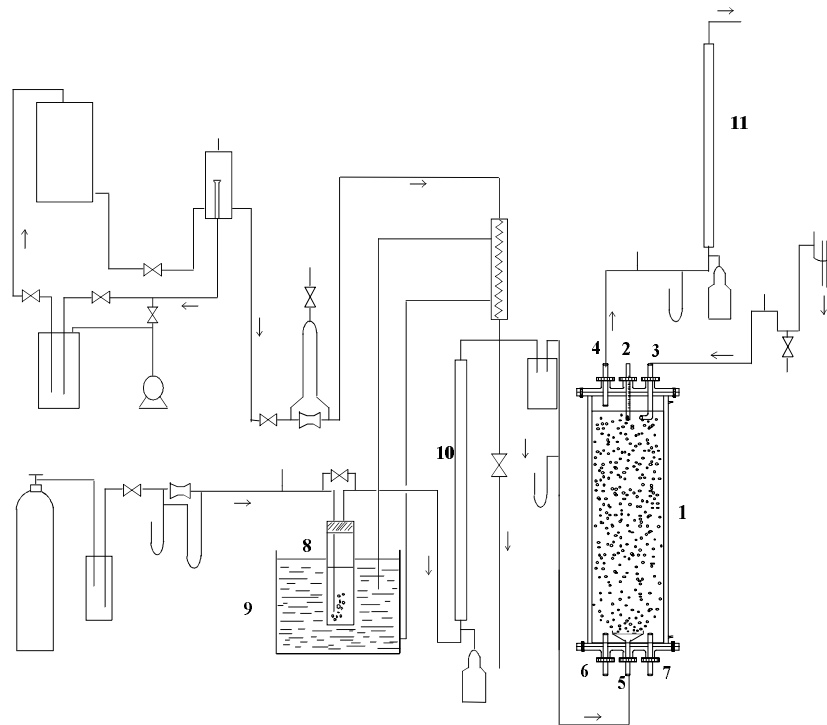


Figure 1. Experimental set-up for measuring absorption of gas.

The gas to be absorbed, pure  $\text{CO}_2$ , was passed through a humidifier at  $25^\circ\text{C}$ , 8, into a thermostated, 9, and entered the contact device at a constant flow rate measured with a bubble flow-meter, 10. Gas outflow was measured with another bubble flow-meter, 11, before its release into the atmosphere. The gas absorption rate was calculated as the difference between inflow and outflow rates. We had employed different gas flow oscillating between  $3 \cdot 10^{-4}$  and  $9 \cdot 10^{-4} \text{ mol/s}$ . The liquid phases used in this work (water and aqueous solutions of sucrose of concentrations up to  $135 \text{ g/L}$ ) were thermostated to room temperature ( $25^\circ\text{C}$ ) before entering the contact device. For batch runs the liquid load was  $12 \text{ L}$ . The contact between the liquid phase and gas phase was in counter current flow.

### Physical properties

Interpretation of the mass transfer data obtained as described above required knowledge of the densities, viscosities and surface tensions of the liquid phases used, and the diffusivities and solubilities of the gas in these phases. The densities and viscosities of the sucrose solutions were measured, at  $25^\circ\text{C}$ , by a pycnometric method and a Schott capillary viscometer (model VAS 350), respectively. The surface tensions were measured by means a Prolabo tensiometer, which uses the Wilhelmy plate method. The diffusivity was calculated from the Joosten and Danckwerts equation [26] obtained different values with sucrose concentration, and

the solubility of CO<sub>2</sub> was calculated from the equation proposed by Linke [27]. The obtained values were shown in the Table 2.

Table 2. Values of physical properties.

	$\rho$ (kg/m <sup>3</sup> )	$\mu \cdot 10^3$ (Pa·s)	$\sigma \cdot 10^3$ (N/m)	$D \cdot 10^9$ (m <sup>2</sup> /s)	$C^e \cdot 10^2$ (mol/L)
Water- CO <sub>2</sub>	997.00	0.896	72.00	1.897	3.36
Sucrose (16 g/L)-CO <sub>2</sub>	1003.14	0.923	72.09	1.768	3.33
Sucrose (50 g/L)-CO <sub>2</sub>	1016.19	1.015	72.29	1.664	3.28
Sucrose (85.7 g/L)-CO <sub>2</sub>	1029.90	1.135	72.49	1.549	3.23
Sucrose (135 g/L)-CO <sub>2</sub>	1048.83	1.327	72.78	1.401	3.15

## THEORY

The calculation procedure followed for the two operation regimes is the one which is shown below.

### Batch processing

The amount of gas absorbed per unit time per unit of liquid phase volume (N) is given by

$$N = \frac{dC}{dt} = k_L a (C^e - C) \quad (1)$$

where the driving force for the transfer process ( $C^e - C$ ) is calculated as the difference between the concentration of the gas in the interface,  $C^*$ , in equilibrium with pure gas (solubility) and the concentration of the gas in bulk liquid. This concentration was evaluated through a mass balance of the experimental absorption rates with the time. Dependence of N with the time for different gas flow rates is shown in Fig. 2, as an example. The volumetric mass transfer coefficient is calculated as the amount of gas absorbed in function of the time, if the volume of the column and the concentrations of the gas at the interface and in the bulk liquid are known.

### Continuous processing

The absorption rate is defined by Eq. (1). As the concentration in the liquid bulk varies throughout the column its value is calculated through a logarithmic average. The volumetric mass transfer coefficient is then:

$$k_L a = \frac{N}{C_1} \ln \frac{C^e}{C^e - C_1} \quad (2)$$

In comparison with the batch regime if we consider  $C_1/N$  represents the contact time. In this case is obtained a single coefficient from each stationary trial, while of the batch system is obtained aa averaged coefficient from each trial, which corresponds to all process variations with the time.

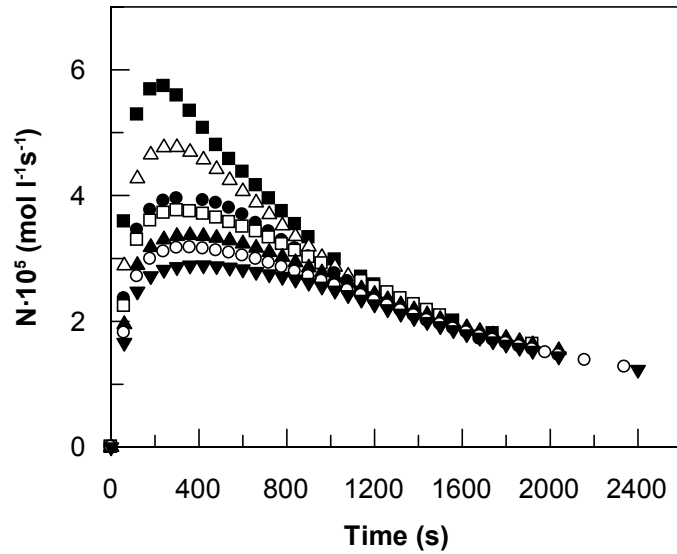


Figure 2. Time-dependence of  $N$  for different gas flow rates. plate 1 and  $[sucrose]=50$  g/L.  
 (■)  $q_G = 8.281 \cdot 10^{-4}$  mol/s; ( $\Delta$ )  $q_G = 6.662 \cdot 10^{-4}$  mol/s; (●)  $q_G = 5.853 \cdot 10^{-4}$  mol/s;  
 (□)  $q_G = 5.011 \cdot 10^{-4}$  mol/s; ( $\blacktriangle$ )  $q_G = 4.516 \cdot 10^{-4}$  mol/s; (○)  $q_G = 4.263 \cdot 10^{-4}$  mol/s.  
 ( $\blacktriangledown$ )  $q_G = 3.848 \cdot 10^{-4}$  mol/s.

As it was observed in batch regime, the value of  $N$  varies with the gas flow rate and furthermore it is seen affected by the liquid flow rate as can be seen in the Fig. 3.

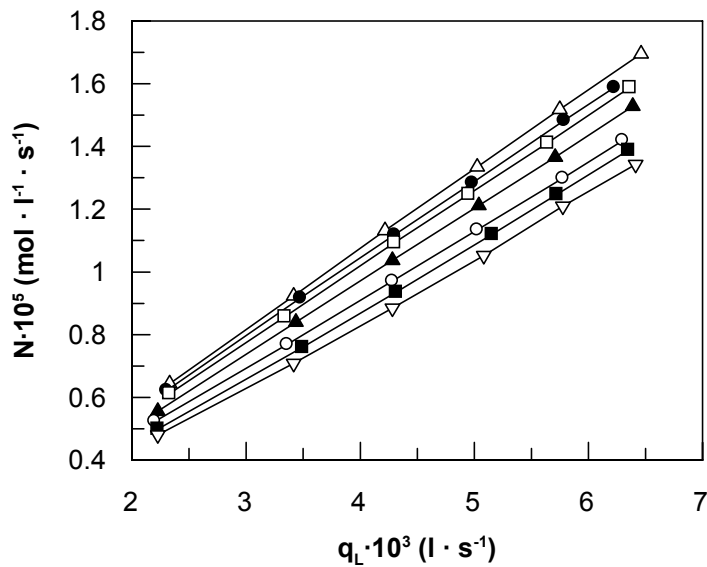


Figure 3.  $q_L$ -dependence of  $N$  for different gas flow rates.  $[Sucrose]=16$  g/L, plate 2.  
 (○)  $q_G = 8.204 \cdot 10^{-4}$  mol/s; (●)  $q_G = 6.907 \cdot 10^{-4}$  mol/s; (□)  $q_G = 5.983 \cdot 10^{-4}$  mol/s;  
 (■)  $q_G = 5.329 \cdot 10^{-4}$  mol/s; ( $\Delta$ )  $q_G = 4.723 \cdot 10^{-4}$  mol/s; ( $\blacktriangle$ )  $q_G = 4.333 \cdot 10^{-4}$  mol/s;  
 ( $\blacktriangledown$ )  $q_G = 3.910 \cdot 10^{-4}$  mol/s

## RESULTS AND DISCUSSION

### Comparison of batch regime with continuous regime

When they are used liquid phases of variable concentrations, the density and viscosity of the same is modified and, consequently, the liquid flow rate that enters the column changes, at the same pressure drop. In order to homogenise the treatment of the experimental results, it has been opted for using the manometric height in place of the liquid flow rate in order to identifying. The not stationary period of the continuous regime of this process. A series of previous trials were made in those which, yet existing liquid flow rate, the amount of gas absorbed is measured in function of the time from the moment in which both phases were put in touch. The process evolves until reaching the stationary regime (witness of the values with the time).

For any one of these processes the number of absorbed moles,  $w$ , varies with the time until reaching a constant value, such as is observed in the Fig. 4 for different conditions of operation. The evolution of these systems is analogous to the one which corresponds to the not stationary regime though in this case the absorption intensity arrives to be annulled, as is logical, when the liquid phase is saturated. These trials permit us to assure that the stationary regime is reached with times of the order of 1800 s, therefore the measurements were accomplished always letting to pass at least this period in continuous regime.

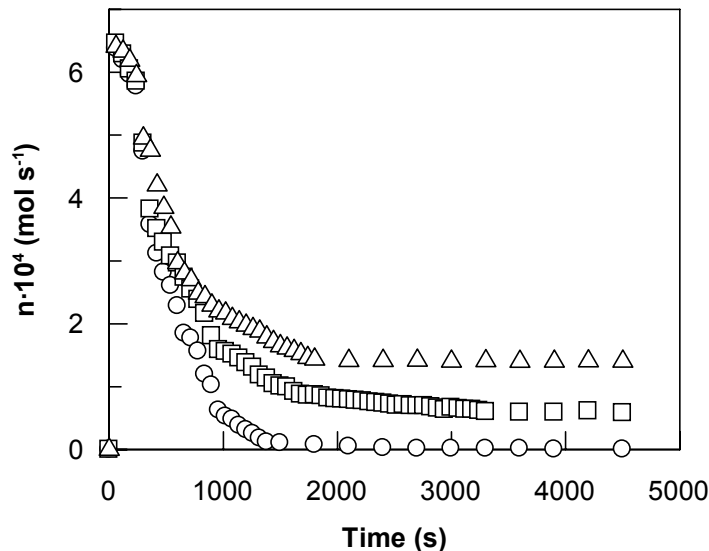


Figure 4. Time-dependence of  $n$  for different manometric heights,  $q_G = 6.500 \cdot 10^{-4}$  mol/s. Plate 0. (O)  $\Delta H = 0$  mm; (□)  $\Delta H = 50$  mm; (Δ)  $\Delta H = 250$  mm.

A similar behavior is appreciated if we represented the number of absorbed moles per unit time per unit of liquid phase volume for the not stationary process of the continuous regime. The absorption velocity goes through a maximum for the different gas flow rates and liquid flow rates, even in the batch process. The extent of the maximum is less pronounced for high gas flow rates. The comparison between the continuous and batch regime can be seen in Fig. 5. There are presented the results of the batch process for three gas flow rates, in the form of values of  $N$  (mol/Ls),

obtaining coincident results to high values of the gas flow rate. The values reached in stationary regime, for the same gas flow rate, are located in the same figure for the time that corresponds to the residence time.

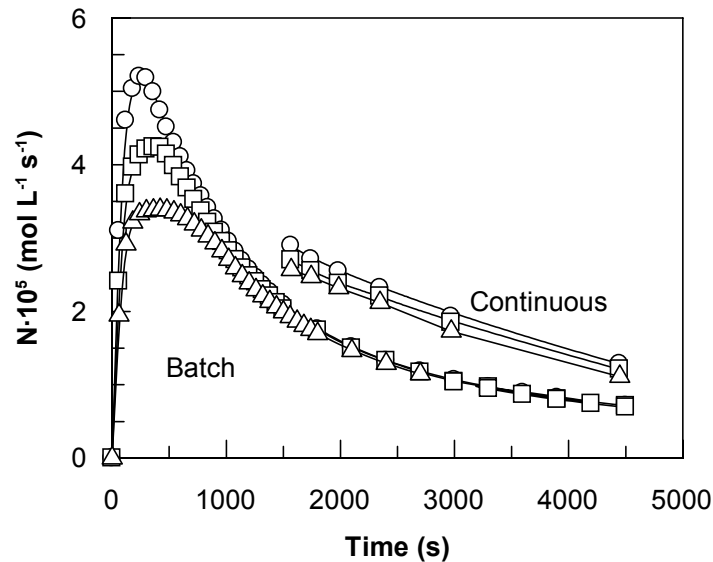


Figure 5. Time-dependence of  $N$  for different gas flow rates and operation regimes.  
 (O)  $q_G = 7.00 \cdot 10^{-4}$  mol/s; (□)  $q_G = 5.10 \cdot 10^{-4}$  mol/s; (Δ)  $q_G = 4.01 \cdot 10^{-4}$  mol/s.

For a volume of 10.29 L, that always it is constant, and different liquid flow rates different we have different residence times (Table 3).

Table 3. Values of residence time for different liquid flow rates

$q_L$ ( $m^3 s^{-1}$ )	$\tau$ (s)
2.314	4447
3.463	2972
4.385	2347
5.184	1985
5.903	1743
6.564	1568

To the sight of the Fig. 5, it is possible to think that the absorption velocity is greater in continuous regime, but this is not exact, since the absorption velocity is obtained in a given time (residence time) in continuous regime and it is a prompt value, while the absorption velocity in batch regime is the value middle from the initial instant until the moment in which is indicated. However the differences between the absorption rates do not justify the differences that result for the mass transfer coefficients, since they would have to coincide to be obtained with the same driving force. The complexity of the flow in the hidrodynamic analysis of the system can justify the differences of these coefficients.



## Batch regime

We have accomplished absorption trials in water to see the behavior of the system. The trials in batch regime have been extended initially until a total time of 5000 seconds. To the sight of the results, it has been taken the decision of reducing the time of the trial until a maxime value of 1800 seconds (30 minutes), the value of the concentration in the bulk liquid stays practically constant since that time.

The volumetric mass transfer coefficients have been evaluated following the method described in the previous paragraph. The values of the mass transfer coefficients vary for each plate with the gas superficial velocity and the sugar concentration that modifies the physical properties of the absorbent solution, fundamentally the viscosity. The variation of  $k_L a$  with this property for the different gas flow rates is presented in the Fig. 6. It is observed that the coefficient decreases with viscosity and increases with the gas flow rate. In this figure the effect of the pore size is not be shown, but we have found a variation with the pore size: the volumetric coefficient increases when the pore diameter decreases.

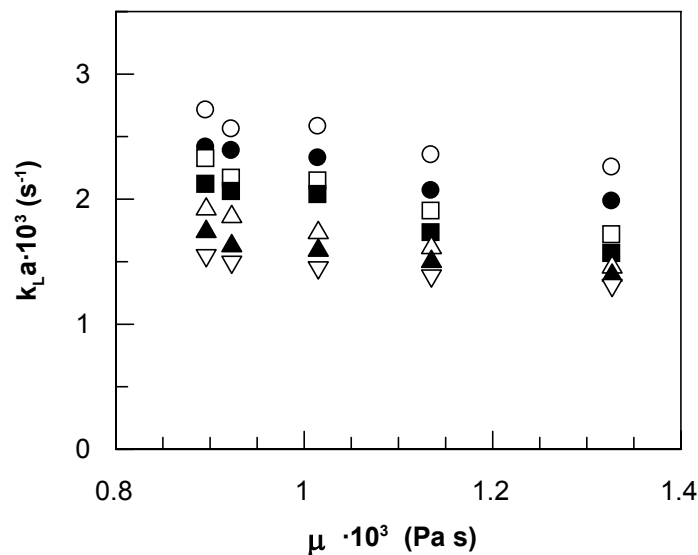


Figure 6. Dependence of  $k_L a$  on viscosity for different superficial gas velocities. Plate 2.  
 (○)  $u_G = 1.930 \cdot 10^{-3}$  m/s; (●)  $u_G = 1.529 \cdot 10^{-3}$  m/s; (□)  $u_G = 1.333 \cdot 10^{-3}$  m/s;  
 (■)  $u_G = 1.175 \cdot 10^{-3}$  m/s; (△)  $u_G = 1.040 \cdot 10^{-3}$  m/s, (▲)  $u_G = 0.953 \cdot 10^{-3}$  m/s;  
 (▽)  $u_G = 0.863 \cdot 10^{-3}$  m/s.

In order to quantify the influence of each variable on the volumetric mass transfer coefficient, it has been opted for correlating the values of obtained  $k_L a$  through an equation from the type: being the parameter value  $K_1$ , equal to  $2.45 \text{ m}^{-3.42} \text{ kg}^{0.77} \text{ s}^{0.63}$  independently of the employed porous plate. The deviation of the results calculated with respect to the experimental values is, in all the cases, inferior to the 9.6 %.

$$k_L a = K_1 \rho^{-0.82} \mu^{-0.46} u_g^{0.69} \sigma^{0.70} d_p^{-0.19} \quad (3)$$

In the Fig. 7 is represented our correlation together with correlations of other authors being shown the difference of employing a correlation to other.

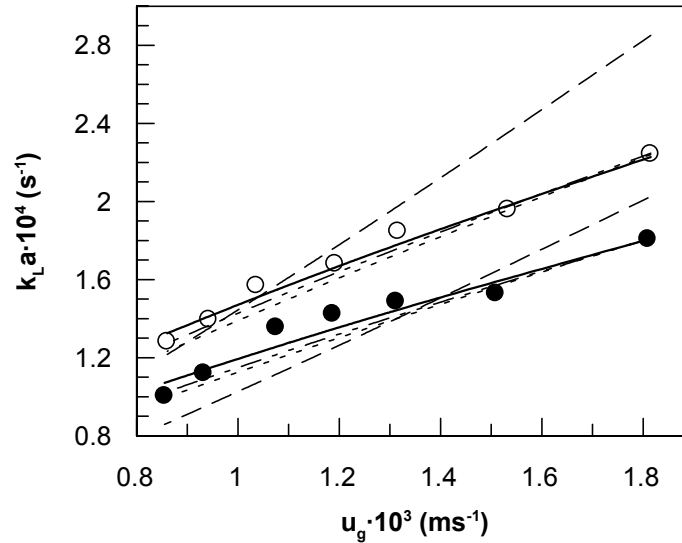


Figure 7. Comparison of the volumetric transfer coefficient data with proposed correlation. (O) experimental data of  $k_L a$  for [sucrose] = 16 g/L; (●) experimental data of  $k_L a$  for [sucrose] = 135g/L; (—) our correlation; (····) Sotelo correlation<sup>[14]</sup>; (- - -) Zhao correlation<sup>[22]</sup>; (— · —) Hikita correlation<sup>[21]</sup>;

Values of the  $k_L a$  estimated for our equation showed good agreement with the observed  $k_L a$  values, and values of the  $k_L a$  estimated for equation of other authors, in some cases, showed disagreement with observed  $k_L a$  values.

### Continuous regime

The values of the coefficient,  $k_L a$ , are determined through the equation (2). The values of these coefficients increase with the gas and liquid flow rates. The increase in the sugar concentration (and by so much of the viscosity of the solution) provokes, as is expected, a meaningful decrease of the coefficients but also the such representations slope increases considerably. So the increase in the slope is appreciated in function of the pore diameter and of the gas flow rate, therefore to high gas flow rate must be compensated the negative effect in the increase of the viscosity. This must be the cause of the difficulty of the adjustment of the liquid flow rate as variable. In the Fig. 8 is represented  $k_L a$  for the different used porous plates, and for one of the seven used gas flow rate as compared to the viscosity of the liquid phase. Observing the graph can be appreciated that the coefficient always decreases when the viscosity increases.

After proving the influence of the physical properties of the absorbent solution as well as that of the liquid and gas flow rates, the  $k_L a$  obtained values are correlated through a similar equation to the equation (3) used in batch regime. The new equation includes as variable the liquid flow rate,  $q_L$ . These results show that the liquid flow rate is an important variable in the continuous absorption regime with

$$k_L a = K_2 \rho^{-0.82} \mu^{-0.46} \sigma^{0.70} u_g^{0.69} d_p^{-0.19} \quad (4)$$

equal exponents to that of batch regime.

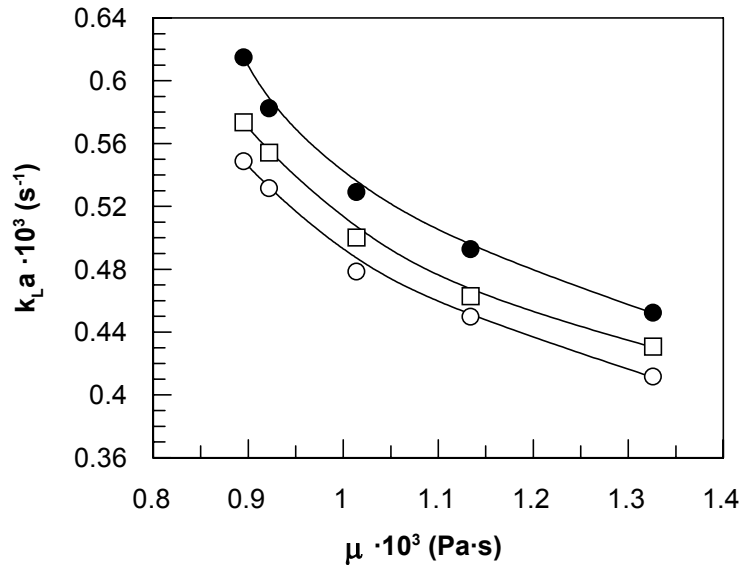


Figure 8. Dependence of  $k_L a$  on viscosity.  $U_g = 1.853 \cdot 10^{-3}$  m/s;  $q_L = 4.437 \cdot 10^{-3}$  m/s; (O) plate 0; (□) plate 1 (●) plate 2.

The variation observed in  $K_2$  is linear with respect to the liquid flow rate, this variation can be attributed to the modification that is produced in the hydrodynamics of the system due to the traffic of the liquid in concurrent with the gas, something that it was not occurring in the batch regime. The relationship is expressed by the following equation:

$$K_2 = d \cdot q_L + e \quad (5)$$

the values of the constants  $d$  and  $e$  are  $135 \text{ m}^{-6.42} \text{ kg}^{0.58} \text{ s}^{1.63}$  and  $0.015 \text{ m}^{-3.42} \text{ kg}^{0.58} \text{ s}^{0.63}$  respectively. The deviation of the results  $k_L a$  calculated with respect to the experimental values is, in 100% of the cases, inferior to the 9.8%.

## CONCLUSIONS

After analyzing the experimental results obtained operating in continuous regime and batch regime in a bubble column, it has been found a capable equation of reproducing the behavior of the system based on the physical properties of the absorbent solution, the gas and liquid flow rates and the pore size ( $d_p$ ). The advantage of this equation as compared to the most of the proposals additionally authors is that is capable to try the different operation regimes with the same efficiency.

## NOTATION

$a$	Effective interfacial area, ( $\text{m}^2$ )
$C$	Gas concentration in the liquid phase, ( $\text{mol L}^{-1}$ )
$C^e$	Gas solubility in the liquid phase, ( $\text{mol L}^{-1}$ )
$d$	Parameter in Eq. (5), ( $\text{m}^{-6.42} \text{ kg}^{0.58} \text{ s}^{1.63}$ )
$d_p$	Pore diameter, (m)

D	Gas diffusivity in the liquid phase, ( $\text{m}^2 \text{s}^{-1}$ )
$\Delta H$	Manometric height, (mm)
e	Parameter in Eq. (5), ( $\text{m}^{-3.42} \text{kg}^{0.58} \text{s}^{0.63}$ )
$k_L$	Individual mass transfer coefficient, ( $\text{m s}^{-1}$ )
$k_L a$	Volumetric mass transfer coefficient, ( $\text{s}^{-1}$ )
$K_1$	Parameter in Eq (3), ( $\text{m}^{-3.42} \text{kg}^{0.58} \text{s}^{0.63}$ )
$K_2$	Parameter in Eq (4), ( $\text{m}^{-3.42} \text{kg}^{0.58} \text{s}^{0.63}$ )
N	Absorption flux per volume, ( $\text{mol l}^{-1} \text{s}$ )
$q_L$	Liquid flow, ( $\text{mol s}^{-1}$ )
$q_g$	Gas flow rate, ( $\text{mol s}^{-1}$ )
$u_g$	Superficial gas velocity, ( $\text{m s}^{-1}$ )

### ***Greek symbols***

$\mu$	Viscosity of liquid phase, ( $\text{kg m}^{-1} \text{s}^{-1}$ )
$\rho$	Density of liquid phase, ( $\text{kg m}^{-3}$ )
$\sigma$	Surface tension of liquid phase, ( $\text{N m}^{-1}$ )
$\zeta$	Residence time, (s)

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**Keywords:** Mass transfer coefficients; physical absorption, bubble column, continuous regime, batch regime.