Size of Bubbles Leaving a Three-Phase Fluidized Bed Containing Low Density Particles*

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Abstract The size of bubbles leaving a three-phase fluidized bed containing low density particles was investigated photographically with a video camera. The influence of gas and liquid flow rates, size and density of solid particles on the size of bubbles was also studied. The size of the bubbles is almost independent of gas velocity, but decreases with increasing liquid velocity. The bubble coalescence tends to occur in the bed rather than the bubble column. The size of bubbles can be roughly estimated from the bubble size in the bubble column, physical properties of the particles, and operating conditions.

Keywords : Bubble diameter, Bubble size distribution, Three phase fluidized bed, Bioreactor, Low density particle

1. INTRODUCTION

Three-phase fluidized beds used as bioreactors have several advantages over such reactors as stirred tank reactors, column reactors with mechanical mixing, etc. Mild mixing of solid particles by liquid or gas flow leads to a small loss of microorganisms and carrier enzymes due to low shear stress. Moreover, it is easy to prevent the contamination resulting from the mixing of another microorganisms. Three-phase fluidized bed processes, therefore, have been widely applied to the productions of various bio-products. However, very little information has been obtained concerning the three-phase fluidized beds of low density particles such as immobilized microorganisms, enzymes and biofilm particles used in biotechnological processes.

Schügerl (1989) and Lee, Miyahara and Takahashi (1990) reviewed the three-phase fluidized beds used as bioreactors and pointed out that the behavior of three-phase fluidized beds of low density particles is different from that of conventional ones of

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heavy particles, such as glass beads or metal particles.

In the present study, the size of bubbles from a three-phase fluidized bed of low density particles, which is related to gas-liquid mass transfer characteristics of bioreactor such as gas-liquid interfacial area and volumetric mass transfer coefficient, is investigated. Especially, the size of bubbles leaving the three-phase fluidized bed, which plays a significant role in the suspension of solid particles in the freeboard region, was examined.

2. EXPERIMENTAL

The schematic diagram of the experimental apparatus is shown in Fig.1. The main column is the same as that used by Miyahara, Lee and Takahashi (1990), with dimensions ; 6.5cm and 157cm in I.D. and height respectively and constructed of Perspex pipe. A perforated plate with nominal pore size of 0.5mm and 30 holes was used as the gas distributor. Standard mesh (300 μ m) was placed on the top of the column and liquid entrance to prevent the elutriation and leakage of particles, respectively. Pressure taps (17 in number) were mounted normal to the column wall to measure the axial static pressure profiles in the bed. A viewing box filled with water was also provided in order to eliminate optical distortion during recording of the bubbles with a video camera.

Fig.1 Schematic diagram of experimental apparatus.

Compressed humidified air and deionized water were used as gas and liquid phases, respectively. Solid particles which have small terminal velocities in liquid were used, namely, two types of 6-nylon particles, polystyrene particles and activated carbon particles. The properties of the particles are listed in Table 1. Water was circulated through the particle filled column and reservoir by means of a pump.

Table 1 Properties of particles employed

	d_p [mm]	ρ_p [kg/m ³]	U_t [mm/s]	Re_t' [-]
6-Nylon	6.32	1115.6	148	110
	9.52	1127.4	190	234
Activated carbon	0.65	1428.5*	44.3	12.4
Polystyrene	3.1	1043.0	46.7	6.47

* wet density

Air and water were introduced cocurrently from the bottom of the column and their flow rates were measured with orifice flow meter. The bubbles formed at the gas distributor left the three-phase fluidized bed in larger or smaller size than that at the moment of formation due to the coalescence or the breakup in the bed.

The bubbles leaving the three-phase fluidized bed were photographed near the fluidized bed boundary via a video camera. The major axis a and the minor axis b of each bubble were measured in the monitor to estimate its size.

For comparison, the size of bubbles in a bubble column, namely, gas-liquid system, was also measured, where both gas and liquid flow upward.

In most bioreactors, mild condition leading to low gas and liquid velocities was employed in order to improve the biological reaction rate and to prevent excessive sloughing of biofilms, and particle-particle attrition or particle elutriation.

In the present study, superficial gas and liquid velocities were in the ranges of $0.085 \leq U_{gc} \leq 2.10$ cm/s, and $0.51 \leq U_{lc} \leq 5.5$ cm/s, respectively; the operation regime was the complete three-phase fluidized or the partial suspension regime shown by Miyahara et al. (1990). Experiments were carried out at room temperature (20 ± 2 °C).

3. RESULTS AND DISCUSSION

The size of the bubbles was estimated as the equivalent spherical diameter and the volumetric mean diameter was calculated from the following equations using measured major and minor axes of each bubble which was assumed to be ellipsoidal in the shape (Miyahara, Matsuba and Takahashi, 1982).

$$\bar{d}_{30} = \left(\frac{\sum 6 V_b / \pi}{n} \right)^{1/3} \quad (1)$$

where

$$V_b = (\pi/6) a^2 b = (\pi/6) d^3 \quad (2)$$

Rearrangement of Eq.(2) gives the equivalent spherical diameter of bubble as follows

$$d = (a^2b)^{1/3} \quad (3)$$

The number of bubbles measured was about 100, a level which is presumed to be reliable for statistical analysis under the present experimental conditions (Koide, Hirahara and Kubota, 1966).

Fig.2 shows the result of volumetric mean diameter of bubbles calculated from Eq.(1) against the gas velocity through a hole in the distributor for the bubble column which has no solid particles. The solid line in the figure is the result of the bubble column without liquid flow obtained by Miyahara, Matsuba and Takahashi (1982). The volumetric mean diameters of bubbles were found to be independent of the gas velocity. This behavior shows a little deviation from the result of Miyahara, Matsuba and Takahashi (1982). It is attributed to the differences in operating conditions; this study was performed in the gas-liquid continuous flow system, while theirs was carried out in a non-flowing liquid medium.

Fig.3 shows the volumetric mean diameter of bubbles leaving the three-phase fluidized bed. The volumetric mean diameter of bubbles was also independent of the gas velocity, as in the bubble column. On the other hand, Ha and Kim (1980) investigated the bubble characteristics in a three-phase fluidized bed of glass beads. They showed that the bubble size in the bed increased with increasing gas velocity; their result is different from that of this study. This difference may be due to the fact that their study was performed for bubbles in a bed of heavy particles.

The volumetric mean diameter of bubbles leaving the bed decreases with increasing liquid velocity as shown in Fig.4. The result agrees with that of Rigby et al.(1970) who measured the bubble size in the bed by means of an electroresistivity probe. Similar results were also obtained by Page and Harrison(1972) who investigated the size distribution of bubbles leaving the three-phase fluidized bed of glass beads photographically.

Rigby et al. (1970) reported that the decrease in bubble size with increasing liquid velocity is related to the decrease in bed viscosity which plays a significant

Fig.2 Effect of gas velocity on volumetric mean diameter of bubbles in bubble column.

Fig.3 Effect of gas velocity on volumetric mean diameter of bubbles leaving the three-phase fluidized bed.

Fig.4 Effect of liquid velocity on volumetric mean diameter of bubbles leaving three-phase fluidized bed.

role in bubble characteristics.

Bruce and Revel-Chion(1974) studied the size of bubbles leaving the three-phase fluidized bed of glass beads of diameters; 2, 4, 6 and 8 mm. They reported that, for particles of diameters 2 and 4 mm at constant gas velocity, the size of bubbles emerging from the bed decreased when the liquid velocity increased. For large particles of 6 and 8 mm in diameter at constant gas velocity, however, the bubble diameter decreased as the liquid velocity increased, reaching a minimum at about 13.20cm/s. Further increase in liquid velocity caused the bubble diameter to increase.

In this study, however, the effect of the particle size on the size of bubbles was found to be negligible.

Fig.5 shows the cumulative size distribution of bubbles leaving the three-phase fluidized bed and in the bubble column. In each case, it is supposed that the data may follow a logarithmic normal probability distribution law.

Fig.5 Cumulative size distribution of bubbles leaving three-phase fluidized bed and bubble column.

The size of bubbles emerging from the three-phase fluidized bed, however, varies slightly with the different species of particles, and shows a little difference from that of the bubble column, that is, the slope of the line is more gentle than that in the bubble column. This result suggests that in the three-phase fluidized bed not only the coalescence, but the breakup of bubbles occurs incessantly as compared with the bubble column.

The logarithmic normal probability distribution function can be represented as follows;

$$\frac{d(1-R)}{d(\ln d/\bar{d}_g)} = \frac{1}{\sqrt{2\pi} \ln s} \exp \left[-\frac{\{\ln(d/\bar{d}_g)\}^2}{2(\ln s)^2} \right] \quad (4)$$

where $\bar{d}_g$ is the geometric mean diameter :

$$\bar{d}_g = (d_1 \cdot d_2 \cdot d_3 \cdot \dots \cdot d_n)^{1/n} \quad (5)$$

and s is the geometric standard deviation:

$$s = \exp \left[\frac{\sum \{\ln(d/\bar{d}_g)\}^2}{n} \right]^{1/2} \quad (6)$$

The relationship between $\bar{d}_{ji}$ and $\bar{d}_g$ of the logarithmic normal probability distribution is (Mugele and Evans, 1951)

$$\bar{d}_{ji} = \bar{d}_g \exp \{ (i+j)(\ln s)^2 / 2 \} \quad (7)$$

Fig.6 shows the geometric standard deviation of bubbles s . From the figure the geometric standard deviation slightly decreases with liquid velocity, and approaches the value of 1 in all cases. This suggests that the bubble size decreases with increasing liquid velocity as shown in Fig.4, and its distribution becomes comparatively homogeneous with liquid velocity.

The ratio of volumetric mean diameter of bubbles emerging from the three-phase fluidized bed to that in the bubble column is plotted against the Reynolds number in Fig.7. It can be seen from the figure that the ratio decreases with the Reynolds number, that is, it decreases with liquid velocity for all particles, and results in a

Fig.6 Geometric standard deviation.

Fig.7 $(\bar{d}_{30})_T/(\bar{d}_{30})_B$ vs. Re

value less than 1. This means that at high liquid velocities, the volumetric mean diameter of bubbles emerging from the fluidized bed is smaller than that from the bubble column. From this result, it is supposed that at low liquid velocities, the bubble shows a tendency to coalesce through the bed because it has the higher apparent viscosity due to the insufficient expansion of the bed. With increase in liquid velocity, however, the apparent viscosity of the bed decreases, the turbulent eddy and the instability of the interface between particulate phase and bubbles increases, which would be expected to reduce the bubble size. Consequently, at higher liquid velocities the tendency of the breakup of bubbles will be greater than that of coalescence, therefore bubble can split easily. From this figure, it can be noticed that the bubble coalescence tends to occur in the experimental range of the present study as compared with the bubble column.

Fig.8 shows the relationship between the Reynolds number and the modified particle Reynolds number at the volumetric mean bubble diameter ratio of about 1 in Fig.7. From this figure, it can be anticipated that the correlation between the Reynolds number and the modified particle Reynolds number is as follows:

$$Re \propto Re_t'^{0.346} \quad (8)$$

Fig.8 Re vs. Re_t'

Fig.9 Correlation of volumetric mean diameter of bubbles leaving three-phase fluidized bed.

Fig.9 shows the volumetric mean bubble diameter ratio against $Re/(Re_t')^{0.346}$. The volumetric mean bubble diameter ratio decreases by increasing the Reynolds number or decreasing the modified particle Reynolds number. It can also be supposed that the bubble size leaving the three-phase fluidized bed can be roughly estimated from, the bubble size in the bubble column, physical properties of particles and operation conditions. Though it is desirable to find the correlation using large columns and other liquids, it is suggested at this stage that the results obtained in this work should be useful for scale-up and/or other liquids because the volumetric mean bubble diameter ratio is correlated by dimensionless groups such as Re and Re_t' .

4. CONCLUDING REMARKS

In this study, from the video photographs of bubbles leaving the three-phase fluidized bed of low density particles, effects of gas and liquid velocities, size and density of solid particles on the size of bubbles were experimentally investigated, and the following results were obtained:

- 1) The size of the bubbles is almost independent of gas velocity, but is reduced by increase in liquid velocity.
- 2) The bubble size distribution can be expected to follow the logarithmic normal probability distribution law.
- 3) There is a high degree of bubble coalescence as compared to the bubble column.
- 4) The size of bubbles leaving the three-phase fluidized bed can be roughly estimated from, the bubble size in the bubble column, physical properties of solid particles and operation conditions.

NOMENCLATURE

a =major axis of bubble
 b =minor axis of bubble
 D_c =column diameter
 d =equivalent spherical bubble diameter

[m]
 [m]
 [m]
 [m]

$\bar{d}_g$ =geometric mean diameter

of bubbles

$\bar{d}_{ji}$ =mean diameter of bubbles

d_p =diameter of solid particle

$\bar{d}_{30}$ =volumetric mean diameter

of bubbles

[m]
 [m]
 [m]
 [m]
 [m]

n =number of bubbles	[—]	U_{gh} =gas velocity through hole	[m/s]
R =cumulative number fraction greater than stated size [%] ,	[—]	U_{lc} =superficial liquid velocity	[m/s]
Re =Reynolds number ($=D_c U_{lc} \rho_l / \mu_l$)	[—]	U_t =terminal velocity of solid particle	[m/s]
Re'_t =modified particle Reynolds number ($=d_p U_t (\rho_p - \rho_l) / \mu_l$)	[—]	V_b =bubble volume	[m ³]
s =geometric standard deviation	[—]	μ_l =viscosity of liquid	[Pa·s]
U_g =superficial gas velocity	[m/s]	ρ_l =density of liquid	[kg/m ³]
		ρ_p =density of solid particle	[kg/m ³]

Subscripts

T : three-phase fluidized bed

B : bubble column

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