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Hydrodynamics and Mixing in Three-Phase Fluidized Bed Bioreactors

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ABSTRACT

The effects of solid particles on hydrodynamics and mixing in a three-phase fluidized bed bioreactor were discussed. The gas holdup, bubble size, and liquid-phase axial dispersion coefficient were measured in a 0.25-m id bubble column bioreactors containing low-density particles. The presence of low-density solid particles slightly increased gas holdup. The decrease in average bubble diameter with solid presence was found. For the three-phase system, the liquid-phase axial dispersion coefficients were higher than for the two-phase system.

We extended a model for a gas holdup developed for a gas–liquid two-phase bubble column bioreactor to a gas–liquid–solid three-phase fluidized bed bioreactor. Using the present data and available data in the literature, the predictions of the proposed model were examined. The proposed model agreed with a wide range of the experimental data. A theoretical correlation for liquid-phase axial dispersion coefficient was developed using Kolmogoroff's theory of isotropic turbulence. Reasonable agreement was obtained between the predicted and experimental values of axial dispersion coefficient.

Index Entries: Three-phase fluidized bed bioreactor; gas holdup; liquid-phase axial dispersion coefficient; bubble size.

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NOMENCLATURE

D_c reactor diameter, m
D_p bubble size, m
E_z liquid-phase axial dispersion coefficient, m^2/s
g gravitational acceleration, m/s^2
l_c characteristic length, m
U_g superficial gas velocity, m/s
U_l superficial liquid velocity, m/s
V_g gas volume in three-phase fluidized bed bioreactor, m^3
V_l liquid volume in three-phase fluidized bed bioreactor, m^3
V_s solid volume in three-phase fluidized bed bioreactor, m^3
Q_g volumetric gas flow rate, l/min
ϵ energy dissipation rate per unit mass of liquid, W/kg
μ_l liquid viscosity, Pa·s
$\overline{\mu}_l$ apparent viscosity of solid-liquid two-phase system, Pa·s
ρ_l liquid density, kg/m^3
ρ_s solid density, kg/m^3
$\overline{\rho}_l$ mean density of solid-liquid two-phase system, kg/m^3
$\overline{\rho}$ mean density of gas-liquid two-phase system or gas-liquid-solid three-phase system, kg/m^3
ϕ_g gas holdup
ϕ_l liquid holdup in gas-liquid-solid three-phase system $(=1 - \phi_s - \phi_g)$
ϕ_l' liquid holdup in solid-liquid two-phase system $(=1 - \phi_s')$
ϕ_s solid holdup in gas-liquid-solid three-phase system
ϕ_s' solid holdup in solid-liquid two-phase system
χ constant

INTRODUCTION

Gas-liquid-solid three-phase fluidized bed reactors are widely used in petrochemical and coal processing. Their recent application is in the field of biotechnology processes. Suspended solids in biological processes are cells of microorganisms immobilized on solid-support particles, microbial, plant or animal cells, and filamentous organisms. Much attention has been devoted to the phenomena of three-phase fluidized bed bioreactors (e.g., *1-8*). However, there are conflicting results indicating an increase or decrease in gas holdup and mass-transfer rates with solids loading as described later in detail. It is clear that the published information on three-phase fluidized bed bioreactors is far from complete.

The knowledge of hydrodynamic characteristics is crucial for the design of three-phase fluidized bed bioreactors, and they have been extensively studied (*2,9*). Nevertheless, very few address the theoretical aspects. Although model representation of hydrodynamic and mixing behavior is quite difficult, theoretical studies can be expected to provide

insight into the mechanism of phenomena in three-phase fluidized bed bioreactors. Even if theoretical studies cannot elucidate the complete picture of the phenomena, they may at least improve our understanding of the behavior of three-phase fluidized bed bioreactors.

A three-phase (gas–liquid–solid) fluidized bed bioreactor can be modeled as a hypothetical two-phase (gas-[liquid + solid]) bioreactor. The continuous phase contains the solid–liquid suspension, whereas the dispersed phase consists of the rising gas bubbles.

The purpose of this work is to measure hydrodynamics and mixing in a three-phase fluidized bed bioreactor containing low-density particles. A few studies have been conducted on suspended solids having the density close to that of the liquid nutrient medium. Low-density particle systems, such as polymeric gel particles with cells entrapped within their inner porous structure and particles with biofilm attached to their surface, are common for bioprocesses. In the cases of mycelial broths, the cells grow to sizes of several hundred microns, and they must be treated as low-density suspensions. The gas holdup, bubble size, and liquid-phase axial dispersion coefficient were measured in a 100-L column bioreactor containing polycarbonate solid particles.

A model for gas holdup in a bubble column bioreactor developed by Kawase et al. (*10*) was extended to a three-phase fluidized bed bioreactor. A theoretical correlation for the axial dispersion coefficient was also derived. The proposed correlations were discussed using a wide range of the data obtained in this work and the available experimental results in the literature.

EXPERIMENTAL SECTION

Experiments were carried out in a bubble column bioreactor having an inside diameter of 0.25 m and total height of 2.0 m. The reactor was operated in a semibatch manner with known volumes of liquid and solid forming the batch through which air was continuously sparged. Tap water was used as the liquid. The superficial gas velocity was studied in the range of 0.0037 to 0.031 m/s. All measurements were performed in the bubbly flow and the churn-turbulent flow regimes (*11*). The sparger consisted of a ring sparge (ring diameter = 0.24 m) with 12 holes of 1.0 mm diameter.

Fluidized bed particles consisted of cylinders of polycarbonate (3.0×2.5 mm length and diameter) having a density of 1050 kg/m^3. The effect of solids loading was studied in the range 0–8 vol%. Fractional gas holdup was measured by noting the difference between the dispersed and clear liquid heights.

Bubble sizes were determined photographically. Average bubble diameters were obtained as the volume-to-surface mean bubble diameters. A

minimum of 30 bubbles were examined to estimate the average bubble diameter.

The values of liquid-phase axial dispersion coefficient, which is a quantitative description of the mixing in three-phase fluidized bed bioreactors, were measured using a pulse technique. A pulse of acid solution (HCl solution) of known volume and concentration was added to the liquid surface at the top of the column. A pH probe was placed near the reactor bottom, and the pH responses were recorded with the help of a chart recorder. The values of liquid-phase axial dispersion coefficient were determined by matching the experimental response curve with the theoretical one (*12*).

THEORY

Gas Holdup

We apply the model presented by Kawase et al. (*10*) for a gas–liquid two-phase bubble column bioreactor. It is based on the concept of turbulent natural convection. The liquid circulation induced in a bubble column bioreactor is considered to be owing to the density difference resulting from local variations of gas holdup. In their analysis, the energy dissipation rate in Newtonian fluids is given as a function of gas holdup:

$$\epsilon = 2^{16/3} (\mu_l/\rho_l)^{1/3} g^{4/3} (\Delta \rho / \bar{\rho})^{4/3} \quad (1)$$

In developing a model for the gas holdup in the three-phase fluidized bed bioreactor, the gas–liquid–solid three-phase system is treated as a hypothetical gas–liquid two-phase system. The solid suspension is postulated to be a pseudo-homogeneous liquid.

In order to apply Eq. (1) to the gas–liquid–solid three-phase system, the viscosity and density must be modified. In other words, it is required to replace the liquid viscosity and density, μ_l and ρ_l, with those for the pseudo-homogeneous liquid, $\bar{\mu}_l$ and $\bar{\rho}_l$, respectively. The viscosity and density of pseudo-homogeneous liquids are expressed as functions of the volume fraction of solids, ϕ_s'. This modification may be valid when the solid phase is in uniformly suspended condition. If a solid distribution is uniform, the present model may be applicable to high-density particle systems as well as low-density particle systems.

The viscosity of suspensions, $\bar{\mu}_l$, is assumed to be dependent on the volume fraction of solid ϕ_s'. Kawase and Ulbrecht (*13*) proposed a theoretical expression for the apparent viscosity of suspensions using the cell model. Their correlation for Newtonian suspensions is written as:

$$\bar{\mu}_l = \mu_l [(1 + 8.203 \phi_s'^5)/(1 - 2.478 \phi_s' + 18.456 \phi_s'^5 - 20.326 \phi_s'^6)] \quad (2)$$

This correlation is in excellent agreement with Rutgers' curve (*14*) obtained by consolidating all the data of the many investigators. At very low solid concentration ($\phi_s' \rightarrow 0$), this equation reduces to the linear form:

$$\bar{\mu}_l = \mu_l (1 + 2.48\, \phi_s') \tag{3}$$

This is very close to Einstein's expression.

The mean density for solid–liquid two-phase systems (pseudo-homogeneous liquids) is given as:

$$\bar{\rho}_l = \rho_s \phi_s' + \rho_l \phi_l' = \rho_s \phi_s' + \rho_l (1 - \phi_s') \qquad (\phi_s' + \phi_l' = 1) \tag{4}$$

Meanwhile, the mean density of gas–liquid–solid three-phase systems (pseudo-homogeneous suspension phase containing bubbles) is estimated by

$$\begin{aligned} \bar{\rho} &= \rho_s \phi_s + \rho_l \phi_l + \rho_g \phi_g \qquad (\phi_s + \phi_l + \phi_g = 1) \\ &\approx \rho_s \phi_s + \rho_l \phi_l \qquad (\rho_g << \rho_s, \rho_l) \\ &= (1 - \phi_g)\, [\rho_s \phi_s' + \rho_l (1 - \phi_s')] \end{aligned} \tag{5}$$

It should be noted that in order to derive Eq. (5), the following relationship between the solid holdup in a solid–liquid two-phase system (pseudo-homogeneous liquid), ϕ_s', and that in a gas–liquid–solid three-phase system, ϕ_s, is used.

$$\phi_s / \phi_s' = [V_s/(V_s + V_l + V_g)]/[V_s/(V_s + V_l)] = 1 - \phi_g \tag{6}$$

Although ϕ_s' can be calculated from the solid loading, ϕ_s must be estimated from the measurements of ϕ_l and ϕ_g. Since, therefore, the resulting correlation for gas holdup should be given as a function of ϕ_s' rather than ϕ_s, $\bar{\rho}_l$ and $\bar{\rho}$ are also given as functions of ϕ_s', Eqs. (4) and (5).

The difference between the density of the suspension phase and that of the three-phase is:

$$\Delta \rho = \bar{\rho}_l - \bar{\rho} = \phi_g\, [\rho_s \phi_s' + \rho_l (1 - \phi_s')] \tag{7}$$

From Eqs. (5) and (7), we have:

$$\Delta \rho / \bar{\rho} = \phi_g / (1 - \phi_g) \tag{8}$$

This is the same as the relationship for gas–liquid two-phase systems (*10*).

Substituting Eq. (8) into Eq. (1) and rearranging yield:

$$\phi_g / (1 - \phi_g) = 2^{-4}\, (\bar{\mu}_l / \bar{\rho}_l)^{-1/4}\, g^{-1}\, \epsilon^{3/4} \tag{9}$$

where μ_l and ρ_l in Eq. (1) are replaced by $\bar{\mu}_l$ and $\bar{\rho}_l$, respectively. Knowledge of the value of ϵ is required to determine the gas holdup using Eq. (9). The energy dissipation rate in a three-phase fluidized bed bioreactor is obtained by considering the energy input rate and the energy recovery rate owing to the increase in potential energy of the liquid phase (*15*). The energy dissipation rate per unit mass of liquid is:

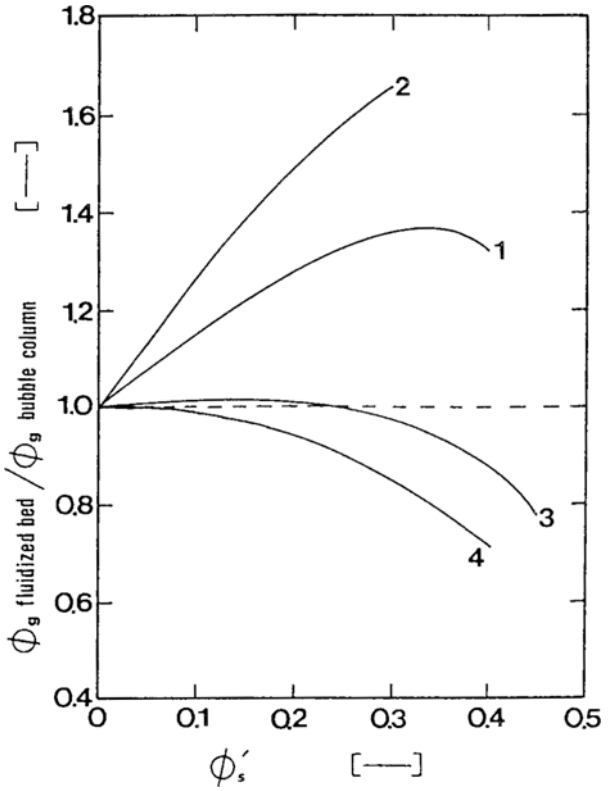

Fig. 1. Gas holdup in three-phase fluidized bed bioreactor ($\rho_1 = 1000$ kg/m^3, $\mu_1 = 0.001$ Pa·s and $U_g = 0.01$ m/s). 1: $\rho_s = 2500$ kg/m^3, $U_1 = 0$ m/s; 2: 2500 kg/m^3, 0.01 m/s; 3: 1050 kg/m^3, 0 m/s; 4: 750 kg/m^3, 0.01 m/s.

$$\epsilon = g\,((U_1 + U_g)\,\{(1-\phi_g)\,\phi_s'\,\rho_s + [1-\phi_g - (1-\phi_g)\,\phi_s']\,\rho_1\} - U_1\rho_1)\,/$$
$$\{1-\phi_g - (1-\phi_g)\,\phi_s'\}\,\rho_1$$

In bubble column bioreactors ($\phi_s' = 0$ and $U_1 \to 0$), Eq. (10) reduces to:

$$\epsilon = gU_g \tag{11}$$

Since ϵ is obtained as a complex function of ϕ_g (Eq. [10]), the resulting correlation, Eq. (9), gives ϕ_g implicitly, requiring access to a computer to use it conveniently.

Figure 1 shows the change in the gas holdup with the solids concentration. The present model predicts that the gas holdup in a fluidized bed reactor of high density particles ($\rho_s = 2500$ kg/m^3) increases with increasing solids concentration for $\phi_s' < 0.32$. A maximum value of ϕ_g is reached at $\phi_s' = 0.32$. At higher values of ϕ_s'(> 0.32), the gas holdup decreases with an increase in ϕ_s'. This is probably because the gas holdup experiences two opposing effects from the presence of solids. Initially, when ϕ_s' increases, the apparent density of the suspension increases, and hence,

local turbulence intensity and bubble breakup nature increase. Consequently, the average bubble diameter decreases, and as a result, the gas holdup increases. Further increase in ϕ_s' significantly increases the apparent viscosity of the suspension and the bubble coalescence nature. Therefore, the gas holdup decreases with an increase in ϕ_s'. The overall effect of the above two factors results in a maximum value in ϕ_g. Similar trends were observed by Khare and Joshi (*16*). Up to a loading values of 0.6 wt%, ϕ_g increased with solids loading (alumina and glass particles, $\rho_s = 4000$, 2500 kg/m^3). Above this loading, ϕ_g continuously decreased with an increase in the solids loading. The proposed model, Eq. (9), can theoretically predict this peculiarity of gas holdup in three-phase fluidized bed bioreactors.

When the density of solids is close to that of the liquid ($\rho_s = 1050$ kg/m^3), the gas holdup in three-phase fluidized bed reactors is slightly higher than in bubble column reactors up to ϕ_s' value of 0.22. Above this value of ϕ_s', that is lower compared to ϕ_g in bubble column bioreactors.

In the case of low-density particles ($\rho_s = 750$ kg/m^3), the gas holdup decreases continuously with ϕ_s'.

It can be also seen from Fig. 1 that the liquid flow ($U_l = 0.01$ m/s) enhances the effect of the presence of solids on gas holdup.

Liquid-Phase Axial Dispersion Coefficient

Kawase and Moo-Young (*17*) obtained a relationship for the liquid-phase axial dispersion coefficient in a gas–liquid two-phase system using Kolmogoroff's theory of isotropic turbulence.

$$E_z = l_c^{4/3} (\chi^4 \epsilon)^{1/3} \tag{12}$$

where $l_c = D_c/4$ and $\chi = 2$. Substitution of Eq. (9) into Eq. (12) gives:

$$E_z = (2g)^{4/9} D_c^{4/3} (\bar{\mu}_l/\bar{\rho}_l)^{1/9} [\phi_g/(1 - \phi_g)]^{4/9} \tag{13}$$

This correlation enables the calculation of the liquid-phase axial dispersion coefficient in a three-phase fluidized bed bioreactor as a function of gas holdup. Eq. (13) indicates an increase in E_z with increased ϕ_g.

RESULTS AND DISCUSSION

Gas Holdup

Figure 2 shows the variation of gas holdup with superficial gas velocity. The gas holdup was found to increase slightly only when a small volume fraction of solid particles was added. This trend agrees well with the predictions of Eq. (9). Recently, Bly and Worden (*6*) measured the gas holdup in a three-phase fluidized bed bioreactor. In their work in which gel beads ($\rho_s = 1050$ kg/m^3) were used, as well as the present experimental

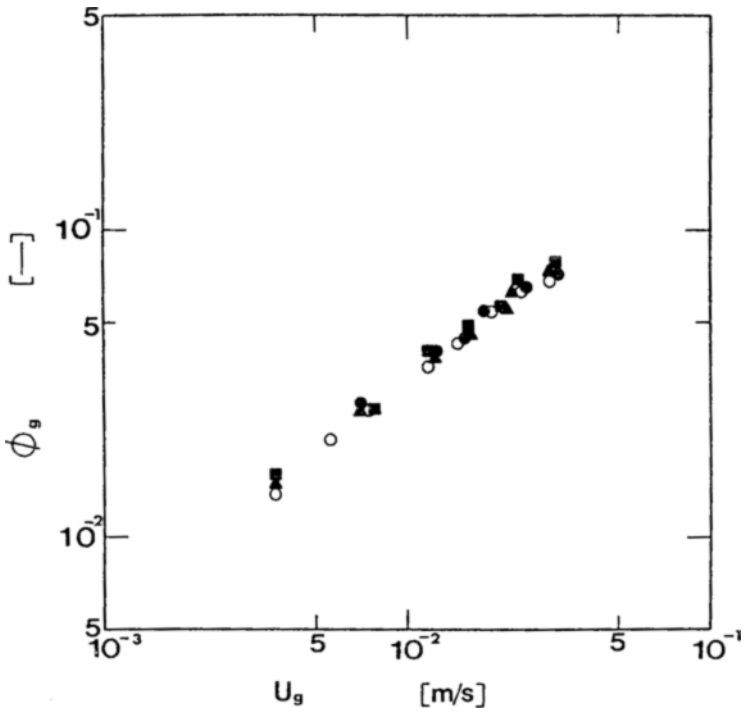

Fig. 2. Gas holdup in a three-phase fluidized bed bioreactor. ○ 0%, ■ 2%, ▲ 5%, ● 8%.

data and the predictions of Eq. (9), the gas holdup slightly increased with an increase in the solids concentration, and the liquid velocity, U_l, enhanced the effect of solids loading. In the literature, there are differences in trends of the gas holdup that have not been rigorously analyzed. Bukur et al. (*18*) found that the gas holdups in the presence of 30 wt% solids (iron oxide particles, $\rho_s = 5100$ kg/m^3) are about 25% larger than those obtained without solids during batch mode of operation. Sada et al. (*19,20*) also obtained similar findings in a slurry bubble column with distilled water and electrolyte solutions (alumina particles, $\rho_s = 3850$, 2240 kg/m^3). An increase in gas holdup owing to solids loading (alumina-silica cylinder, $\rho_s = 1250$ kg/m^3) was found in a loop three-phase fluidized bed reactor by Herskowitz and Merchuk (*21*). These results are in agreement with predictions of the proposed model. On the other hand, a decrease in gas holdup with an increase in solids concentration has been reported. Tang and Fan (*22*) (polystyrene, acrylic, acetate, and nylon particles, $\rho_s = 1050 - 1300$ kg/m^3), Smith and Ruether (*23*) (glass particles, $\rho_s = 2420$, 3990 kg/m^3), and Koide et al. (*24*) (glass particles, $\rho_s = 2500$ kg/m^3) found the presence of solid particles decreased ϕ_g. Their results are not consistent with the present data. It should be noted that in larger column reactors ($D_c > 0.218$

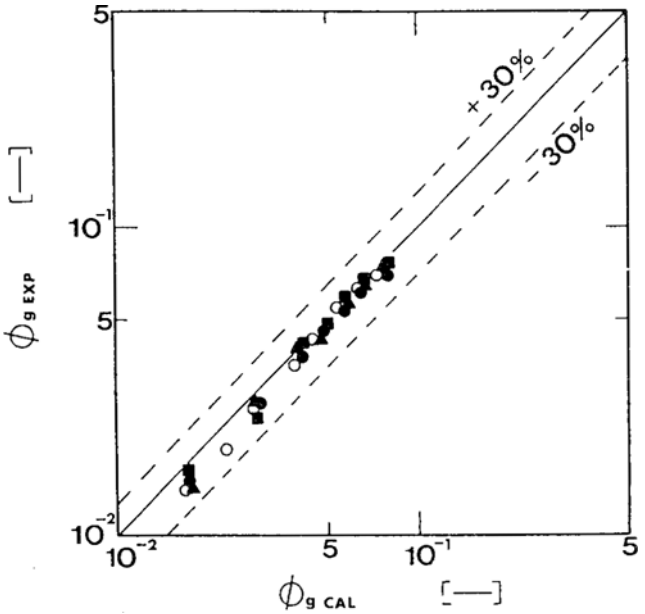

Fig. 3. Comparison between present experimental data and predicted values of gas holdup (symbols as in Fig. 2).

m), only very little change of ϕ_g with solid presence was reported by Koide et al. (24).

The coalescing nature may be increased by increasing the apparent liquid viscosity with solid particles. Therefore, the reduction of the gas holdup is expected. In this work, on the contrary, the addition of low-density particles slightly increased the gas holdup. They might enhance not only bubble coalescence, but also bubble breakup and, hence, result in small bubbles that offset gas holdup reduction resulting from large bubbles. As a result, the influence of solid particles on overall gas holdup was insignificant in this work. Eq. (9) predicts weak dependence of the gas holdup on the apparent viscosity. Since the power dependency for μ_l is $-1/4$, an increase in apparent viscosity owing to the presence of solids is expected to decrease the gas holdup only slightly.

Figure 3 compares the present experimental data with Eq. (9). As seen in the figure, the proposed correlation is found to fit the data reasonably well, the average deviation being about 7.6%.

In Fig. 4, the predicted values of ϕ_g by Eq. (9) are compared with the data obtained in three-phase fluidized bed reactors (no net liquid flow, $U_l = 0$) (24–26). The ranges for ϕ_s', μ_l, and $\rho_s =$ are 0.02–0.425, 0.000894–0.0176 Pa·s, and 817–2830 kg/m^3, respectively. The ϕ_g values calculated by Eq. (9) agree with the experimental data with an average error of 21.7%.

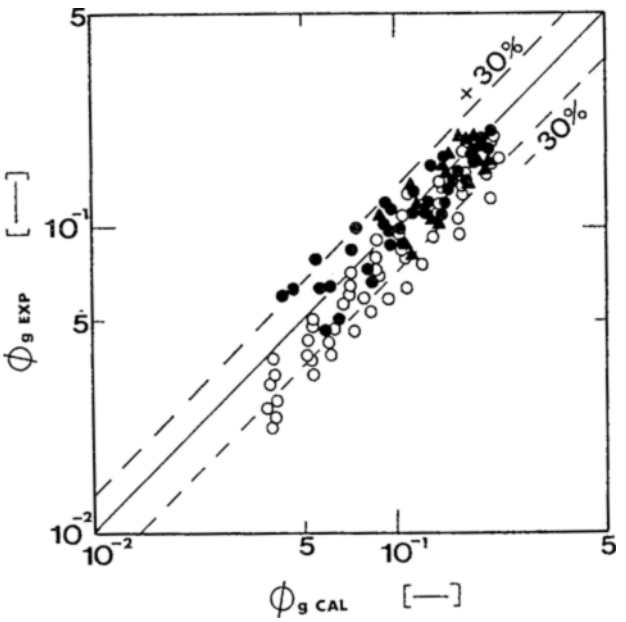

Fig. 4. Comparison between experimental data in the literature and predicted values of gas holdup ($U_l = 0$ m/s). ● Koide et al. (1984) (*24*) $\rho_s = 2500$ kg/m^3, $\phi_s = 0.04$–0.08 (water, glycerol aq. sol.). ○ Muroyama et al. (1990) (*26*) $\rho_s = 817$ kg/m^3, $\phi_s = 0.0568$–0.425 (water). ▲ Nigam and Schumpe (1987) (*25*) $\rho_s = 2830$ kg/m^3, $\phi_s = 0.02$–0.16 (water).

The gas holdups in three-phase fluidized bed reactors ($U_l \neq 0$) (*5,22,23,27,28*) are compared in Fig. 5 with Eq. (9). The data presented in the figure cover the range:

$$0.047 \leq \phi_s' \leq 0.40$$
$$700 \leq \rho_s \leq 3990 \text{ kg/m}^3$$
$$0.0009 \leq \mu_l \leq 0.0066 \text{ Pa·s}$$
$$0.001 \leq U_l \leq 0.01 \text{ m/s}$$

The average error of prediction of Eq. (9) is 25.1%.

Koide et al. (*24*) proposed an empirical correlation for ϕ_g, which agreed with their own data with an average error of only 5.7%. However, their complicated correlation fitted the experimental data in a slurry bubble column reported by Yasunishi et al. (*27*) with an average deviation of 28%.

Bubble Size

Figure 6 illustrates the effect of solid particles on D_p in the fluidized bed bioreactor of low-density particles. The bubble size in the two-phase systems decreased with increasing gas flow rate. In the three-phase systems, a lesser dependence with gas flow rate was found. From the figure, it is clear that the bubble size was smaller in the presence of particles as

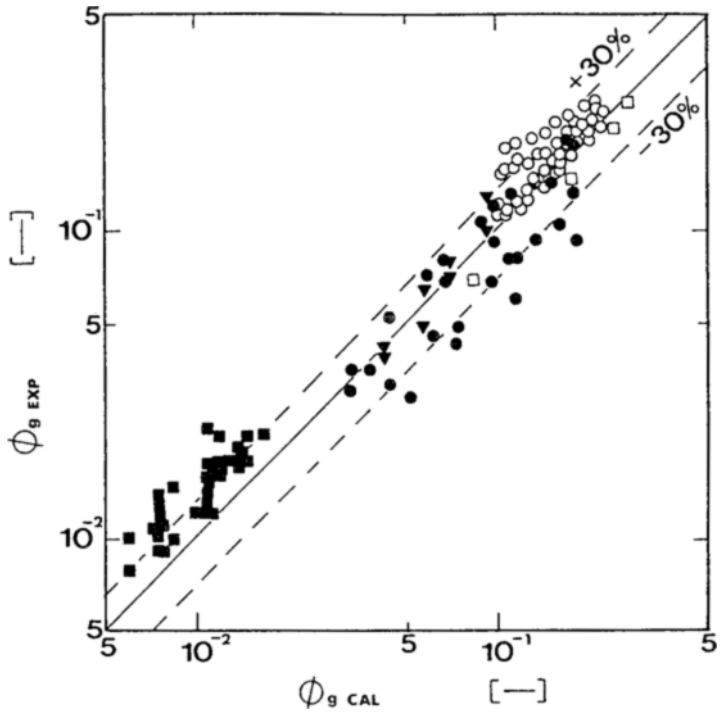

Fig. 5. Comparison between experimental data in the literature and predicted values of gas holdup ($U_1 \neq 0$ m/s). □ Smith and Reuther (1985) (*23*) ρ_s = 2420, 3990 kg/m³, ϕ_s=0.10 (water). ■ Yasunishi et al. (1986) (*27*) ρ_s = 2500 kg/m³, ϕ_s=0.13–0.40 (water, glycerol aq. sol.). ▼ Tang and Fan (1990) (*32*) ρ_s = 1050 kg/m³, ϕ_s=0.047, 0.137 (water). ■ Bajpai et al. (1990) (*5*) ρ_s = 1020 kg/m³, ϕ_s=0.16– 0.23 (electrolyte sol.) ○ Jamialahmadi and Muller-Steinhagen (1991) (*28*) ρ_s = 700–1360 kg/m³, ϕ_s=0.05, 0.10 (water).

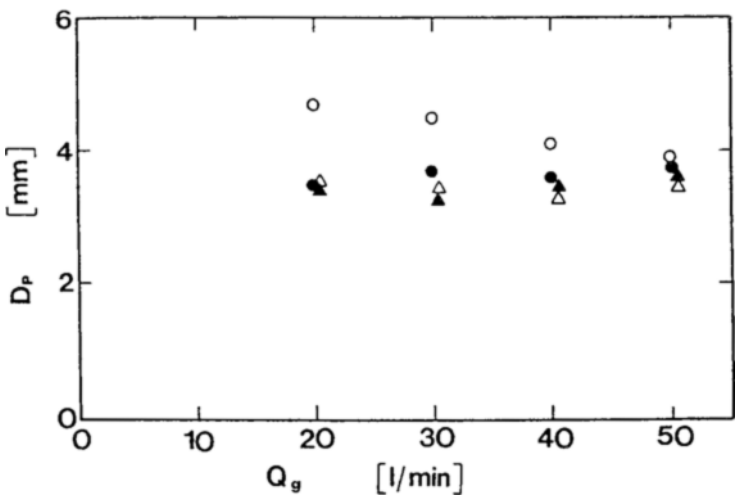

Fig. 6. Bubble size in a three-phase fluidized bed bioreactor. ○ 0%, ● 2%, ▲ 5%, △ 8%.

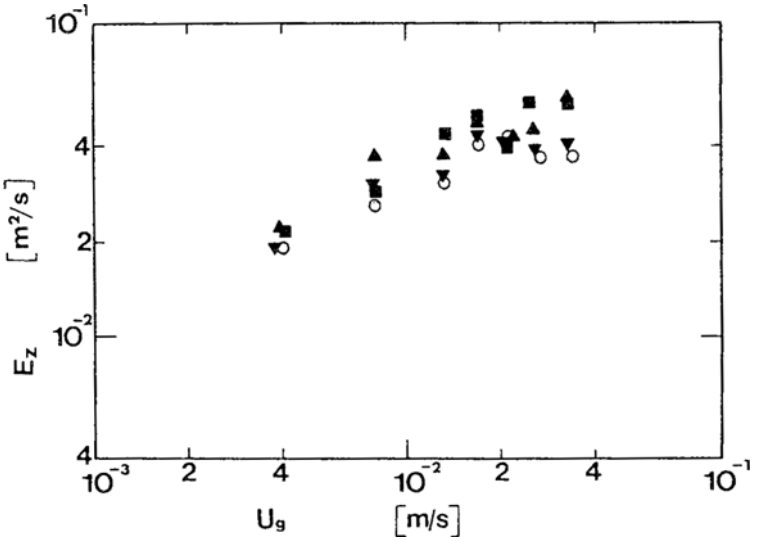

Fig. 7. Liquid-phase axial dispersion coefficient in a three-phase fluidized bed bioreactor. ○ 0%, ▲ 2%, ▼ 5%, ■ 8%.

compared to that in the absence, which is consistent with the data reported in the literature (*29*). At low gas flow rates, the bubble size reduction with solids addition was significant. The reduction in bubble size as marginally affected by changing the solids concentration from 2 vol% to 8 vol%. The smaller bubble sizes have relatively lower rising velocities and result in higher gas holdup. This is in agreement with the gas holdup data in Fig. 2.

It should be noted that the photographic technique used to measure bubble size in this work was limited to the bubbles rising near the bioreactor wall. Since large bubbles tended to rise in the center core of the bioreactor, the average bubble size might be somewhat underestimated.

Axial Dispersion Coefficient

The experimental values of E_z plotted against the superficial gas velocity are shown in Fig. 7. This figure indicates that that E_z values increased owing to the presence of small amounts of solid particles. The mixing intensity in the three-phase systems is suspected to be higher than in the two-phase system. In general, decreasing bubble size decreases the axial liquid-phase mixing. This contrasts with the present result. In this work, the presence of particles enhanced the large-scale liquid circulation flow intensity and, hence, increased the liquid-phase axial dispersion coefficient. The present results are consistent with the data reported by El-Temtamy et al. (*30*). However, contradictory results were obtained by Kim and Kim (*31*). Recently, Tang and Fan (*32*) found an increase in E_z in the three-phase system with a porous plate gas dis-

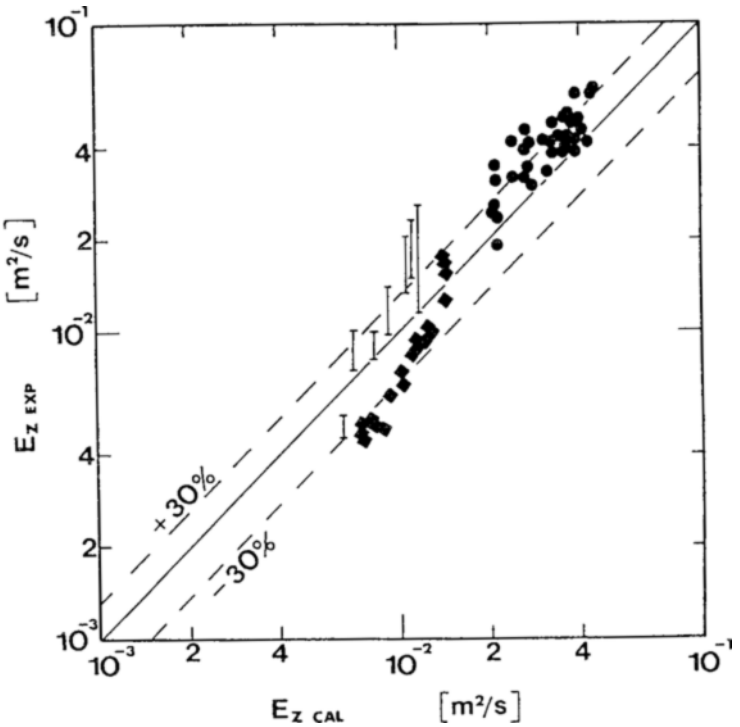

Fig. 8. Comparison between experimental and predicted values of liquid-phase axial dispersion coefficient. ● present work, ◆ Matsumoto et al. (1988) (*34*), I Kojima et al. (1986) (*33*).

tributor, but the reduction of E_z with a concentric gas–liquid distributor owing to the presence of low-density particles. Incidentally, several studies have been devoted to aeration devices, since they sometimes have significant effects on the mixing in bioreactors. However, no clear understanding of the effect of gas sparger designs has been presented.

The present model, Eq. (13), is compared with the present data and the available experimental data (*33,34*) in Fig. 8. Reasonable agreement was found. Most of the data lie within 30% of the correlation.

CONCLUSIONS

The solids loading of polycarbonate solid particles was found to affect the gas holdup only slightly. The gas holdup in the gas–liquid–solid system was slightly higher than in the gas–liquid system. The bubbles in the presence of solids were smaller than those in the absence of solids. Furthermore, the addition of particles having low density was found to increase the liquid-phase axial dispersion coefficient. These results are probably

the result of increased local turbulence with solid particles and, as a result, increasing overall liquid motion.

The proposed model for gas holdup could describe the peculiarity of gas holdup caused by solids loading. The proposed correlations for gas holdup and axial dispersion coefficient were discussed in light of the results of this work and the published literature. They were found to fit data reasonably well over a wide range of ρ_s, μ_l, and ϕ_s'.

Because of the complexity of cellular systems, our research began with inert well-defined particles. Our continuing work will involve studies on cellular systems themselves.

As described previously, there are a number of discrepancies among the results reported in the literature that make a unified analysis difficult. In a rigorous fashion, modeling gas holdup and liquid-phase axial dispersion coefficient in a three-phase fluidized bed bioreactor is not possible for now. Since the effects of particle size, particle shape, and particle surface properties are not considered in developing the proposed correlations, they should be preliminary. Further progress in the modeling of gas holdup and liquid-phase axial dispersion coefficient is highly desirable. However, they provide a successful theoretical framework and a promising methodology for further study. In order to develop more precise correlations, synergistic effects of particle size, particle shape, and particle surface properties must be studied in a systematic manner. Furthermore, Eq. (10) should be modified. Eq. (10) has been widely used to estimate the energy dissipation rate in a three-phase fluidized bed reactor (*15,35,36*). In the presence of solids, viscous dissipation occurs at the solid–liquid interface (*29*). This additional energy dissipation mode is not considered in Eq. (10) for simplification. This work was undertaken as a starting point for systematic studies of the influence of solids loading on the performance of a three-phase fluidized bed bioreactor.

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