# Gas-Liquid Mass Transfer in a Bubble Column with Suspended Solids

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Nguyen-tien et al. (1985) recently presented a comprehensive study of gas-liquid mass transfer in three-phase fluidized beds. Glass spheres ranging from 0.05 to 8 mm dia. were fluidized in a 0.14 m dia. column H by cocurrent upward flow of water and air. In beds of small particles ($d_s \leqq 1$ mm) not capable of bubble disintegration, the volumetric mass transfer coefficients $k_La$ were independent of the particle size and could be empirically correlated by the superficial gas velocity $u_G$ and the solids volume fraction in the suspension $\phi_s$:

$$k_La = 0.39\left(1 - \frac{\phi_s}{0.58}\right)u_G^{0.67} \tag{1}$$

Equation 1 describes 147 data points for glass spheres ($0.05 \leqq d_s \leqq 1.0$ mm; $\phi_s \leqq 0.29$) with 11% mean deviation. Nguyen-tien et al. suggested the term in brackets to account for the effect of suspension viscosity, which could not be measured.

It is the purpose of this paper to check on the applicability of Eq. 1 to suspension bubble columns, i.e., fluidization by gas-induced liquid motion with no net liquid flow. Specifically, whether the relation holds also for particle mixtures is investigated.

## Experimental

Glass spheres with uniform diameters of 0.05, 0.1, and 0.2 mm, and densities of 2,770, 3,050, and 2,830 kg · m$^{-3}$, respectively, were used. The 0.2 mm spheres were mixed with an equal volume of either the 0.1 mm or the 0.05 mm spheres to form particle mixtures. The measurements were carried out in a jacketed glass column of 0.095 m ID equipped with a perforated plate sparger (19 mm × 1 mm holes). The column was thermostat-controlled to 20 ± 0.5°C and operated at 0.9 m height of the gas-free suspensions. The gas velocities ranged up to 0.1 m · s$^{-1}$. The lower limit was given by minimum fluidization velocities of 0.014, 0.024, and 0.038 m · s$^{-1}$ for the spheres of 0.05, 0.1 and 0.2 mm dia., respectively, reported by Schumpe et al. (1984). For the particle mixtures, the required gas velocities were less than for the 0.2 mm dia. spheres.

The gas holdups were determined from the observed dispersion heights. The axial profile of the solids concentration was investigated by withdrawing a sample of the suspension in the bottom zone 4 cm above the gas distributor. The solids fraction was determined by gravimetric analysis. The sampling was repeated a second time after readjusting the concentration. The measured volume fraction at the bottom, $\phi_{s,o}$, was compared to the average value $\bar{\phi}_s$ calculated from the solids loading. The volumetric mass transfer coefficients were measured by the dynamic method. Nitrogen was sparged to desorb the dissolved oxygen. At disengagement of the last nitrogen bubbles and before significant sedimentation of the particles occurred, a preadjusted air flow was fed into the column by switching two magnet valves. The increase of the liquid phase oxygen concentration was measured with a polarographic oxygen electrode (WTW EO 90) with a response time of about 3 s. It was inserted horizontally at half the dispersion height since this positioning minimizes the effects of the start-up period and imperfect mixing (Linek et al., 1979). The observations at relative saturations in the range of $0.5 < C_E/C_E^* < 0.9$ were fitted to the following relation:

$$\ln(C_E^* - C_E) = -\frac{k_La}{\epsilon_L}t + \text{const} \tag{2}$$

Schumpe (1985) has shown that the results of this procedure agree with a more involved approach considering the electrode response time, and with independent measurements by the steady state method.

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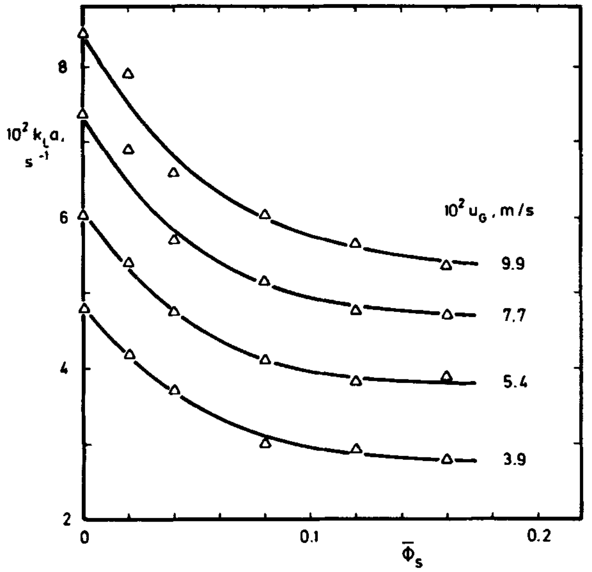

**Figure 1. Volumetric mass transfer coefficients in mixed suspensions of 0.1 and 0.2 mm dia. spheres.**

## Results

With increasing solids loading, the volumetric mass transfer coefficients and gas holdups decrease strongly. The typical trends are demonstrated in Figures 1 and 2 for mixtures of 0.1 and 0.2 mm dia. spheres. There is no $k_La$ increase at small particle concentrations, but a rather steep decline. The slope then decreases, and at mean solids fractions exceeding 10% the volumetric mass transfer coefficients show little dependence on the solids loading. The gas holdups follow the same nonlinear trend, as shown in Figure 2, i.e., the $k_La$ decrease is basically a decrease in interfacial area caused by coalescence to large bubbles with high rise velocities.

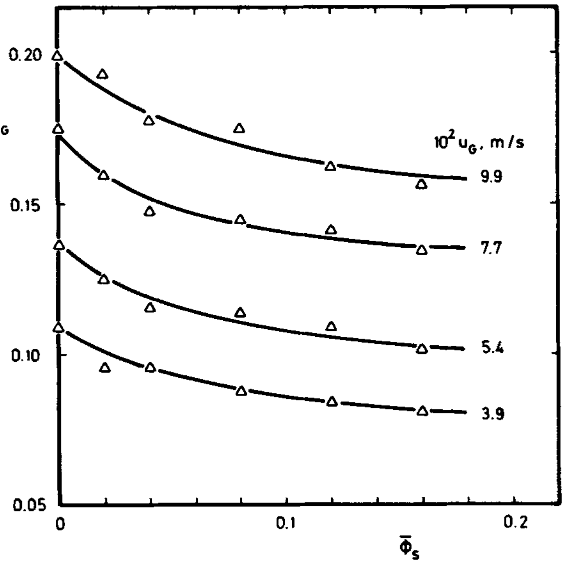

**Figure 2. Gas holdups in mixed suspensions of 0.1 and 0.2 mm dia. spheres.**

## Discussion

From the nonlinearity of the plot in Figure 1 one may conclude that Eq. 1 suggested by Nguyen-tien et al. for three-phase fluidized beds does not hold for suspension bubble columns with batch operation of the liquid phase. Indeed, the predictions of Eq. 1 are high, particularly at intermediate solids concentrations if introducing the mean solids fraction $\bar{\phi}_s$. However, one has to consider that in three-phase fluidization the solids concentration is rather homogeneous in the bulk of the bed and varies only in the upper particle entrainment zone, whereas without net liquid flow a strong axial concentration profile is developed. Schumpe et al. (1984) suggested that in short suspension bubble columns the bottom concentration rather than the mean concentration should be considered. They reasoned that large bubbles formed by coalescence in this zone of high solids concentration, i.e., high effective suspension viscosity, would not break up again during their small residence time. Hence the solids fractions at the gas distributor $\phi_{s,o}$ should be introduced in Eq. 1 rather than the average values.

The measured solids fractions at the bottom are compared with the sedimentation-dispersion model. By integration of the relation for the local solids fraction,

$$\phi_s(z) = \phi_{s,o} e^{-Pe_s z} \tag{3}$$

the value at the bottom ($z = 0$) can be related to the known average solids fraction:

$$\phi_{s,o} = \bar{\phi}_s Pe_s / (1 - e^{-Pe_s}) \tag{4}$$

The measured values were found to be in close agreement to Eq. 4, with Peclet numbers calculated from the relation suggested by Kato et al. (1972):

$$Pe_s = \frac{u_{ss}L}{D_s} = \frac{u_{ss}L}{u_G d_R} \frac{13Fr}{1 + 8Fr^{0.85}} \tag{5}$$

The settling velocity $u_{ss}$ is

$$u_{ss} = 1.2 u_T \left(\frac{u_G}{u_T}\right)^{0.25} \left(\frac{1 - \bar{\phi}_s}{1 - \phi_s^*}\right) \tag{6}$$

where $\phi_s^*$ is the solids volume fraction in the gas-free suspension at a concentration of 0.1 g/cm$^3$.

In the case of particle mixtures the experimental $\phi_{s,o}$ values compare well with the weighted mean of the predictions for the small and the large particles at the same total volume fraction. This is demonstrated in Figure 3 for mixtures of 0.1 and 0.2 mm dia. spheres. The values calculated for different gas velocities are adequately represented by a single line, whereas the experimental results depend slightly on the gas velocity. Nevertheless, the agreement is reasonable and one may conclude that the solids fractions at the bottom are much higher than the average values. The nonlinear variation of $\phi_{s,o}$ with $\bar{\phi}_s$ may explain the nonlinearity observed in Figure 1.

Introducing the solids fractions at the bottom in Eq. 1, 200 volumetric mass transfer coefficients measured in suspensions of single and mixed size glass spheres ($\phi_{s,o} \leqq 0.29$) are predicted with 10.3% mean error. The error is even less than in the case of

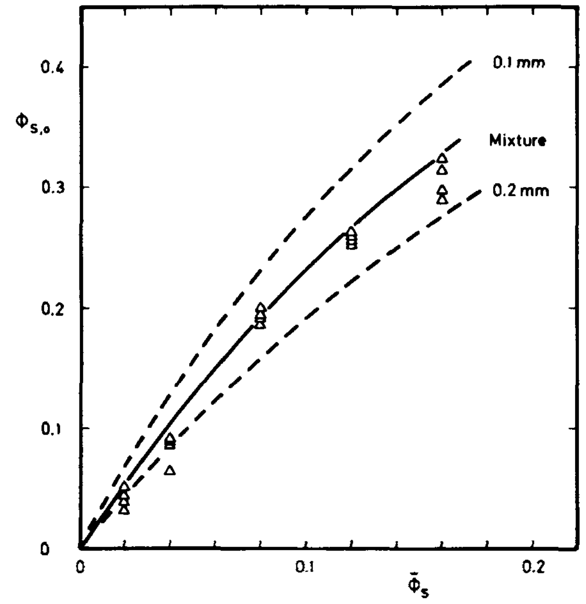

**Figure 3. Measured solids fractions at the bottom in suspensions of particle mixtures.**
△ 0.1 + 0.2 mm dia. sphere mix
---- —— predictions

the original data for three-phase fluidized beds to which Eq. 1 was fitted. A parity plot is given in Figure 4. Included are some data reported for the 0.05 mm dia. spheres in a column of 0.3 m dia. (Schumpe et al., 1984). For measurements at $\phi_{s,o}$ values that exceed the concentration range covered by Eq. 1, the predictions are systematically low. However, this might result from the smaller column diameter used in this study. The large bubbles formed in the concentrated suspensions are subjected to a stronger wall effect.

## Conclusions

The volumetric mass transfer coefficients in suspension bubble columns and three-phase fluidized beds can be described by the same relation, eq. 1, if the solids fractions at the bottom are considered. It is concluded that at moderate dispersion heights the suspension properties at the gas distributor determine the overall column performance.

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## Notation

$C_E$ = reading of oxygen electrode in arbitrary units
$C_E^*$ = equilibrium value of $C_E$ at $t \to \infty$
$D_s$ = solids dispersion coefficient, $m^2 \cdot s^{-1}$
$d_R$ = column diameter, m
$d_s$ = solids diameter, m
$Fr$ = Froude number $[=u_G/(gd_R)^{0.5}]$
$k_La$ = volumetric mass transfer coefficient with respect to dispersion $(G/L/S)$ volume, $s^{-1}$

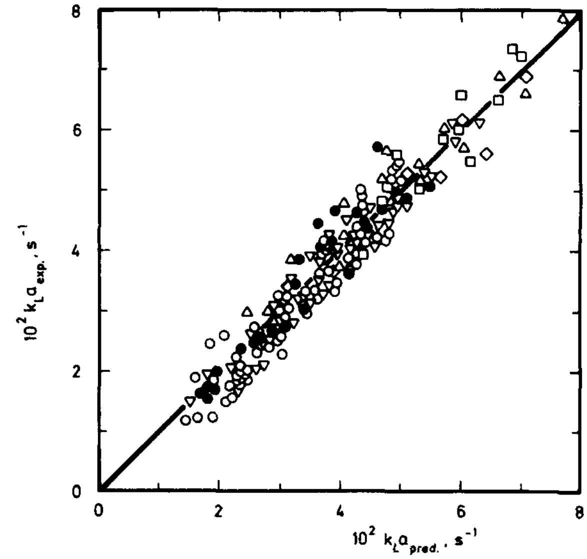

**Figure 4. Parity plot for Eq. 1 with $\phi_{s,o}$ in place of $\bar{\phi}_s$.**
◇ 0.05 + 0.2 mm dia. sphere mix
△ 0.1 + 0.2 mm dia. sphere mix
○ 0.05 mm dia. spheres
▽ 0.1 mm dia. spheres
□ 0.2 mm dia. spheres
● 0.05 mm dia. spheres at $d_R$ = 0.3 m (Schumpe et al., 1984)

$L$ = dispersion height, m
$Pe_s$ = Peclet number $(=u_{ss}L/D_s)$
$u_G$ = superficial gas velocity, $m \cdot s^{-1}$
$u_{ss}$ = hindered settling velocity, $m \cdot s^{-1}$
$u_T$ = terminal velocity of single particle, $m \cdot s^{-1}$
$t$ = time, s
$z$ = dimensionless axial coordinate

### *Greek letters*

$\epsilon_G$ = relative gas holdup
$\epsilon_L$ = relative liquid holdup
$\underline{\phi_s}$ = volume fraction of solids in gas-free suspension
$\bar{\phi}_s$ = axial average of $\phi_s$, determined from solids loading
$\phi_{s,o}$ = $\phi_s$ at the bottom ($z$ = 0)

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