Bubble Characteristics in the Radial Direction of Three-Phase Fluidized Beds

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Introduction

Bubble characteristics such as bubble size and rising velocity in three-phase fluidized beds provide the basic information for analyzing phase holdup, axial and radial mixing, and heat and mass transfer characteristics. The relationship between the bubble size and its rising velocity (Massimilla et al., 1961; de Lasa et al., 1984; Lee et al., 1984; Matsuura and Fan, 1984) and the bubble size distribution (Kim et al., 1977; Page and Harrison, 1972; Ha and Kim, 1980; Matsuura and Fan, 1984) in three-phase fluidized beds have been investigated. On the other hand, Kim and Kim (1987) observed bubble size reduction by addition of floating bubble breakers.

Rigby et al. (1970) observed that the radial distribution of bubble holdup and frequency has its maximum at the center of a column under the condition of fully developed flow. By contrast, Morooka et al. (1982, 1986) observed that the local gas holdup decreased linearly with the square of the dimensionless radial distance as in bubble columns (Ueyama and Miyauchi, 1979; Joshi and Sharma, 1979). Lee and de Lasa (1987) adopted a polynomial equation to predict the local gas holdup and bubble frequency in a bed of 250 μm glass beads. A U-shaped optical fiber probe has been developed by de Lasa et al. (1984) to measure bubble characteristics in three-phase fluidized beds. The optical fiber probes are convenient for controlling the light transmission in the bed (Lee and de Lasa, 1984, 1987). Information on radial nonuniformity of bubble rising velocity, bubble size and local gas holdup in three-phase fluidized beds is very sparse.

Therefore, in this study, the bubble holdup, mean velocity, and mean bubble chord length in a three-phase fluidized bed of 0.254 m-ID with four different particle sizes (0.4–6.0 mm) have been determined in the radial direction by means of a U-shaped optical fiber probe.

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Experimental

Experiments were carried out in a Plexiglass column of 0.254 m in diameter and 2.5 m high. The liquid and gas distributor is identical in arrangement to that in a previous publication (Kang and Kim, 1986). The particle sizes and operating conditions are summarized in Table 1. About 30 kg of glass beads with a density of 2,500 kg/m^3 were fluidized by water in the liquid phase and by air in the gas phase.

The bubble chord length and velocity were measured by means of a two-channel optical fiber probe made of plastic fiber 250 μm in diameter and located at 0.325 m above the distributor, approximately equal to 0.4–0.7 times expanded bed heights. The active elements of the probe were made by bending the fiber into U-shape and encasing the entire fiber except the U-shaped bend in a stainless steel tube of 6 mm in diameter (Jones and Delhaye, 1976; de Lasa et al., 1984). An He-Ne laser unit (Japan laser, JLHRT2OU) was used as the light source and two photo-transistors (ST-KLBII) were used as light intensity for electrical voltage converters. The sampling time intervals of the signals were 50 μs between the first and second tips, and 500 μm at each of the tips individually with a reduction of time intervals until they gave no further effects on the results. The signals were observed by an oscilloscope. The high-level signals corresponding to the gas phase are 2.25 V and the lower-level signals corresponding to the liquid phase are near zero. Total sampling time was 90 s, which is equivalent to about 50–500 bubbles at each operating condition.

The distance between the two tips was 2 mm. Therefore, bubbles with a length below 2 mm are eliminated because very small bubbles may strike each of the two tips, having some time gap. Moreover, the bubble with a length above 2 mm must have the gas phase at both tips and leave from the lower tip before passing through the upper tip. The local gas holdup, mean bubble chord length, and mean rising velocity were calculated using the relations proposed by Park et al. (1969), and Lee and de Lasa (1987).

Table 1. Operating Conditions and the Particles Used in This Study

d_p(mm)	Density (kg/m^3)	U_l(m/s)	U_g(m/s)
0.4	2,500	0.006–0.018	0.01–0.06
1.0	2,500	0.020–0.060	0.01–0.10
2.3	2,500	0.040–0.010	0.01–0.14
6.0	2,500	0.060–0.012	0.01–0.14

The local mean bubble velocity and bubble chord length are, respectively, the arithmetic mean and 50% of the cumulative value of a log-normal distribution. To estimate size and velocity of descending bubbles, any bubble with the maximum circulation velocity above 1.0 m/s was also eliminated, as calculated from the circulation cell model (Joshi and Sharma, 1979) with bubble velocity at the center of column.

Results and Discussion

Gas-phase holdup

Radial distribution of local gas-phase holdup in three-phase fluidized beds is shown in Figure 1. Gas-phase holdup decreases with an increase in radial distance. Larger bubbles are considered to be formed due to the bubble coalescence in the bed of 2.3 mm glass beads (Kim et al., 1972, 1975) and tend to rise along the center of the bed due to the wall effect, while being accompanied by many small bubbles. In the present study, the radial profile of gas-phase holdup is flatter than that of Morooka et al. (1982), since the wall effect might be less in the present larger column size, Figure 1.

The gas-phase holdup profile has been approximated by the following relation in previous studies (Miyauchi and Shyu, 1970; Ueyama and Miyauchi, 1979; Morooka et al., 1982) in both bubble columns and three-phase fluidized beds.

$$\epsilon_g/\bar{\epsilon}_g = \frac{n+2}{n}[1-(r/R)^n] \qquad (1)$$

In this study, the average gas-phase holdup is represented as:

$$\bar{\epsilon}_g = 3.697 d_p^{0.309} U_l^{-0.022} U_g^{0.701} \qquad (2)$$

The n value in Eq. 1 has been correlated with particle size in the bubble coalescing regime (0.4, 1.0, 2.3 mm glass beads) since the radial gas holdup profiles were found to be nearly flat in the bubble disintegrating bed (6.0 mm glass beads):

$$n = 409 d_p^{0.653} \qquad (3)$$

Miyauchi and Shyu (1970) reported the value of n as 8, but other studies (Ueyama and Miyauchi, 1979; Morooka et al., 1982) have represented its value as 2. The values of n in Eq. 3 ranged from 2.47 to 7.74; depending on particle size, the radial profile of gas-phase holdup may be flattened with an increase in particle size. The local gas-phase holdups of the present system in the bubble coalescing regime can be determined from Eqs. 1, 2 and 3 with a regression coefficient of 0.95.

The radial distribution of gas-phase holdup in the bed of 6 mm glass beads is shown in Figure 2. The radial profile of gas phase holdup is found to be much flatter than that in the bed of 2.3 mm glass beads. Bubbles formed by the distributor may be disintegrated by their axial mixing with larger particles and are dispersed uniformly throughout the bed of 6 mm particles as observed by Kim et al. (1975).

The radial profile of gas-phase holdup in the bed of 6 mm glass beads is far less affected by gas velocity than that in the bed of 2.3 mm glass beads. However, the radial gas holdup profile starts to develop at a higher gas velocity in the former, since bubbles are coalesced.

Bubble chord length distribution

In the present study, bubble chord length distribution is found to be log-normal, Figure 3a, as in the previous studies (Darton and Harrison, 1974; Kim et al., 1977; Ha and Kim, 1980; Matsuura and Fan, 1984). As shown in the figure, bubble chord

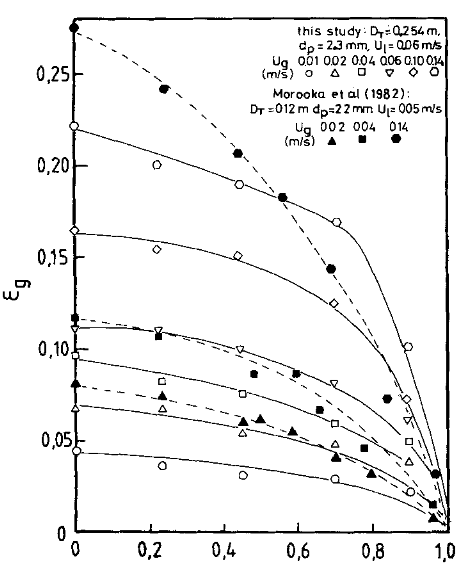

Figure 1. Radial distribution of gas-phase holdup in the three-phase fluidized beds.

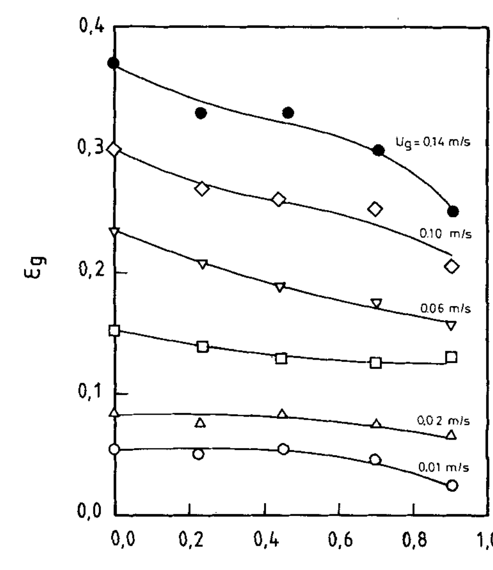

Figure 2. Radial distribution of gas-phase holdup in the bed of 6.0 mm glass beads at U_l = 0.10 m/s.

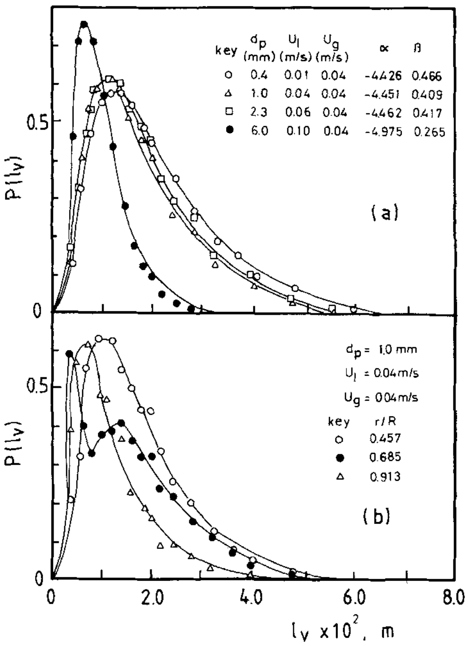

Figure 3. Bubble chord length distribution.
a) For various particle size
b) For various radial distance in the bed of 1 mm glass beads

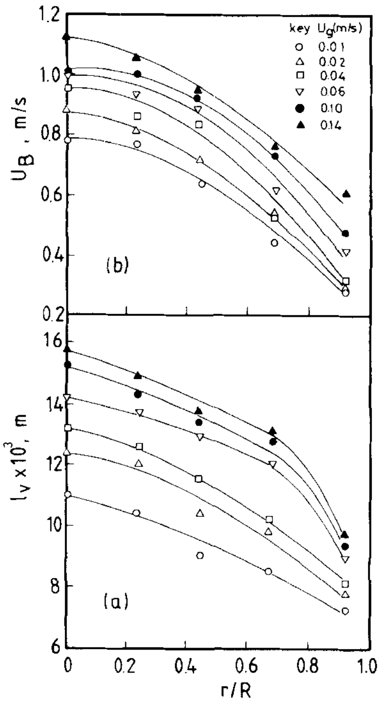

Figure 4. Radial distribution of vertical bubble length and bubble velocity in the bed of 2.3 mm glass beads, at $U_l = 0.06$ m/s.

length distribution is wider in the bubble coalescing regime ($d_p = 0.4, 1.0, 2.3$ mm) than in the bubble disintegrating regime ($d_p = 6.0$ mm). In the bubble coalescing regime, the bubble size distribution becomes narrower with an increase in particle size.

The bubble size distribution is found to be bimodal at the dimensionless distance r/R of 0.685 in the bed of 1 mm glass beads, Figure 3b. The first peak represents the small bubbles descending along the recirculating flow and the second peak represents the ascending large bubbles which overcome the down flow near the wall. However, bubble chord length distribution becomes a log-normal distribution at the dimensionless distance r/R of 0.915, since the population of large bubbles decreases because of the wall effect. This phenomenon can be observed in the bubble coalescing regime ($d_p = 0.4, 1.0, 2.3$ mm).

Local bubble chord length and bubble velocity

The radial distributions of bubble chord length and bubble velocity in the bed of 2.3 mm glass beads are shown in Figure 4. As in the case of gas-phase holdup, bubble size decreases with an increase in radial distance. Consequently, since coalesced large bubbles tend to rise along the center of the column due to the wall effect, their velocities decrease with an increase in radial distance, Figure 4b. Since liquid recirculation flow exists in a fluidized bed (Morooka et al., 1982), it can be anticipated that the bubble velocity in the peripheral region would be far lower than that in the center region of the column due to the inhibition effect of liquid down flow on small bubbles. Moreover, it can be observed visually that some bubbles descend with the liquid near the wall of the column.

The radial distributions of the bubble chord length and of the bubble velocity in the bed of 6 mm glass beads are flatter than those in the gas phase holdup. The bubble chord length and its rising velocity do not decrease significantly along the radial direction; since the bed of 6 mm glass beads is known as the bubble disintegrating bed (Kim et al., 1975), the bubbles formed from the distributor are disintegrated into a comparatively uniform size in the bed. Therefore, liquid recirculating flow may not exist at any gas velocity up to 0.06 m/s. Above this velocity, liquid recirculating flow may exist since bubble velocity decreases significantly with an increase in radial distance.

Present data for mean bubble chord length can be correlated by the following equation which was proposed by Koide et al. (1979) from the radial profile of gas-phase holdup in a bubble column.

$$\frac{l_{vc} - l_v}{l_{vc}} = \frac{l_{vc} - l_{v_w}}{l_{vc}} \left(\frac{r}{R}\right)^m \tag{4}$$

where l_{vc} is the bubble chord length at the center of the bed and l_{vw} is the bubble chord length estimated by extrapolation of the value from r to R. The values of l_{vc}, l_{vw} and m can be correlated by the following equations.

$$l_{vc} = 2.667 \times 10^{-3} d_p^{-0.300} U_l^{-0.072} U_g^{0.221} \tag{5}$$

$$l_{vw} = 4.295 \times 10^{-3} d_p^{-0.129} U_l^{-0.060} U_g^{0.124} \tag{6}$$

$$m = 3.770 d_p^{0.121} \tag{7}$$

Using Eqs. 4 to 7, local bubble chord length can be estimated with a regression coefficient of 0.86. As illustrated by Eq. 6, the gas phase velocity does not greatly affect the vertical bubble length estimated at the column wall.

In addition, the radial profile of bubble rising velocity can be predicted in the bubble coalescing regime by the same form of

Eq. 1 with a regression coefficient of 0.86, since the rising velocity is an inverse function of the gas-phase holdup.

$$U_B = \overline{U}_B \frac{k+2}{k}\left[1 - \left(\frac{r}{R}\right)^k\right] \quad (8)$$

where

$$\overline{U}_B = 1.772\, d_p^{0.060}\, U_l^{0.024}\, U_g^{0.225} \quad (9)$$

$$k = 16.062\, d_p^{0.228} \quad (10)$$

Since bubbles may be broken up by small liquid eddies in turbulent flow, the bubble chord length should be reduced by an increase in liquid velocity which may produce the small-scale eddies in three-phase fluidized beds (Kang and Kim, 1986) as illustrated by Eq. 5. However, liquid velocity did not significantly affect the bubble rising velocity (Eq. 9) since the bubble velocity mainly depends on bubble size and bed viscosity.

Acknowledgment

We acknowledge a Grant-in-Aid of research from the Korea Science and Engineering Foundation.

Notation

D_T = column diameter, m
d_p = particle diameter, m
l_v = local vertical bubble length, m
l_{vc} = vertical bubble length at the center of column, m
l_{vw} = estimated vertical bubble length from extrapolating the value from r to R, m
$P(l_v)$ = probability density function of vertical bubble length, as

$$1/(\sqrt{2\pi}\,\beta(l_v)\exp\left[\frac{(\ln(l_v)-\alpha)^2}{2\beta^2}\right]$$

r = radial distance from the center of column, m
R = column radius, m
U_B = bubble rising velocity, m/s
$\overline{U}_B$ = mean bubble rising velocity in radial direction, m/s
U_g = superficial gas velocity, m/s
U_l = superficial liquid velocity, m/s

Greek letters

ϵ_g = local gas-phase holdup
$\overline{\epsilon}_g$ = mean gas-phase holdup

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Manuscript received Dec. 1, 1987, and revision received July 8, 1988.