Mass Transfer in Continuous Bubble Columns with Floating Bubble Breakers

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Bubble column reactors are widely used in the fields of petrochemical, biochemical, and chemical engineering processes. They are comparable to other multiphase reactors because of their simplicity, low operating costs, and higher heat and mass transfer coefficients between phases. Since a continuous bubble column reactor consists of a dispersed gas phase in the continuous liquid phase, the bubble size and its phase holdup are needed to estimate the phase contacting capability between the phases, which plays a vital role in the effective performance of such reactors.

The volumetric mass transfer coefficients in batch and continuous bubble columns have been determined by several investigators (Akita and Yoshida, 1973; Deckwer et al., 1974, 1982, 1983; Godbole et al., 1984). It has been found that the axial dispersion model can be applied to analyze the volumetric mass transfer coefficient from knowledge of the concentration profile along the entire height of the column. In general, the volumetric mass transfer coefficient in the bubble columns increases with increasing gas velocity and decreases with an increase in the liquid viscosity, whereas the liquid velocity has a minor effect on the volumetric mass transfer coefficient. However, Tang and Fan (1990) found a significant effect of liquid velocity on the liquid-side mass transfer coefficient in a three-phase fluidized bed containing low-density particles.

Recently, Kim and Kim (1990) have obtained a much higher volumetric mass transfer coefficient in three-phase beds with floating bubble breakers. One of the effective ways to increase the volumetric mass transfer coefficient in the bubble columns is to decrease the bubble size and consequently increase the bubble phase holdup by means of floating bubble breakers in the column.

In the present study, the gas holdup and gas–liquid mass transfer coefficient in bubble columns with floating bubble breakers have been measured. The effects of bubble breakers on the gas holdup and volumetric mass transfer coefficient in continuous bubble columns with non-Newtonian liquids (carboxyl methyl cellulose solutions) have been determined.

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Experimental Method

Experiments were conducted in a 2 m long acrylic column of 0.15 m ID. Details of the experimental apparatus can be found elsewhere (Kang et al., 1985). The gas and liquid phase holdups were determined by Eqs. 1 and 2 from knowledge of the pressure drop through the column and the amount of the bubble breaker, which can be calculated from the known weight and density of the breaker.

$$\epsilon_g + \epsilon_l + \epsilon_F = 1.0 \tag{1}$$

$$\frac{\Delta P}{L} = (\epsilon_g \rho_g + \epsilon_l \rho_l + \epsilon_F \rho_F) g \tag{2}$$

Dissolved oxygen concentration in the liquid samples taken at five axial locations along the column (0, 0.11, 0.36, 0.61, and 0.81 in dimensionless axial coordinates) were measured by a dissolved oxygen meter by way of solenoid valves. The liquid sampling was repeated at three radial positions (0, 0.5, and 1.0 in dimensionless radial coordinate, r/R) at each sampling tap located axially at the wall of the column. In measuring the dissolved oxygen concentration, the sampled liquids were stirred by a magnetic stirrer and were maintained at a constant temperature. Recycled liquid from a weir at the top of the column was fed to a purge column where the dissolved oxygen in the liquid was desorbed by pure compressed nitrogen gas. The temperature was maintained at 20 ± 0.5°C, and the dissolved oxygen concentration was maintained below 1.0×10^{-4} mol/L in the purge column.

Air and aqueous solutions of carboxy methyl cellulose (CMC) were used for the gas and liquid phases, respectively. A floating bubble breaker was made of a cylindrical acrylic tube (0.019 m long, 0.015 m OD) in which a copper rod was inserted to adjust the density of the floating bubble breaker to 1,300 kg/m^3.

Liquid viscosity

The effective viscosity of a non-Newtonian liquid ($0.72 \leq n \leq 1$) can be written in the form of the Ostwald-de Waele relation (Kawase and Moo-Young, 1986; Schumpe et al., 1989) as

$$\mu_{eff} = \kappa \dot{\gamma}_{eff}^{n-1} \tag{3}$$

Since CMC solutions have a pseudoplastic flow behavior, the effective shear rate due to the rise of bubbles with velocity relative to the liquid flow can be described as in Eq. 4 in upward concurrent operation (Schumpe et al., 1989).

$$\dot{\gamma}_{eff} = 2{,}800\left(U_g - U_l \frac{\epsilon_g}{\epsilon_l}\right) \tag{4}$$

In this equation, the units of the shear rate and fluid velocity are 1/s and m/s, respectively. Therefore, the effective shear rate and viscosity can be obtained from Eqs. 3 and 4 with the predetermined fluid consistency index, K, and flow behavior index, n. The values of K and n of CMC solutions used in this study are listed in Table 1.

Volumetric mass transfer coefficient

The volumetric mass transfer coefficients in the bubble columns were determined by employing the axial dispersion model (Deckwer et al., 1974, 1983), which is based on the oxygen balance in the liquid phase. The final form of this model can be written as Eq. 5

$$\frac{1}{Pe}\frac{d^2c}{dx^2} - \frac{dc}{dx} + St(c^* - c) = 0 \tag{5}$$

with the following boundary conditions

$$\text{at } x = 0, c = c_0 + \frac{1}{Pe}\frac{dc}{dx}\bigg|_{x=0} \tag{6}$$

$$\text{at } x = 1, \frac{dc}{dx}\bigg|_{x=0} \tag{7}$$

where

$$Pe = \frac{U_l L}{D_Z \epsilon_l}, S_t = (k_L a)\frac{L}{U_l}, x = Z/L \tag{8}$$

$$C^* = a + bx \tag{9}$$

Table 1. Liquid Physical Properties and Fluids Flow Rate Range

Solution	ρ_l kg/m^3	$K \times 10^3$ Pa · s^n	n	$\sigma_l \times 10^3$ N/m	$U_g \times 10^2$ m/s	$U_l \times 10^2$ m/s
Water	1,000	1.000	1.000	72.9	2.16–9.5	3.5–10.5
CMC(1)	1,001	21.69	0.882	73.2	2.16–9.5	3.5–10.5
CMC(2)	1,002	43.82	0.847	73.3	2.16–9.5	3.5–10.5
CMC(3)	1,003	71.64	0.825	73.6	2.16–9.5	3.5–10.5

In Eq. 9, C^* is a equilibrium concentration of oxygen, and a and b can be written as Eqs. 10 and 11, respectively:

$$a = \frac{Y}{H}(P_T + \rho_l g \epsilon_l L) \tag{10}$$

$$b = -\frac{Y}{H}\rho_l g \epsilon_l L \tag{11}$$

Results and Discussion

It has been generally understood that the larger particles can have a bubble breaking potential in three-phase fluidized beds, which prompted the concept of a bubble breaker in bubble columns to increase the contact efficiency of gas bubbles with a liquid medium.

The size of a bubble and its rising velocity increase as the bubble rises in the column, resulting from bubble coalescence and growth. It can be effective, therefore, to break the rising bubbles above a certain level of the column, since the amount of the bubble breaker is small so as not to increase the pressure drop as well as not to alter the structure of the bubble column. The size of the bubble breaker must be large enough to break the enlarged rising bubble, while its density can be adjusted to be relatively low, in order to float above a certain level of the column.

Gas-phase holdup

The effects of floating bubble breakers on the gas holdup are shown in Figure 1. The gas holdups increase with an increase in the volume fraction of bubble breakers in the column, especially at relatively higher gas velocities. The increment of the gas holdup with the addition of bubble breakers has been up to 13% in a column with viscous CMC solutions. The gas holdup in the present continuous bubble column with floating bubble breakers has been correlated with the experimental variables as

$$\epsilon_g = 0.331 U_g^{0.663} \mu_{eff}^{-0.189} (1 + \epsilon_F)^{1.872} \tag{12}$$

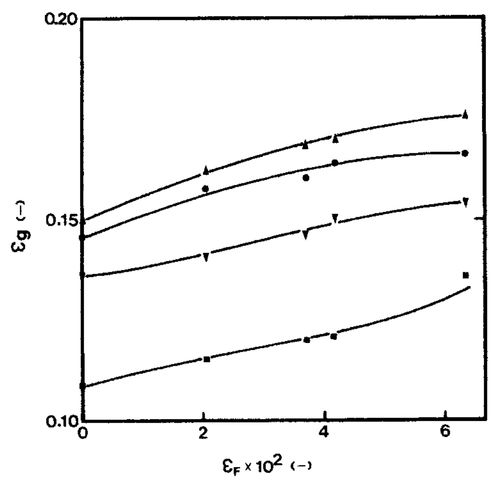

Figure 1. Effects of ϵ_F on ϵ_g in viscous bubble columns with floating bubble breakers.

	●	▲	■	▼
U_l m/s	0.035	0.056	0.056	0.105
U_g m/s	0.076	0.095	0.058	0.095
Liquid	CMC(1)	CMC(1)	CMC(2)	CMC(3)

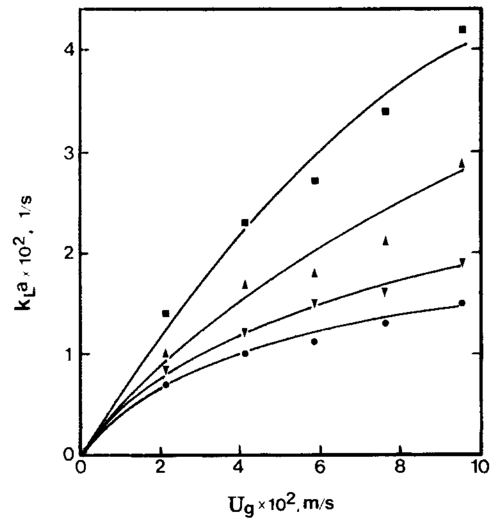

Figure 2. Effects of U_g on k_La in viscous bubble columns with floating bubble breakers.

	●	▲	■	▼
U_l m/s	0.105	0.056	0.105	0.105
ϵ_F	0	0.021	0.042	0.063
Liquid	CMC(3)	CMC(2)	CMC(1)	CMC(3)

with a correlation coefficient of 0.941. In Eq. 12, the units of U_g and μ_{eff} are m/s and Pa · s, repectively.

Volumetric mass transfer coefficient

The volumetric mass transfer coefficient has been obtained by fitting the analytical solution of Eq. 5 to the dissolved oxygen concentration profile in the axial direction of the column (Deckwer et al., 1974). A mean value of the oxygen concentration of three sampled liquids taken from different radial positions at a given axial location has been considered as the oxygen concentration at that axial coordinate. For the determination of two parameters, *St* and *Pe*, included in Eq. 5, Powell's algorithm, which is efficient for finding the minimum of a function of several variables, has been employed (Powell, 1964; Kim and Kim, 1990; Tang and Fan, 1990).

Thus, the obtained volumetric mass transfer coefficient, k_La, increases with an increase in the gas velocity in viscous bubble columns with bubble breakers, as shown in Figure 2. It has been found that the addition of bubble breakers can reduce the bypassing of large bubbles, especially at the higher gas velocities, not only by breaking the bubbles but also by creating small-scale eddies. The gas–liquid mass transfer rate has been reported to be promoted strongly by small-scale eddies, which can be generated at the expense of energy dissipation in the column (Lamont and Scott, 1970). The axial dispersion coefficients of viscous liquid media obtained from the parametric fitting were somewhat lower than those in the column without bubble breakers (Deckwer et al., 1974; Kelkar and Shah, 1985; Kang et al., 1987).

The effects of liquid viscosity on k_La in bubble columns with bubble breakers can be seen in Figure 3. The decrease in k_La is mainly due to the decrease in the gas holdup with increasing liquid viscosity. It can be noted from Figure 3 that the benefit in mass transfer coefficient that arises from the addition of bubble breakers in the column, reduces with an increase in the liquid viscosity. This effect may occur because the mobility of floating breakers becomes restricted in the viscous liquid medium.

As can be seen in Figure 4, the volumetric mass transfer coefficients increase with an increase in the volume fraction of bubble breakers in all the cases studied. The values of k_La obtained in the bubble columns without bubble breakers have agreed well with those calculated from the correlations of Deckwer et al. (1982) and Godbole et al. (1984) within 30% error bounds (Figure 3). The increment of the k_La has been up to 25% with an increase in the volume fraction of bubble breakers up to 0.063 in the column. This increase of the k_La is, of course, mainly attributed to the frequent breakage of bubbles in the column as the number of bubble breakers increases. The increase of the volume fraction of bubble breakers can, in addition, cause the increase in the turbulence intensity generated by the bubble breakers, since the floating breakers disturb the steady upward flow of the gas–liquid mixture in the column (Kang and Kim, 1986).

However, it should be considered that the higher volume fraction of bubble breakers can result in a decrease in the gas–liquid contact area in the column, similar to that in the case of suspended solid particles (Shah et al., 1982). It has been reported that the k_La values show their maxima with an increase in the volume fraction of bubble breakers in three-phase fluidized beds, where the holdups of solid particles are in the range of 0.3–0.6 (Kim and Kim, 1990).

The relation between bubble size and the amount of bubble breakers can be found from the experimental results of Kim and Kim (1987) in three-phase fluidized beds. The bubble length decreases with an increase in the amount of bubble breakers; however, it increases again with further increase in the amount of bubble breakers, exhibiting a minimum value. It can be

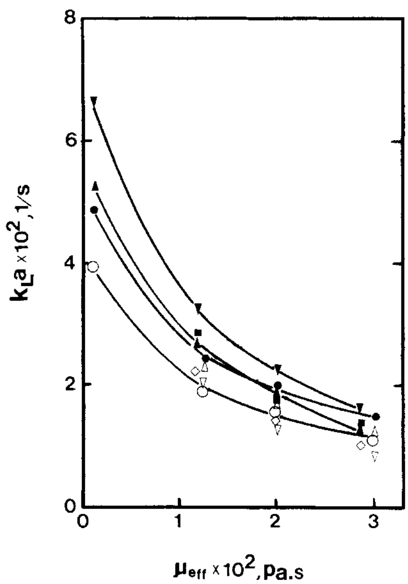

Figure 3. Effects of μ_{eff} on k_La in bubble columns with floating bubble breakers.

	●*	○*	▼*	▲*	△**	◇**	■†	▽†
U_l m/s	0.056	0.056	0.105	0.105	—	—	—	—
U_g m/s	0.058	0.058	0.076	0.076	0.058	0.076	0.058	0.076
ϵ_F	0.042	0	0.063	0	0	0	0	0

*This study; **Deckwer et al. (1982); †Godbole et al. (1984)

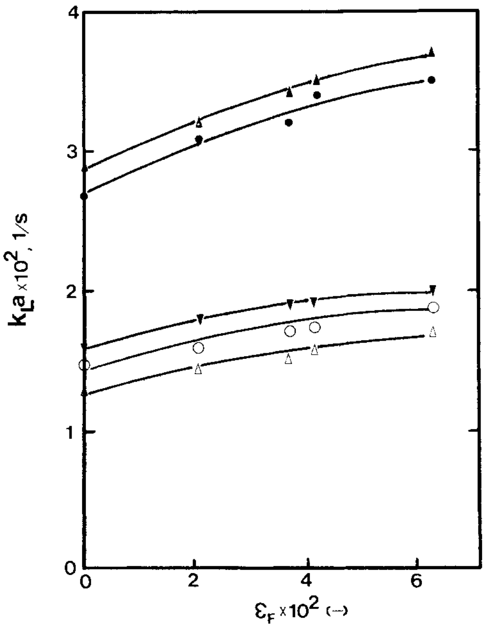

Figure 4. Effects of ϵ_F on k_La in viscous bubble columns with floating bubble breakers.

	●	▲	■	▼	▼
U_l, m/s	0.105	0.035	0.056	0.105	0.056
U_g, m/s	0.076	0.095	0.058	0.095	0.095
Liquid	CMC(1)	CMC(3)	CMC(2)	CMC(3)	CMC(1)

understood that the gas holdup and volumetric mass transfer coefficient increase with decreasing bubble size. From Tang and Fan (1990), the k_La value decreases with an increase in the solid holdup from 0.047 to 0.137 in a three-phase bed of polystyrene particles.

It can be inferred, therefore, that the upper limit of the volume fraction of bubble breakers may exist for the maximum k_La and lower pressure drop. The values of k_La in the continuous bubble columns with floating bubble breakers have been correlated in terms of the experimental variables as in Eq. 13 with a correlation coefficient of 0.928. In Eq. 13, the units of the k_La, fluid velocity, and liquid viscosity are 1/s, m/s, and Pa.s, respectively.

$$k_La = 4.47 \times 10^{-2} U_g^{0.782} U_l^{0.160} \mu_{eff}^{-0.407} (1 + \epsilon_F)^{3.64} \tag{13}$$

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Notation

a = constant, Eq. 9
b = constant, Eq. 9
C = oxygen concentration, mol/L
C_o = initial oxygen concentration, mol/L
C^* = equilibrium concentration, mol/L
D = column diameter, m
g = gravitational acceleration, m/s^2
H = Henry's constant, atm·L/mol
K = fluid consistency index, Pa·s^n
k_La = volumetric mass transfer coefficient, 1/s
L = column height, m
n = flow behavior index
Pe = Peclet number, Eq. 8
ΔP = pressure drop in column, Pa
P_T = pressure at top of column, Pa
S_t = Stanton number, Eq. 8
U = superficial velocity, m/s
x = dimensionless distance, Eq. 8
z = axial distance, m

Greek letters

ϵ = holdup
ρ = density, kg/m^3
μ = viscosity, Pa·s
γ = shear rate, 1/s

Subscripts

eff = effective
F = floating bubble breaker
g = gas
l = liquid

Literature Cited

Akita, K., and F. Yoshida, "Gas Holdup and Volumetric Mass Transfer Coefficient in Bubble Columns," *Ind. Eng. Chem. Process Des. Dev.*, **12**, 76 (1973).

Chang, S. K., Y. Kang, and S. D. Kim, "Mass Transfer Characteristics of Three-Phase Fluidized Beds," *J. Chem. Eng. Japan*, **19**, 524 (1986).

Deckwer, W.-D., R. Burckhart, and G. Zoll, "Mixing and Mass Transfer in Tall Bubble Columns," *Chem. Eng. Sci.*, **29**, 2177 (1974).

Deckwer, W.-D., K. Nguyen-Tien, A. Schumpe, and Y. Serpemen, "Oxygen Mass Transfer into Aerated CMC Solutions in a Bubble Column," *Biotech. Bioeng.*, **24**, 461 (1982).

Deckwer, W.-D., K. Nguyen-Tien, G. B. Kelkar, and Y. T. Shah, "Applicability of Axial Dispersion Model to Analyze Mass Transfer Measurements in Bubble Columns," *AIChE J.*, **29**, 915 (1983).

Godbole, S. P., A. Schumpe, Y. T. Shah, and N. L. Carr, "Hydrodynamics and Mass Transfer in Non-Newtonian Solutions in a Bubble Column," *AIChE J.*, **30**, 213 (1984).

Kang, Y., and S. D. Kim, "Radial Dispersion Characteristics of Two- and Three-Phase Fluidized Beds," *Ind. Eng. Chem. Process Des. Dev.*, **25**, 717 (1986).

Kang, Y., I. S. Suh, and S. D. Kim, "Heat Transfer Characteristics of Three-Phase Fluidized Beds," *Chem. Eng. Commun.*, **34**, 1 (1985).

Kang, Y., W. M. Lim, and S. D. Kim, "Axial and Radial Mixing Characteristics in Bubble Columns," *Hwahak Konghak*, **25**, 460 (1987).

Kawase, Y., and M. Moo-Young, "Influence of Non-Newtonian Flow Behavior on Mass Transfer in Bubble Columns with and without Draft Tubes," *Chem. Eng. Commun.*, **40**, 67 (1986).

Kelkar, B. G., and Y. T. Shah, "Gas Holdup and Backmixing in Bubble Column with Polymer Solutions," *AIChE J.*, **31**, 700 (1985).

Kim, J. O., and S. D. Kim, "Bubble Characteristics in Three-Phase Fluidized Beds of Floating Bubble Breaker," *Particulate Sci. Technol.*, **5**, 309 (1987).

———"Mass Transfer Characteristics in Three-Phase Fluidized Beds with Floating Bubble Breakers," *Can. J. Chem. Eng.*, (1990).

Lamont, J. C., and D. S. Scott, "An Eddy Cell Model of Mass Transfer into the Surface of a Turbulent Fluid," *AIChE J.*, **16**, 513 (1970).

Powell, M.J.D., "An Efficient Method for Finding the Minimum of a Function of Several Variables without Calculating Derivatives," *Computer J.*, **7**, 155 (1964).

Schumpe, A., W.-D. Deckwer, and K. D. P. Nigam, "Gas–Liquid Mass Transfer in Three-Phase Fluidized Beds with Viscous Pseudoplastic Liquids," *Can. J. Chem. Eng.*, **67**, 873 (1989).

Shah, Y. T., B. G. Kelkar, S. P. Godbole, and W.-D. Deckwer, "Design Parameters Estimations for Bubble Column Reactors," *AIChE J.*, **28**, 353 (1982).

Tang, W. T., and L. S. Fan, "Gas–Liquid Mass Transfer in a Three-Phase Fluidized Bed Containing Low-Density Particles," *Ind. Eng. Chem. Res.*, **29**, 128 (1990).

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