Effect of Reactor Diameter on Gas-Liquid Mass Transfer in Three-Phase Fluidized Beds

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In a recent paper Nguyen-tien et al. (1985) reported on a comprehensive gas-liquid mass transfer study in fluidized particle beds. The measurements were carried out in a 0.14 m dia. reactor using a water-air-glass bead system. The oxygen liquid phase concentration profiles along the reactor (2.65 m long) were measured. Reliable volumetric liquid-side mass transfer coefficients (k_La) could be derived by employing an appropriate model that considers axial dispersion in the liquid phase and the nonisobaric conditions in the reactor due to the hydrostatic head. In the investigation of Nguyen-tien et al., a wide range of gas and liquid velocities, bed heights, and spherical particle sizes (0.05 to 8 mm) was covered. A new correlation for k_La in fluidized beds of particles of ≤ 1 mm dia. has been proposed.

In the study of Nguyen-tien et al. the reactor diameter was not varied. However, when scaling up multiphase reactors, the reactor diameter may have a significant influence on reactor performance (Shah and Deckwer, 1984). It is thought in particular that the ratio of column diameter to particle diameter D_C/d_S is of importance and may serve as a criterion in scale-up and design considerations. If this ratio is too small, the measured data such as volumetric mass transfer coefficients may not be representative for multiphase flow in larger equipment.

In order to confirm that the results and conclusions reported by Nguyen-tien et al. can also be applied to larger diameter reactors, it appeared desirable to carry out additional measurements in a larger column. If an effect of the reactor diameter or the D_C/d_S ratio, respectively, occurs at all, it will be particularly relevant for particles of larger size. Therefore, the additional runs were done in a reactor of 0.3 m dia. with glass spheres of 8 mm dia. Thus the D_C/d_S ratio was increased from 17.5 in the study of Nguyen-tien et al. to 37.5 in these measurements.

The O_2 absorption experiments were carried out in the same setup as described by Nguyen-tien et al. (1985) except that the absorption column with the fluidized bed and the regeneration column had a diameter of 0.3 m. In addition, the bottom of the absorption column was differently constructed, as shown in Figure 1. The 0.14 m dia. reactor of Nguyen-tien et al. was equipped with a conical bottom, the liquid entering at the center and the air being sparged through a ring distributor with 29 holes of 1 mm diameter. In the 0.3 m dia. column of this study a plane bottom was used. The gas was sparged by a perforated plate having 616 openings of 1 mm dia. 115 kg of 8 mm dia. glass spheres ($\rho_S = 2.51 \times 10^3$ kg/m^3) were employed, giving a height to the expanded three-phase fluidized bed (H_S) of about 1.3–1.4 m. The three-phase bed was followed by a two-phase zone of about 1.5 m height. In the 0.14 m dia. reactor a significant influence of the liquid flow rate on k_La was not observed for 8 mm particles. Therefore, only one liquid velocity of 0.095 m/s was employed. The mean gas velocity was varied from 0.016 to 0.14 m/s. The steady state oxygen concentration was measured along the reactor length at six positions in the fluidized particle bed and at four positions in the subsequent two-phase zone.

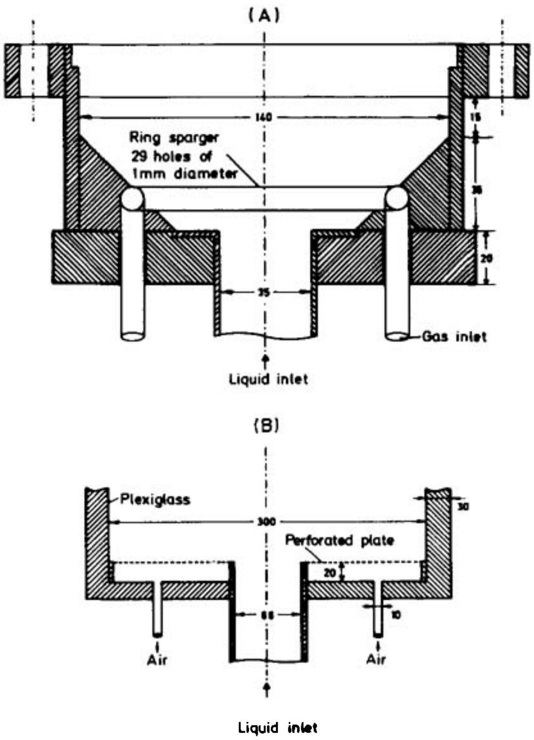

Figure 1. Gas sparger: (*A*) Nguyen-tien et al. (1984); (*B*) this study.

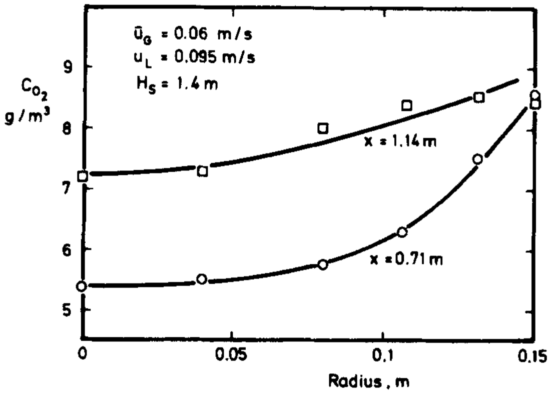

Figure 2. Radial profiles of oxygen in liquid phase.

When measuring the O_2 concentration, not only an axial profile but also a radial dependency was observed, particularly in the lower and middle parts of the fluidized particle bed. As an example, Figure 2 shows the radial profiles for a gas velocity of 0.06 m/s measured at 0.71 and 1.14 m above the gas and liquid inlets. Although radial profiles are more probable to occur in larger diameter reactors, it is evident that the radial profiles observed are a result of the phase distribution system used in these measurements. One could observe visually that glass spheres near the wall were not fluidized up to a height of about 0.3 m above the bottom plate. Thus a zone of conical shape with stagnant particles was formed above the gas sparger. The water entering through the central pipe could not fluidize particles in that region. It is understood that mass transfer rates are low in the conical region with stagnant particles as compared to the fluidized bed. Therefore, the actual length of the three-phase bed available for mass transfer was reduced.

For determination of the volumetric mass transfer coefficients k_La from the measured O_2 concentration data, two models were used, the back-flow cell model (BFCM) and the plug-flow model (PFM) (Nguyen-tien et al., 1985). As the radial profiles are approximately of parabolic shape, the concentration values measured at $r/R = 0.7$ were used. The height of the three-phase fluidized bed was reduced by 0.2 m to account for the formation of the stagnant particle zone above the sparger. Under this condition, a satisfactory description could be achieved when fitting the BFCM to the axial profiles.

The PFM as employed by Nguyen-tien et al. did not yield height-dependent k_La values even if a pointwise evaluation was used instead of fitting the whole profile. Therefore this model was used to calculate k_La from the inlet concentration and the concentration measured at the highest position in the bed. One can expect that in this case the effects caused by the particular phase distribution system used in the 0.3 m dia. reactor are less important and probably negligible.

The calculated mass transfer data are plotted vs. the gas velocity in Figure 3. The k_La values obtained from both methods do not differ much and are in satisfactory agreement with the data of Nguyen-tien et al. reported for the 0.14 m dia. column. Figure 3 also shows k_La values measured in the 0.3 m reactor for the two-phase system without particles (bubble column). These data are in reasonable agreement with the correlation proposed by Shah et al. (1982). Only at lower gas velocities are slightly increased k_La values found. This increase is due to homogeneous (bubbly) flow as the applied gas sparger generated uniformly small bubbles at low gas velocities. The correlation of Shah et al. (1982) considers mainly k_La data from the churn-turbulent flow regime.

In summary, it can be concluded that the volumetric mass transfer coefficients (k_La) measured in three-phase fluidized

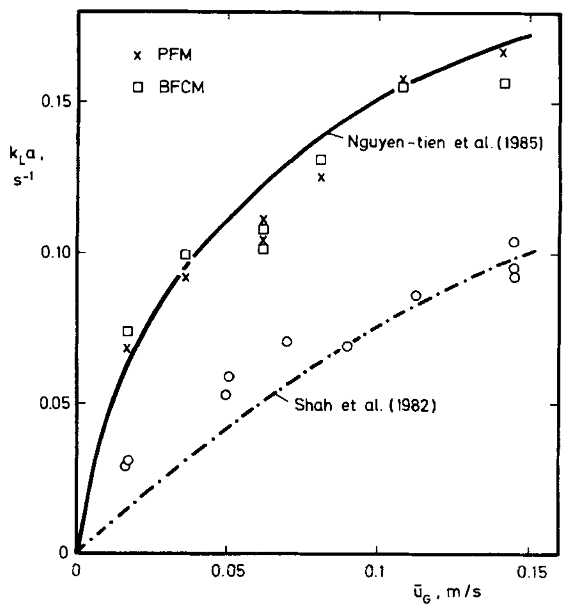

Figure 3. k_La data in 0.3 m dia. reactor. Comparison with Nguyen-tien et al. (1985) for three-phase bed and Shah et al. (1982) for bubble column.

particle beds in a 0.14 m dia. reactor can also be applied to larger diameter reactor. It is particularly thought that the k_La correlation developed by Nguyen-tien et al. for beds of small particles ($d_S \leqslant$ 1 mm) can be used successfully in larger scale equipment.

ACKNOWLEDGMENT

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NOTATION

D_C = column diameter, m
d_S = particle diameter, m
H_S = height of three-phase bed, m
k_La = volumetric mass transfer coefficient, s^{-1}
$\bar{u}_G$ = average linear gas velocity (volumetric gas flow/reactor cross section), m·s^{-1}
u_L = linear liquid velocity (volumetric liquid flow/reactor cross section), m·s^{-1}
x = axial coordinate, m

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