AXIAL HOLDUP DISTRIBUTIONS OF GAS AND SOLID PARTICLES IN THREE-PHASE FLUIDIZED BED FOR GAS-LIQUID (SLURRY)-SOLID SYSTEMS

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Key Words: Chemical Reactor, Slurry Reactor, Three Phase Fluidization, Fluidized Bed, Bubble Column, Mixing, Dispersion Coefficient, Gas Holdup, Solid Holdup, Entrainment

The behavior of gas and solid particles was studied in a so-called three-phase fluidized bed of 12-cm diameter. The mean gas holdups in the dense and lean regions and the mean solid holdup in the dense region were correlated by experimental equations. The axial distribution of solid holdup in the lean region was analyzed by the sedimentation diffusion model. The mean settling velocity, v_p, and the longitudinal dispersion coefficient of solid particles, E_p, were correlated with previous data for bubble columns with suspended solid particles, respectively, as follows.

For $U_g/v_t<30$:

$$v_p=v_t\{1+1.5(U_g/v_t)\}^{0.3}\psi_l^{2.5}$$

For $(d_pv_t/v_l)=0.3$–2.5:

$$U_gD_T/E_p=\{13Fr/(1+8Fr^{0.85})\}\{1+0.009(d_pv_t/v_l)Fr^{-0.8}\}$$

For $(d_pv_t/v_l)=2.5$–640:

$$U_gD_T/E_p=\{13Fr/(1+8Fr^{0.85})\}(1+0.023Fr^{-0.8})$$

where D_T = column diameter; d_p = particle diameter; $Fr=U_g/\sqrt{gD_T}$; g = acceleration of gravity; U_g = superficial gas velocity; v_t = terminal velocity of solid particle; v_l = kinematic viscosity of liquid or slurry; ψ_l = (volume fraction of liquid or slurry)/(1-volume fraction of gas).

Introduction

For three-phase systems where the density is in the order of solid > liquid > gas, solid particles are suspended or fluidized by cocurrent upward flows of gas and liquid.[5]

When the terminal velocity of solid particles is considerably smaller than the bubble rising velocity, solid particles are suspended mainly by the gas flow. This type of bubble column with suspended solid particles[8] is normally operated with a continuous feed and discharge of solid particles.

When the terminal velocity of solid particles exceeds several $\text{cm}\cdot\text{s}^{-1}$, a certain liquid velocity is needed to maintain the fluidization of solid particles. This is a so-called three-phase fluidized bed[5,9] and is operated batchwise with respect to solid particles.

Recently, these three-phase contactors have been attracting much attention as the main reactor in catalytic, biological and coal conversion[16] processes. Most research studies, however, are on bubble columns with suspended solid particles. The general characteristics of the three-phase fluidized bed are still poorly understood.[17]

The so-called three-phase fluidized bed is divided into two regions. One is a lower region (dense region) with concentrated solid particles and the other is a freeboard region (lean region) with an axial distribution of solid holdup.[2,14] Prediction of the solid holdup in the lean region is essential for design of the height of liquid outlet. El-Temtamy and Epstein[3] analyzed the axial distribution of solid holdup in the lean region on the basis of their wake model. At higher gas velocities, however, bubble wakes may lose their identity, and the movement of solid particles becomes more diffusive.

In this work, water and slurries are used as the fluidizing media. The holdups of gas and solid particles in the three-phase fluidized bed are measured. The holdups of gas in the dense and lean regions and the holdup of solid particles in the dense region are correlated by experimental equations. Axial distribution of solid holdup in the lean region is analyzed by the sedimentation diffusion model.[4,8,15] The mean settling velocity and the longitudinal dispersion coef-

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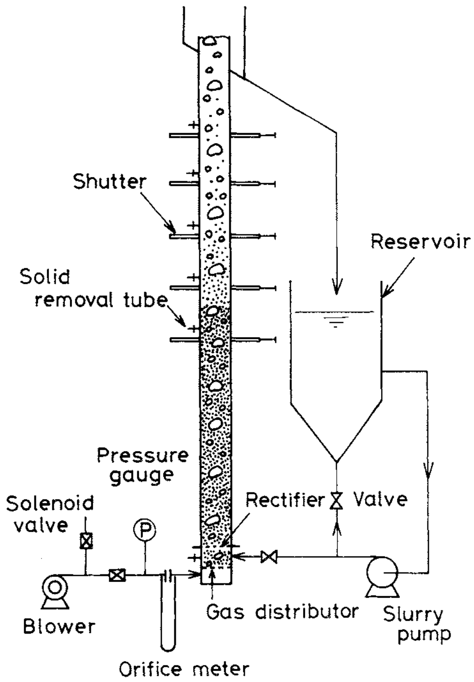

Fig. 1. Schematic diagram of experimental apparatus.

ficient of solid particles are correlated with previous data[8] for bubble columns with suspended solid particles.

1. Experimental

A schematic view of the experimental equipment is shown in **Fig. 1**. The main column, made of transparent acrylic resin, was 12 cm in diameter and 216 and 266 cm in height. The gas distributor was a double perforated plate in which a 28-mesh metal net was sandwiched to prevent the fall of solid particles. The perforated plates has 48 holes of 1-cm diameter drilled in a trigonal arrangement of 1.5-cm pitch. A flow rectifier with 48 holes of 1-cm diameter was installed just above the inlet of liquid or slurry.

Air and tap water or slurry were used as the gas phase and the liquid phase, respectively. The slurry was a mixture of water and porous silica alumina particles (low-alumina cracking catalyst produced by Shokubai Kasei Kogyo, Co., Ltd., Japan; weight-mean diameter = 25–60 μm; apparent dry density = 0.9 $Mg\cdot m^{-3}$; apparent density soaked with water = 1.5 $Mg\cdot m^{-3}$). The viscosity of the slurry was measured with a coaxial cylindrical viscometer. The slurry of $\bar{C}=0.4$ kg-soaked particles/kg-slurry showed the Bingham plastic property with a yield stress of 1.0 Pa. The concentration of fine particles in the slurry was constant in any part of the apparatus during steady-state operation. As the coarse solid phase, glass spheres and porous alumina particles were used. The properties of liquid and solid are listed in **Table 1**.

As shown in Fig. 1, five horizontal shutter plates[13] were installed at 52, 78, 103, 129 and 154 cm above the gas distributor in the 216-cm tall column and at 102, 128, 153, 173 and 204 cm in the 266-cm tall column. The hole size of each plate was the same as the column inside diameter. Airtightness of the shutter plate was ensured by a lubricated O-ring. After steady state was attained in the bed, the gas and liquid or slurry flows were stopped simultaneously. At that moment, all shutter plates were closed. The mean holdup of each phase between two shutter plates was obtained by measuring the volume of trapped gas and solid particles. The slurry was considered to be a homogeneous liquid.

Table 1. Properties of solid particles, liquid and slurry

Particle	ρ_p [$Mg\cdot m^{-3}$]	d_p [mm]	$\bar{C}$ [—]	η_l [mPa·s]	v_t [$cm\cdot s^{-1}$]
Glass sphere	2.52	0.52	0	1	7.8**
		2.2	0	1	29**
			0.4	10	13**
		5.2	0	1	50**
Porous alumina	1.8	1.5	0	1	14.5**
			0.2	2.5	9.9*
			0.4	10	4.8**

* Measured. ** Calculated.

2. Experimental Results and Discussion

2.1 Gas holdup

Figure 2 shows the mean gas holdup in the dense and lean regions. The values of gas holdup in the lean region are larger than those in the dense region but smaller than those in the bubble column,[7] whose gas distributor is a perforated plate with 2-mm holes. In the dense region, coalesced gas bubbles ascend at high velocities. At the upper surface of the dense region, however, these large bubbles are disintegrated by the intense turbulence.

The mean gas holdup in each region is expressed by the following equations.

For the lean region:

$$\bar{\varepsilon}_{gL}=0.3W^{1.3}/(1+1.1W^{1.15}) \tag{1}$$

For the dense region:

$$\bar{\varepsilon}_{gD}=\bar{\varepsilon}_{gL}/(1+0.01Re_t^{0.62}) \tag{2}$$

where

$$W=(gD_T^2\rho_l/\sigma)^{0.198}(gD_T^3/\nu_l^2)^{0.035}(U_g/\sqrt{gD_T})$$

The dimensionless groups and their powers in the above equations were determined on the basis of dimensional analysis and the dependency of gas holdup on experimental variables. It was also assumed

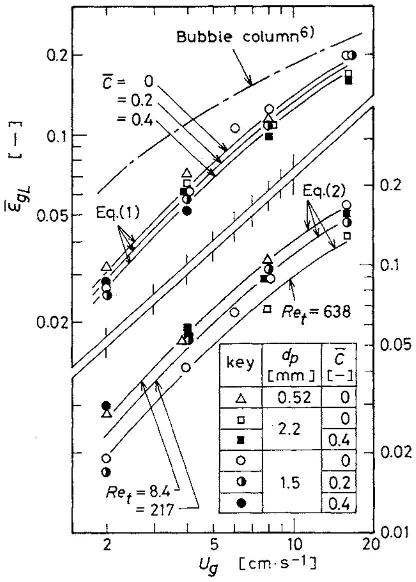

Fig. 2. Mean gas holdup in dense and lean regions.

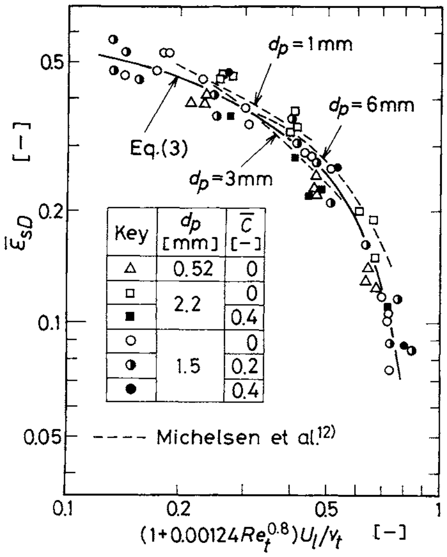

Fig. 4. Correlation of mean solid holdup in dense region.

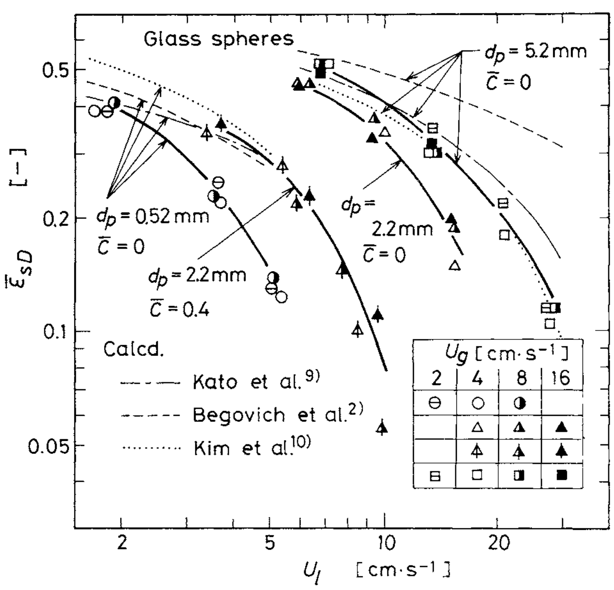

Fig. 3. Mean solid holdup in dense region.

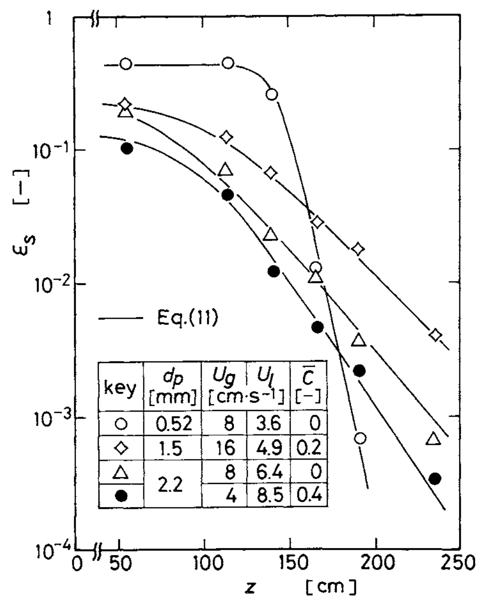

Fig. 5. Axial distribution of solid holdup.

that $\bar{\varepsilon}_{gL}$ was proportional to the -0.07 power of liquid viscosity[6] and was independent of column diameter.[1,9]

2.2 Mean solid holdup in dense region

Figure 3 shows typical data of mean solid holdup in the dense region. The values of $\bar{\varepsilon}_{sD}$ decrease with increasing liquid velocity and slurry concentration and with decreasing particle diameter. No effect of gas velocity on $\bar{\varepsilon}_{sD}$ was observed within experimental error. Calculated values from the experimental equations of Kim *et al.*,[10] Begovich and Watson,[2] and Kato *et al.*[9] are shown in Fig. 3. Agreement of the calculated values from the equations of Kato *et al.*[9] with the experimental ones is the best. However, agreement becomes poor for solid holdup smaller than 0.3, because their equations were originally derived to correlate the holdups of gas and liquid, respectively. As shown in **Fig. 4**, the data of the present work and that of Michelsen and Østergaard[12] are well correlated by the following equation.

$$\bar{\varepsilon}_{sD}=\bar{\varepsilon}_{s0}[1-1.09\{(1+0.00124Re_t^{0.8})U_l/v_t\}^{0.85}] \quad (3)$$

where $\bar{\varepsilon}_{s0}$ is solid holdup in the completely settled bed and is 0.63 for spherical particles of the same size. Equation (3) is valid under the following conditions: $Re_t=8$–3000, $U_l/v_t=0.08$–0.7, and $\bar{\varepsilon}_{sD}>0.15$.

2.3 Behavior of solid particles in lean region

Axial distribution of solid particles **Figure 5** shows sample data for axial distribution of solid holdup. In the lean region, solid holdup decreases exponentially with increasing height and is described by the sedimentation diffusion model.[8,15]

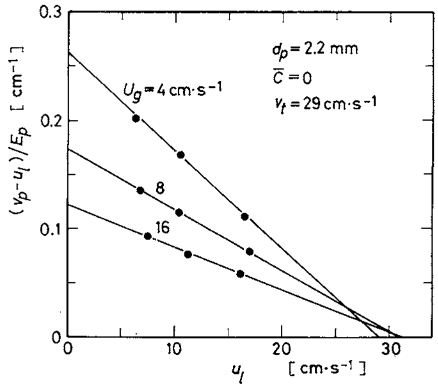

Fig. 6. Calculation of v_p and E_p.

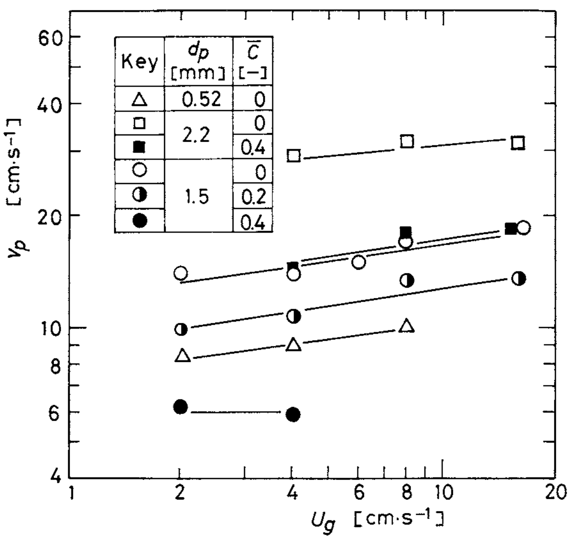

Fig. 7. Mean settling velocity of solid particles in lean region.

$$E_p\frac{d^2\varepsilon_{sL}}{dz^2}+(v_p-u_l)\frac{d\varepsilon_{sL}}{dz}=0 \tag{4}$$

where E_p is the longitudinal dispersion coefficient of solid particles, v_p the mean settling velocity and u_l the actual liquid velocity. When the column is sufficiently long, the solution of Eq. (4) is given as

$$\varepsilon_{sL}/\bar{\varepsilon}_{sD}=\exp\{-(v_p-u_l)(z-L_f)/E_p\} \tag{5}$$

where L_f is the asymptotic height of the dense region. Asymptotic solid holdup at L_f is given by $\bar{\varepsilon}_{sD}$. The relationship between $(v_p-u_l)/E_p$ and u_l is shown in **Fig. 6**. By assuming that v_p and E_p are independent of u_l, they can be obtained from the intercepts on the abscissa and the ordinate.

Mean settling velocity **Figure 7** shows the mean settling velocity of solid particles in the lean region. The values of v_p are nearly proportional to the terminal velocity of the particle and increase slightly with increasing gas velocity. Kato *et al.*[8] measured the mean settling velocity in bubble columns with suspended solid particles of glass spheres, whose mean diameters were 74, 97, 137 and 162 μm. Their

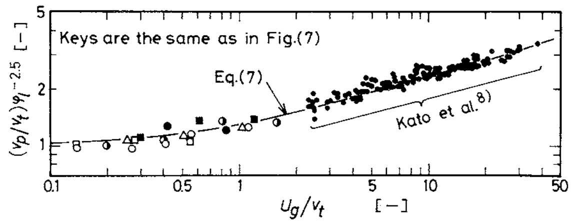

Fig. 8. Correlation of mean settling velocity of solid particles for three-phase systems.

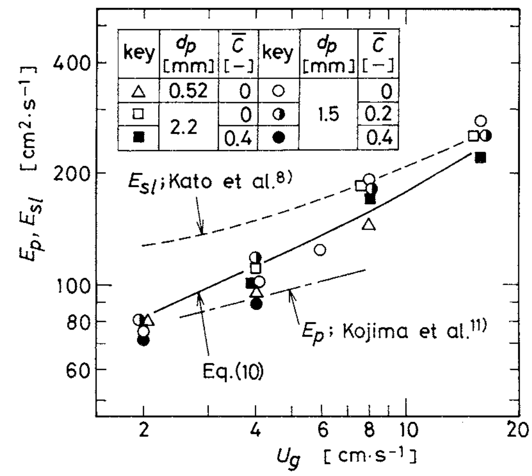

Fig. 9. Longitudinal dispersion coefficient of solid particles in lean region.

data were correlated as

$$v_p=1.33v_t(U_g/v_t)^{0.25}\psi_l^{2.5} \tag{6}$$

where

$$\psi_l=\varepsilon_l/(\varepsilon_l+\varepsilon_s)\ .$$

As shown in **Fig. 8**, the values of v_p in the lean region of the three-phase fluidized bed are linked smoothly to those in the bubble column with suspended solid particles. Thus, v_p is correlated in the following equation in the range of $U_g/v_t<30$.

$$v_p=v_t\{1+1.5(U_g/v_t)\}^{0.30}\psi_l^{2.5} \tag{7}$$

Longitudinal dispersion coefficient of solid particles **Figure 9** shows the longitudinal dispersion coefficient of solid particles in the lean region. The values of E_p are independent of particle size and slurry concentration. The broken line in Fig. 9 is the longitudinal dispersion coefficient of liquid in bubble columns with suspended solid particles[8] and is given by:

$$U_gD_T/E_{sl}=13Fr/(1+8Fr^{0.85}) \tag{8}$$

The experimental values of E_p are lower than those of E_{sl} and are larger than those of E_p given by Kojima and Asano.[11] Kato *et al.*[8] correlated the longitudinal dispersion coefficient of solid particles in three-phase bubble columns as follows.

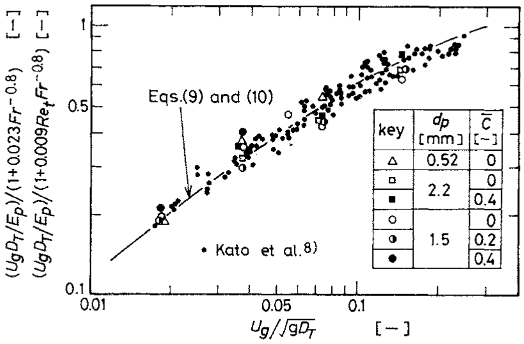

Fig. 10. Correlation of longitudinal dispersion coefficient of solid particles for three-phase systems.

In the range of $Re_t = 0.3$–2.5:

$$U_gD_T/E_p = \{13Fr/(1+8Fr^{0.85})\}(1+0.009Re_tFr^{-0.8}) \tag{9}$$

Comparison of Eq. (9) with the data of E_p in the lean region of the three-phase fluidized bed indicates that the correction factor in Eq. (9) should be replaced by $(1+0.023Fr^{-0.8})$. Thus, in the range of $Re_t = 2.5$–640, the following equation can be obtained.

$$U_gD_T/E_p = \{13Fr/(1+8Fr^{0.85})\}(1+0.023Fr^{-0.8}) \tag{10}$$

All values of E_p in the range of $Re_t = 0.3$–640 are plotted in **Fig. 10**. The solid line in Fig. 10 is the calculated value from Eqs. (9) and (10) and is in good agreement with the experimental ones. The solid lines in Fig. 5 indicate calculated values from the following equations, Eq. (11), with Eqs. (1), (2), (3), (7), (9) and (10).

$$\varepsilon_s = \bar{\varepsilon}_{sD}/[1+\exp\{(v_p-u_l)(z-L_f)/E_p\}] \tag{11}$$

Equation (11) is an experimental one which can be applied over the length of the bed. In this case, the operation is batchwise with respect to solid particles.

Conclusion

The gas and solid holdups in a so-called three-phase fluidized bed was investigated. The mean gas holdup in the dense and lean regions was correlated by Eqs. (1) and (2), respectively. The mean solid holdup in the dense region was expressed by Eq. (3), which also correlated the data of Michelsen and Østergaard.

The axial distribution of solid particles in the lean region was described by the sedimentation diffusion model. The mean settling velocity and the longitudinal dispersion coefficient of solid particles in the lean region of the three-phase fluidized bed were connected with those of bubble columns with suspended solid particles. Over a wide range of experimental conditions, the mean settling velocity and the longitudinal dispersion coefficient of solid particles were correlated by Eq. (7) and Eqs. (9) and (10), respectively.

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Nomenclature

$\bar{C}$	=	concentration of solid particles in slurry	[kg-soaked particles/kg-slurry]
D_T	=	diameter of bed	[m]
d_p	=	diameter of coarse particles	[m]
E_p	=	longitudinal dispersion coefficient of solid particles in the lean region of three-phase fluidized bed or in bubble column with suspended solid particles	[m^2·s^{-1}]
E_{sl}	=	longitudinal dispersion coefficient of liquid in bubble column with suspended solid particles	[m^2·s^{-1}]
Fr	=	Froude number, $U_g/\sqrt{gD_T}$	[—]
g	=	acceleration of gravity	[m·s^{-1}]
L_f	=	height of dense region	[m]
Re_t	=	Reynolds number, $v_td_p\rho_l/\eta_l$	[—]
U_g	=	superficial gas velocity	[m·s^{-1}]
U_l	=	superficial liquid velocity	[m·s^{-1}]
u_l	=	actual liquid velocity, $U_l/(1-\bar{\varepsilon}_g)$	[m·s^{-1}]
v_p	=	mean settling velocity	[m·s^{-1}]
v_t	=	terminal velocity of a particle	[m·s^{-1}]
z	=	axial coordinate	[m]
$\bar{\varepsilon}_{gD}$	=	mean gas holdup in dense region	[—]
$\bar{\varepsilon}_{gL}$	=	mean gas holdup in lean region	[—]
ε_l	=	local liquid or slurry holdup	[—]
ε_s	=	local solid holdup	[—]
$\bar{\varepsilon}_{sD}$	=	mean solid holdup in dense region	[—]
ε_{sL}	=	local solid holdup in lean region	[—]
$\bar{\varepsilon}_{s0}$	=	mean solid holdup in completely settled bed	[—]
η_l	=	viscosity of liquid or slurry	[Pa·s]
ν_l	=	kinematic viscosity of liquid or slurry	[m^2·s^{-1}]
ρ_l	=	density of liquid or slurry	[kg·m^{-3}]
ρ_p	=	density of soaked solid particle	[kg·m^{-3}]
σ	=	interfacial tension	[N·m^{-1}]
ψ_l	=	$\varepsilon_l/(\varepsilon_l+\varepsilon_s)$	[—]

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LONGITUDINAL DISPERSION COEFFICIENT OF LIQUID IN THREE-PHASE FLUIDIZED BED FOR GAS-LIQUID-SOLID SYSTEMS

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Key Words: Chemical Reactor, Fluidization, Fluidized Bed, Three Phase Fluidization, Dispersion, Dispersion Coefficient, Mixing, Multiphase Flow, Peclet Number, Bubble Column

The longitudinal dispersion coefficient of liquid, E_l, was measured in three-phase fluidized beds 12 and 19 cm in diameter. The data of E_l were correlated by expanding the equation for liquid dispersion in the bubble column, and the following equations were obtained.

$$U_g < 9\,\mathrm{cm\cdot s^{-1}}: \quad U_g D_T/E_l = (1 + v_t/U_l)^{0.40 Bo^{0.225}} Pe(\mathrm{BC})$$

$$U_g > 25\,\mathrm{cm\cdot s^{-1}}: \quad U_g D_T/E_l = Pe(\mathrm{BC})$$

In the range of $U_g = 9$–$25\,\mathrm{cm\cdot s^{-1}}$, E_l was approximated by connecting the above equations. The notations are:

$$Pe(\mathrm{BC}) = (\eta_l/\eta_w)^{0.07}[13(U_g/\sqrt{gD_T})/\{1 + 6.5(U_g/\sqrt{gD_T})^{0.8}\}]$$

$Bo = gd_p^2\rho_l/\sigma$, D_T = diameter of bed, d_p = diameter of solid particles, g = acceleration due to gravity, U_g = superficial gas velocity, U_l = superficial liquid velocity, v_t = terminal velocity of solid particle, η_l = viscosity of liquid, η_w = viscosity of water, ρ_l = liquid density and σ = surface tension.

Introduction

In the three-phase fluidized bed, solid particles whose terminal velocity exceeds several centimeters per second are fluidized mainly by the liquid flow.[2] Meanwhile, solid particles whose terminal velocity is lower than several centimeters per second are suspended mainly by the gas flow.[2] This system is called a bubble column with suspended solid particles.

The longitudinal dispersion coefficient of liquid, E_l, in the three-phase fluidized bed has been studied by several investigators.[1,7–9,11,13] Owing to the complexity of three-phase flow, however, only a few correlations of E_l have been proposed to date.

Muroyama *et al.*,[11] measured the liquid dispersion in three-phase fluidized beds of 6 and 10 cm diameter, and correlated the data for dispersed bubble flow and coalesced bubble flow by separate equations. However, the boundary of each flow regime was not specified, and so no correlation was given for the intermediate flow regime between dispersed bubble flow and coalesced bubble flow.

El-Temtamy *et al.*[1] used a three-phase fluidized bed as small as 5 cm in diameter. Their correlation does not account for the effect of gas velocity, through the value of E_l increases with increasing gas velocity drastically in the intermediate flow region.

Recently, Kim and Kim[8] proposed a correlation based on the isotropic turbulence theory. According to their equation, E_l is proportional to the -0.66 power of particle diameter. Therefore, the estimated value of E_l in the three-phase fluidized bed cannot be linked to the observed value of E_l in the bubble column.

In this paper, the longitudinal dispersion coefficient of liquid is measured in three-phase fluidized beds of 12 and 19 cm diameter. The values of E_l are correlated

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