



Effect of some solid properties on gas–liquid mass transfer in a bubble column

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ABSTRACT

The knowledge about the effects of solids on gas–liquid systems and the respective physical mechanisms are not yet totally clarified. In this work, the effect of the solids on the mass transfer characteristics in a bubble column was studied experimentally for the systems air/water/expandable polystyrene (EPS) beads and air/water/glass beads. Volumetric liquid side mass transfer coefficient, $k_L a$, was determined under different solid concentrations (up to 30 vol.%), superficial gas velocities (up to 2.7 mm/s) and mean diameters (1100, 770 and 591 μm for EPS and 9.6 μm for glass beads).

The presence of EPS solids affects negatively $k_L a$ being this effect more pronounced for the smaller particles. Also, a decrease in $k_L a$ occurs when the solid loading increases. Experiments done with large polystyrene particles ($d_p \geq 591 \mu\text{m}$) contaminated with very fine EPS particles ($d_p \cong 0.1 \mu\text{m}$) indicate that very fine particles play an important role on gas–liquid mass transfer.

Mass transfer experiments in a hollow glass spheres three-phase slurry showed a dual effect of solids loading on $k_L a$, contrarily to what happens with the previous particles. These results can be associated with the different surface properties of the particles studied.

An empirical correlation for $k_L a$ on the experimental variables was developed.

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1. Introduction

In many operations of chemical industry processes, one or more components of a gas phase are absorbed into a liquid phase. The phase contacting is often required to trigger reactions among components of the two phases. The complexity and diversity of industrial processes implied that different types of gas–liquid contactors were developed and constructed, such as, bubble columns, pipes/tubes, mechanical agitated tanks, packed columns, plate/tray columns, spray towers, jet (loop) reactors, tubular/ventury ejectors and motionless mixers [1]. Bubble columns are contactors in which a gas or a mixture of gases in the form of a dispersed phase of bubbles moves in a continuous liquid phase. In the liquid, a suspended or fluidized, reactive or catalytic solid can also be present. Thus, in fact, two or three-phase bubble columns exist [2]. These columns can be of different types such as single stage, multi stage, multi channel, with motionless mixers, loop reactors, jet reactors, downflow bubble columns, three-phase fluidized-bed reactors and slurry reactors. Their applications are not limited to the chemical industry; they can be found in biochem-

ical operations, separation of mixtures by rectification, absorption, wastewater treatment [3] and petrochemical industry. Bubble columns are also gaining increasing importance in the field of biotechnology [4].

Parameters such as phase holdup, flow regime, bubble size distribution, coalescence characteristics, gas–liquid interfacial area, interfacial mass transfer coefficients, heat transfer coefficients and dispersion coefficients influence the bubble column reactors design. The mass transfer coefficient is believed to be the most important design variable, followed by the gas holdup and the axial dispersion coefficient [5].

In three phase systems, the presence of solids affects the gas–liquid mixture in different ways: bubble formation and rise [6,7], axial and radial profiles [8,9], mixing and dispersion, mass transfer [10–13], and gas holdup and flow regimes [14,15].

Despite all the research efforts, the knowledge about the effects of solids on gas–liquid systems and the respective physical mechanisms are not yet clarified. In the present work, volumetric mass transfer coefficients, $k_L a$, were obtained for different gas flow rates, solids loading and size. These results were used in order to develop an empirical correlation for $k_L a$, taking into account the different experimental variables. The influence of fine particles on $k_L a$ was also analyzed, in order to clarify some unexpected results.

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Nomenclature

a	gas–liquid interfacial area (m^{-1})
a_i	empirical parameter, $i = 1, 2, 3, 4$
b	linear regression slope ($y_{est} = c + bx$)
b_1	linear regression optimum slope for the left hand side
b_2	linear regression optimum slope for the right hand side
C	oxygen concentration in the liquid (kg/L)
c	linear regression parameter ($y_{est} = c + bx$)
C_0	oxygen concentration in the liquid at $t = 0$ (kg/L)
C^*	oxygen solubility in the liquid (kg/L)
d_p	particle mean diameter (μm) (mm in Eq. (5))
e_s	solid volume fraction
F^*	Test F parameter
k_L	liquid-side mass transfer coefficient (m/s)
$k_L a$	volumetric liquid side mass transfer coefficient (s^{-1})
n_p	number of points for linear regression
n_{p1}	number of points for linear regression for the left hand side
n_{p2}	number of points for linear regression for the right hand side
T	temperature ($^{\circ}\text{C}$)
t	time (s)
u_G	superficial gas velocity (m/s)
y	experimental value
y_{est}	estimated value ($y_{est} = c + bx$)

2. Experimental

2.1. Experimental set-up

The contact device used to perform the mass transfer experiments, at atmospheric pressure, was the bubble column represented in Fig. 1, with the respective dimensions. The device is a perspex cylindrical column covered by a perspex rectangular box to control the temperature (25°C) through water circulation. At the bottom, it is located a gas chamber where the gas enters and then passes through a sparger where the bubbles are formed.

The sparger consists of 13 needles with an inner diameter of 0.3 mm. A scheme of the gas sparger is shown in Fig. 2(a). The shape and size (25 mm) of the needles ensure the formation of small and well-defined bubbles. The needles disposal and the movement of the bubbles enable the suspension of low density solids, and the homogeneity in terms of O_2 concentration of the liquid phase. A concave perspex cylindrical piece (Fig. 2(b)), formed by small cones involving each needle, was placed at the column bottom to allow solids circulation in this section avoiding their deposition, and, consequently, the blockage of the orifices.

A flow valve was used to control the air flow rate that was measured by a KDG Mobrey 2-A-150 R rotameter followed by a manometer, where the gauge pressure was kept at the constant value of 1 bar. Before entering the bubble column, the air is saturated in a humidifier. Dissolved oxygen concentration was measured by a Mettler Toledo In Pro 6100 O_2 sensor (polarographic electrode) connected to a Mettler Toledo O_2 4100 dissolved oxygen meter. The oxygen consumption by the electrode is negligible. A data acquisition board was coupled to the O_2 meter, enabling to read the O_2 concentration on-line.

2.2. Mass transfer experiments

Oxygen mass transfer runs were performed in two and three-phase systems. Compressed air from air gas cylinder and tap water were used as gas and liquid phases, respectively. Expandable polystyrene (EPS) particles of three different mean diameters, d_p , (Styropor[®] VEP 124 (P124) – $d_p = 1100 \mu\text{m}$, Styropor[®] VEP 324 (P324) – $d_p = 770 \mu\text{m}$ and Styropor[®] EP 424 (P424) – $d_p = 591 \mu\text{m}$) with a density of $\rho_p = 1040\text{--}1050 \text{ kg/m}^3$, and hollow glass spheres (with $d_p = 9.6 \mu\text{m}$, and a density of $\rho_p = 1100 \text{ kg/m}^3$) were used as solid phase. The experiments were performed for several superficial gas velocities, u_G , (up to 2.7 mm/s), and different solid concentrations (up to 30 vol.%). The clear liquid height was $h_0 = 0.2 \text{ m}$ for all experiments (no liquid throughput) and the polarographic electrode (O_2 probe) was located transversely 0.1 m from the gas sparger and, approximately, 2 cm from the column wall.

The experimental procedure is initiated by bubbling nitrogen to deoxygenate the liquid phase. When the dissolved oxygen concentration is approximately zero, humidified air is fed into the column and then the dissolved oxygen concentration variation with time, t , is obtained.

The mass balance for oxygen in the liquid is written as:

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

where C^* and C are, respectively, the oxygen solubility and oxygen concentration in the liquid. Assuming the liquid phase homogeneous and C_0 the oxygen concentration at $t = 0$, the integration of the previous equation leads to:

$$\ln(C^* - C) = \ln(C^* - C_0) - k_L a t. \quad (2)$$

The volumetric mass transfer coefficient can now be determined by plotting $\ln(C^* - C)$ against time (t). The experimental results are reproducible with an average relative error of 5%. Further, $k_L a$ results were not influenced by the dynamics of the oxygen electrode since, for all runs, it was verified that its response time ($\sim 35 \text{ s}$) was much smaller than the mass transfer time of the system (ranging from 100 to 650 s).

3. Results and discussion

3.1. $k_L a$ evaluation from experimental data

An example of data treatment is presented below, for air–water–P424 beads and superficial gas velocity $u_G = 2.2 \text{ mm/s}$. The dissolved oxygen concentration curve for this particular experiment is plotted in Fig. 3. One can distinguish three zones on the graph. Zone I, at the beginning, where the O_2 concentration is nearly constant, followed by an intense mass transfer zone where the O_2 concentration rises fast (Zone II). The last zone (Zone III) appears close to the saturation, when the mass transfer rate starts to decline.

Plotting $\ln(C^* - C)$ against time (Fig. 4), one can observe that the applicability of Eq. (2) is valid except in Zone I, related with the electrode response. So, $k_L a$ was determined from the slope in the linear zone. The solubility of oxygen in water (C^*) was taken from the literature [16] and the slope was determined using the statistical method Test F [17]. This method consists in determining the optimum number of points (n_p) for a linear regression of the experimental data. Initially, a short data interval is considered and the linear regression parameters (c and b) are determined as well as a parameter F^* . This parameter is defined as:

$$F^* = \frac{1}{n_p - 1} \sqrt{\frac{\sum (y_{est} - y)^2}{\sum y^2}} \quad (3)$$

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