GAS-LIQUID MASS TRANSFER IN THREE-PHASE FLUIDIZED BEDS

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Introduction

A three-phase fluidized bed (TPFB) is formed when a bed of particles is fluidized by two immiscible fluids. The principal application of TPFB with cocurrent upward flow of gas and liquid appears to be gas absorption, fermentation¹⁾ and catalytic hydrodesulfurization^{9,10)}. There are few useful equations for gas-liquid mass transfer in TPFB.

For TPFB with cocurrent upward flow of gas and liquid, Østergaard and coworkers^{7,11)} have reported as follows: the liquid-phase volumetric coefficient of mass transfer, k_l^*a , for 6-mm glass spheres is about ten times larger than the coefficient for 1-mm spheres at low liquid-flow rates. In beds of 6-mm particles, bubble break-up occurs and the gaseous phase forms a uniform dispersion of small bubbles at low liquid-flow rates. Liquid mixing for beds of 6-mm spheres may be neglected except at high gas-flow rates, whereas beds of 1-mm spheres are characterized by a high degree of mixing.

Muroyama *et al.*⁸⁾ have presented a map for hydrodynamic regions of coalesced bubble, dispersed bubble and slug flow. In beds of 3- to 5-mm spheres, dispersed bubble flow was observed at low gas flow rates.

For beds of 6-mm glass spheres, 6.4 cm on the side in square section and 183 cm in height, Lee and Worthington⁶⁾ have reported data on gas holdup Φ_g , interfacial area a and k_l^*a every 30 cm above the gas distributor.

For upward gas flow through stagnant liquid and 1.1-cm resin spheres, Kito *et al.*⁵⁾ have presented data for Φ_g and a in 0.5 N NaOH solution.

For packed columns with cocurrent upward flow (PCCU), an early report⁴⁾ has presented the empirical equations for a in terms of gas holdup per void fraction Φ_g/ϵ , gas and liquid Reynolds numbers Re_g and Re_l , and a ratio of packing to column diameter d_p/d_T for respective flow regions: bubble (I), churn, pseudospray, bubble (II), pseudopulse and pulse. It was also observed that the empirical equation of k_l^* for packed columns with cocurrent downward flow

(PCCD), in terms of Re_g , Re_l and d_p/d_T given in an earlier report³⁾ is applicable to illustrate the data for respective flow regions in PCCU.

The purpose of this investigation was to obtain the equations of mean bubble diameter, d_b , and k_l^* , for TPFB from the analysis of the literature data using the empirical equations of gas-liquid mass-transfer for PCCD³⁾ and PCCU⁴⁾ given in previous reports.

Analysis of the Data and Discussion

1. Upward flow of gas and liquid

1) Interfacial area and mean bubble diameter

The values of Φ_g and a for beds of 6-mm glass spheres where $\epsilon=0.72$ were measured by Lee and Worthington⁶⁾ using the γ -ray attenuation and light transmission methods, respectively. These values were observed at 17.6, 47.6, 77.6 and 107.6 cm above a gas distributor with four nozzles. **Figure 1** shows the dependence of $ad_p\epsilon/\Phi_g$ on Re_g , where $Re_l=1106$. The thin solid line indicates the data rearranged by the least square method, except for the data at 107.6 cm.

The empirical equation of a for the bubble (I) flow region in PCCU is given in the early report⁴⁾ as follows:

$$ad_p/(\Phi_g/\epsilon)=16Re_l^{0.05}Re_g^{-0.4}(d_p/d_T)^0 \quad (1)$$

The thick solid line representing Eq. (1) is only 10% lower than the thin solid line. This suggests that the liquid forms a continuous phase and that the gas was in uniform bubble distribution in an axial direction of TPFB, similar to the behavior of the bubble (I) flow region in PCCU.

The mean bubble diameter is given by

$$d_b=6\Phi_g/a \quad (2)$$

Thus, Eq. (1) is rewritten as:

$$d_b/d_p=0.38\epsilon Re_l^{-0.05}Re_g^{0.4}(d_p/d_T)^0 \quad (3)$$

This equation supports the literature data⁷⁻¹⁰⁾ which indicate that small bubbles are obtained where d_p is large, and the values of ϵ and Re_g are small.

According to the map of hydrodynamic flow region given by Muroyama *et al.*⁸⁾, the Lee-Worthington data below 77.6 cm are located in the dispersed bubble flow region. However, as mentioned above, the flow pattern for dispersed bubble flow in TPFB shown by

Fig. 1 Correlation of Eq. (1) and the Lee-Worthington data, and rearrangement of the Kito data

Fig. 2 Correlation of Eq. (6) and the Lee-Worthington data

Fig. 3 Dependence of liquid-phase mass transfer coefficient on mean bubble diameter

Murayama *et al.* is similar to that for bubble flow (I) in PCCU.

In Fig. 1, the data for 107.6-cm bed height are very different from the values predicted by Eq. (1). This

supports the observation that the upper part of TPFB was a solid-free bed.

2) Liquid-phase mass-transfer coefficient

The values⁶⁾ of $k_l^* a$ were obtained from the data on absorption rate of carbon dioxide into water every 30 cm above the gas distributor. The equation of k_l^* for respective flow regions in PCCD and PCCU is simply given in the reports^{3,4)} as follows:

$$Sh_l = k_l^* d_p / D_l = 5 \times 10^2 \phi^{0.3} Re_l^{1/3} Re_g^{1/5} Sc_l^{1/2} (d_p/d_T)^{2.2} \quad (4)$$

Assuming that the exponent on (d_p/d_T) in Eq. (4) is 2.1 instead of 2.2, the equation for TPFB of spheres is obtained as

$$Sh_l = 8.9 \times 10^2 Re_l^{1/3} Re_g^{0.2} Sc_l^{1/2} (d_p/d_T)^{2.1} \quad (5)$$

Hence, the equation for $k_l^* a$ is given by Eqs. (1) and (5) as follows:

$$k_l^* a d_p^2 \epsilon / D_l \Phi_g = 1.4 \times 10^4 Re_l^{0.38} Re_g^{-0.2} Sc_l^{1/2} (d_p/d_T)^{2.1} \quad (6)$$

Lee and Worthington⁶⁾ have given the empirical equation for $k_l^* a$ as follows:

$$k_l^* a / \Phi_g = 1.53 \quad (7)$$

Figure 2 shows the correlation of Lee-Worthington-data⁶⁾ by Eq. (6). The thick solid line representing Eq. (6) agrees with the data within $\pm 30\%$ error. But the thin solid line representing Eq. (7) leads to 250% error.

The dependence of k_l^* on d_b is also expressed by combining Eqs. (2) and (6) as follows:

$$(k_l^* / d_b) (d_p^2 \epsilon / D_l) = 2.3 \times 10^3 Re_l^{0.38} Re_g^{-0.2} Sc_l^{1/2} (d_p/d_T)^{2.1} \quad (8)$$

For shallow-pool bubble-columns and bubble stirred tanks, Calderbank and Moo-Young²⁾ have shown that large bubbles ($d_b > 2.5$ mm) give larger k_l^* value than small bubbles ($d_b < 2.5$ mm). The k_l^* value increases as d_b increases in the transition region in which d_b is about 2.5 mm. The dependence of k_l^* value on d_b for TPFB is roughly in agreement with that observed in the transition region by Calderbank and Moo-Young²⁾, but the effect of Re_g and Re_l on k_l^* / d_b cannot be neglected, as is seen from Eq. (8) or Fig. 3.

2. Upward flow of gas through stagnant liquid

The free-expansion bed in which gas flows upward through stagnant liquid is capable of operating in a wide range of Re_g . For the gas-0.5 N NaOH solution system, the data on Φ_g and a are given by Kito *et al.*⁵⁾ for beds of 1.1-cm resin particles, 10 cm in bed diameter, with a sieve plate used as gas distributor. The values of $ad_p\varepsilon/\Phi_g$ are independent of Re_g and comparable to those in continuous liquid flow, as shown in Fig. 1. Thus, the gas-liquid behavior is analogous to that in the transition state between bubble (I) and pseudospray flow in PCCU.

Conclusions

Gas-liquid mass transfer characteristics for three-phase fluidized beds of large spheres are obtained as follows.

1) The mean bubble diameter is predicted by an empirical equation of interfacial area for the bubble (I) flow region in packed columns with cocurrent upward flow.

2) The liquid-phase mass-transfer coefficient is also evaluated from a modification of the empirical equation for packed columns with cocurrent flow given previously.

Nomenclature

a	= interfacial area per unit volume of reactor	[cm ⁻¹]
D	= diffusivity	[cm ² /sec]
d_b	= mean bubble diameter	[cm]
d_p	= diameter of solid particles or packings	[cm]
d_r	= diameter or side of reactor	[cm]
k^*	= mass transfer coefficient	[cm/sec]
Re	= Reynolds number for continuous medium, $d_p u \rho / \mu$	[—]

S	= surface of particles or packings	[cm ²]
Sc	= Schmidt number, $\mu/\rho D$	[—]
Sh	= Sherwood number, $k^* d_p / D$	[—]
u	= linear velocity for continuous medium	[cm/sec]
V	= reactor volume	[cm ³]
ε	= void fraction, $(V_g + V_i)/(V_g + V_l + V_s)$	[—]
μ	= viscosity	[g/cm·sec]
ρ	= density	[g/cm ³]
Φ_g	= gas holdup, $V_g/(V_g + V_l + V_s)$	[—]
ϕ	= surface shape factor of solid particles or packings, S/d_p^2	[—]

<Subscripts>

g	= gas phase
l	= liquid phase

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