Korean J. Chem. Eng., 16(4), 456-461 (1999)

Gas Backmixing in the Dense Region of a Circulating Fluidized Bed

Won Namkung and Sang Done Kim†

Department of Chemical Engineering and Energy & Environment Research Center,
Korea Advanced Institute of Science and Technology, Taejon 305-701, Korea

(*Received 1 December 1998 • accepted 2 June 1999*)

Abstract–The gas backmixing characteristics in a circulating fluidized bed (0.1 m-ID×5.3-m high) have been determined. The gas backmixing coefficient (D_{ba}) from the axial dispersion model in a low velocity fluidization region increases with increasing gas velocity. The effect of gas velocity on D_{ba} in the bubbling bed is more pronounced compared to that in the Circulating Fluidized Bed (CFB). In the dense region of a CFB, the two-phase model is proposed to calculate D_{bc} from the two-phase model and mass transfer coefficient (k) between the crowd phase and dispersed phase. The gas backmixing coefficient and the mass transfer coefficient between the two phases increase with increasing the ratio of average particle to gas velocities (U_p/U_g).

Key words : Fluidized Bed, Gas Backmixing, Two-phase Model

INTRODUCTION

The flow behavior of gas-solid in a fluidized bed largely depends on the fluidization regime. Therefore, the gas mixing characteristics are also affected by the fluidization regime. Information of gas backmixing may be used to evaluate chemical conversion and selectivity at different gas flow rates [Arena, 1997] or gas flow regimes in a fluidized bed [Li and Weinstein, 1989; Namkung and Kim, 1997]. Many experiments have been carried out to determine gas mixing characteristics in a bubbling fluidization regime [van Deemter, 1985]. However, a comprehensive description of gas backmixing in thea high velocity fluidization regime has not yet been developed and controversy over some experimental results still remains. Furthermore, comparison studies of gas backmixing characteristics in the low- and high velocity regimes are relatively sparse.

It has been observed that a core-annulus structure exists in the dilute region of a CFB. Thus, several studies [Brereton et al., 1988; Werther et al., 1992; Amos et al., 1993; Namkung and Kim, 1997] have proposed the core-annulus models to determine the gas mixing coefficient in the dilute region of a CFB. Although the core-annulus structure in the dense region of CFB cannot be observed clearly as in case of the dilute region, annular flow is also observed in the dense region of a CFB [Ishii and Horio, 1991]. In the bubbling fluidization region, two- or three-phase models are used to determine the gas mixing coefficient [van Deemter, 1985]. Yerushalmi [1986] proposed a simple countercurrent model to predict gas backmixing in bubbling and turbulent fluidization regimes. In a CFB, gas backmixing in the center and wall regions of the dense phase has different mixing behavior [Li and Weinstein, 1989; Namkung and Kim, 1998a]. Since the two-phase model can account for different gas mixing behavior in the two regions [Arena, 1997], a more accurate gas backmixing coefficient may be determined in the dense region of a CFB.

In this study, the gas backmixing coefficients in the dense region from bubbling to fast fluidized beds are compared. A two-phase model for the dense region of a CFB is proposed to determine the gas backmixing coefficient (D_{bc}) and the mass transfer coefficient (k) between the crowd and the dispersed regions.

EXPERIMENTAL

The experiments were carried out in a transparent Plexiglas column (0.1 m-ID×5.3 m-high) as shown in Fig. 1. The details of the experimental facilities can be found elsewhere [Namkung and Kim, 1997, 1998a, b]. The solid particles used in the present study were FCC particles and their properties are given in Table 1. The superficial gas velocity (U_g) and the solids circulation rate (G_s) were varied in the range of 1.5-3.0 m/s and 14-45 kg/m^{2}s, respectively. The tracer gas (He) was injected at steady rate as step injection at 1.85 m above the distributor plate through an up-facing injection hole and sampled at four elevations (0.10, 0.25, 0.35, 0.55 m). The tracer injection rate was 0.45-0.56 % by volume of the total air flow rate. The holes of the sampling tube were covered with a screen to prevent leakage of solid particles from the bed. To determine the gas backmixing coefficient (D_{ba}, D_{bc}), the injection and sampling tubes were moved radially across the bed width. At steady state conditions, the tracer gas was sampled at nine different radial positions and the sampled gas was analyzed by gas chromatography (HP 5890II).

RESULTS AND DISCUSSION

The radial concentration distributions of backmixed tracer gas at the different tracer injection points in the dense regions of CFB are shown in Fig. 2. As can be seen in Fig. 2A, a very sharp radial concentration profile is observed with the tracer gas injection at the wall region. However, with the tracer gas injection at the center region, the amount of backmixed tracer gas is very low

†To whom correspondence should be addressed.
E-mail : kimsd@.kaist.ac.kr

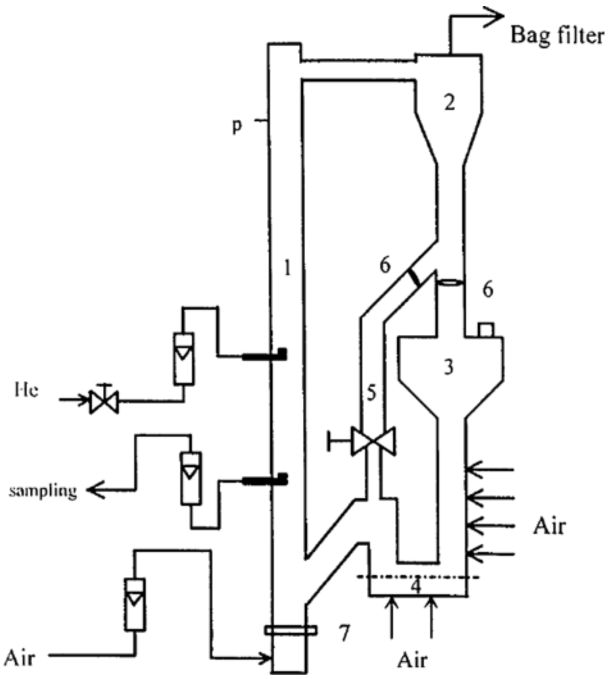

Fig. 1. Schematic diagram of apparatus.

1. Riser	5. Sampling pot
2. Cyclone	6. Butterfly valve
3. Hopper	7. Distributor
4. Loop-seal	

due to the high upward gas-solid flow in the center region of the bed. Therefore, it can be claimed that the gas backmixing characteristics in a center and wall region would be different. As can be seen, the tracer gas can reach the side opposite to the injection point due to the considerable circumferential gas mixing in the bed. In contrast to CFB, the very sharp radial concentration profiles do not exist and the intensity of gas backmixing is more or less uniform over the cross section of the bubbling fluidized bed [Li and Weinstein, 1989].

1. Backmixing Coefficients

1-1. Axial Dispersion Model

The gas backmixing coefficients in the dense regions have been determined by using the dispersion model based on the one-dimensional flow [Li and Weinstein, 1989; Li and Wu, 1991]. The axial dispersion model at steady state can be expressed as :

$$\varepsilon D_{ba}\frac{d^2C}{dx^2} - \varepsilon\frac{U_g}{\varepsilon}\frac{dC}{dx}=0 \tag{1}$$

Table 1. Physical properties of FCC particle

	FCC particle
Mean diameter [μm]	65
Apparent density [kg/m^3]	1720
ε_{mf}	0.494
U_{mf} [m/s]*	0.0027
U_t [m/s]	0.19
Transport velocity [m/s]**	1.4

*calculated by Haider and Levenspiel [1989]'s method
**emptying time method [Namkung et al., 1999]

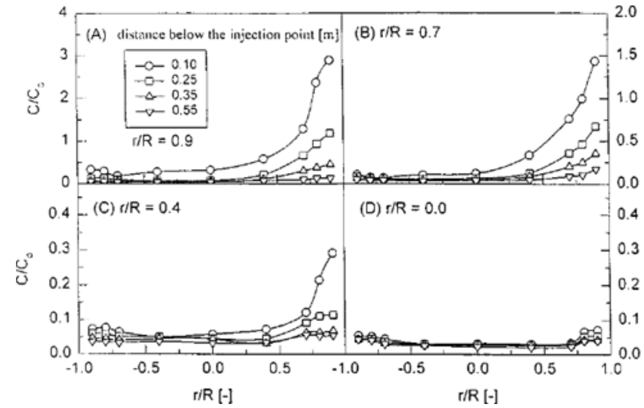

Fig. 2. Backmixed tracer gas concentration in the dense region (U_g=2.5 m/s).

with the following boundary conditions

$$x=0,\quad C=C_o,\quad x=-\infty,\quad C=0 \tag{2}$$

Then, Eq. (1) has a solution of the form

$$\frac{C}{C_0}=\exp\left(\frac{U_g}{\varepsilon D_{ba}}x\right) \tag{3}$$

where C_o and D_{ba} are the concentration of tracer gas at injection level and the gas backmixing coefficient, respectively. The average values of the sampled tracer concentrations along the radial direction (Fig. 2) at the different axial positions are used to calculate D_{ba} from the axial dispersion model.

The effect of gas velocity on the gas backmixing coefficient is shown in Fig. 3 with the data of previous studies (Table 2). Gas backmixing in a fluidized bed is mainly caused by downflow of solids. Nguyen et al. [1981] reported that gas backmixing commences in a bubbling fluidized bed at 3 U_{mf}. As can be seen, D_{ba} increases with increasing gas velocity at lower fluidizing gas velocities since movement of gas-solid and downflow of particles increase with increasing gas velocity.

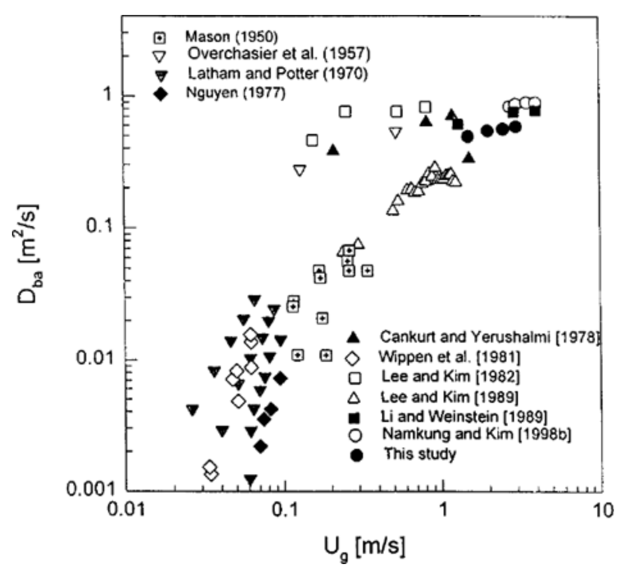

Fig. 3. Variation of gas backmixing coefficient with gas velocity.

Table 2. Summary on experimental conditions of previous studies (Fig. 4)

Workers	Particles	Size [μm]	Density [kg/m^3]	Column dia. [m]	Flow regime	Tracer gas
Mason [1950]	Catalyst	70	1956	0.0762	B	He
		104				
	Glass beads	102	2421			
		155				
Overchashier [1957]	Catalyst	77	1500	0.406	B	He
Latham and Potter [1970]	Glass	75	2400	0.152	B	Mercury vapor
		97	2410			
		125	2420			
Nguyen [1977]	Silica sand	145	2500	0.30	B	He
Cankurt and Yerushalmi [1978]	Catalyst	55	1074	0.152	S, T	CH_4
Wippen et al. [1981]	Sodium carbonite	100	1450	0.19	B	CO_2
Lee and Kim [1982]	Cement raw mill	23.6	2500	0.078	S, T	CO_2
Lee and Kim [1989]	Glass beads	362	2500	0.10	S, T	CO_2
Li and Weinstein [1989]	Catalyst	59	1450	0.152	B, S, T	He
					F, P	
Namkung and Kim [1998b]	Silica sand	125	3055	0.10	F	CO_2
This study	FCC particles	65	1720	0.10	F	He

Flow regime : B, S, T, F, P-bubbling, slugging, turbulent, fast and pneumatic, respectively.

Cankurt and Yerushalmi [1979] reported that gas backmixing diminishes over the turbulent flow regime, and they have defined the gas velocity at which D_{ba} exhibits its maximum value as the transition velocity from bubbling to turbulent fluidized bed. In the fast fluidized bed, D_{ba} increases with increasing G_s but decreases with increasing U_g [Namkung and Kim, 1998a]. To maintain a dense region with increasing U_g, G_s should increase accordingly. Thus, it is difficult to observe the effect of U_g on D_{ba} in the dense region of CFB independently. However, D_{ba} in the fast fluidized bed are higher than those in a turbulent fluidized bed since the increasing rate of D_{ba} decreases as the gas velocity increases [Namkung and Kim, 1998a, b; Li and Weinstein, 1989]. As can be seen in Fig. 3, the effect of U_g on D_{ba} in the bubbling region is more pronounced than that in the turbulent and fast fluidized bed regimes.

The effect of U_g on Peclet number ($Pe_b = U_g D_t / D_{ba}$) is shown in Fig. 4. As can be seen, Pe_b decreases significantly with increasing U_g. However, as U_g is increased above 0.5 m/s, Pe_b values lie in the range of 0.2-0.6 regardless of U_g. In the circulating fluidized bed, the effects of convection and dispersion of gas phase on Pe_b become similar with increasing U_g.

1-2. Two-phase Model in the Dense Region of CFB

As shown in Fig. 2, in the dense region of the CFB, gas backmixing in the center and wall regions of the bed would be somewhat different. To evaluate D_{bc} in the bubbling fluidized bed, van Deemter [1961] proposed a two-phase model. Later, Lee and Kim [1982] adopted this model in the turbulent fluidized bed. However, the proposed model based on the two-phase structure in the dense region of CFB to calculate gas backmixing coefficient is scarce. Therefore, in the present study, the two-phase model in the dense region of CFB is proposed. Although the two regions in the dense region are not observed clearly as in case of the dilute region, the dense phase can be divided into the crowd region in which gas backmixing occurs and the dispersed region where gas flows upward.

The following assumptions are made to simplify the proposed model as shown in Fig. 5. Gas backmixing occurs in the crowd region due to downflow of solids, but it is neglected in the dispersed region. The crowd region exists at the condition of minimum fluidization state. Gas exchange occurs between the crowd and dispersed regions, and the mass transfer coefficient (k) and D_{bc} do not change with the bed height.

Namkung and Kim [1997, 1998a] proposed the core-annulus model to calculate D_{bc} and k in the dilute region of a CFB. For the crowd and dispersed regions at steady state, the following

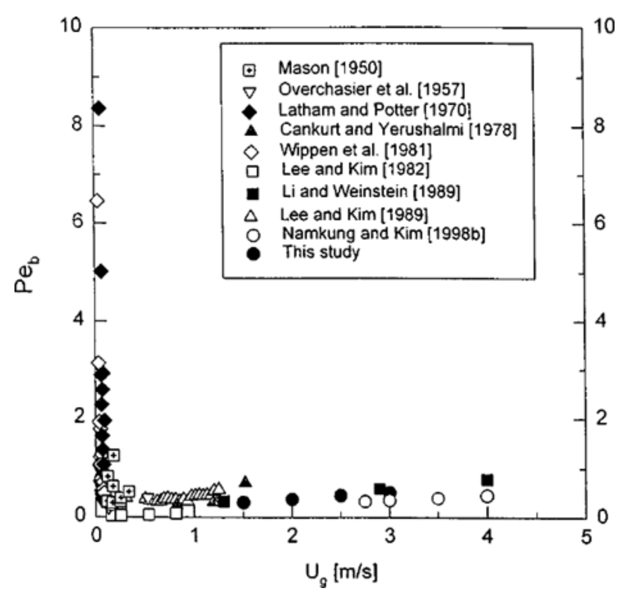

Fig. 4. Variation of Peclet number with gas velocity.

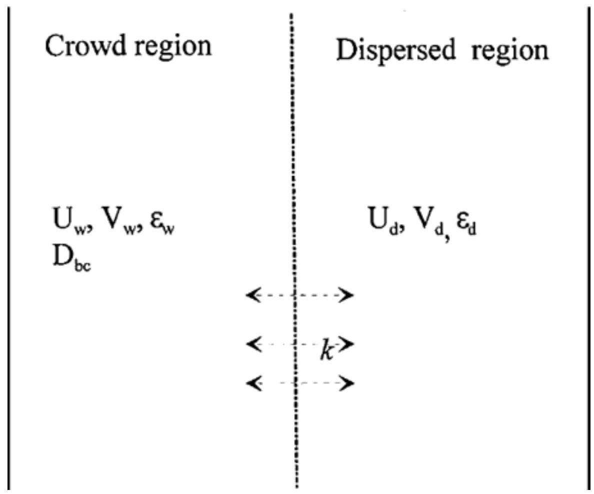

Fig. 5. Two-phase model for gas backmixing in the dense region of the CFB.

can be derived as :

$$\text{dispersed region}: \varepsilon_d U_d \frac{dC_d}{dx} + k(C_d - C_w) = 0 \tag{5}$$

$$\text{crowd region}: \varepsilon_w U_w \frac{dC_w}{dx} - D_{bc}\varepsilon_w \frac{d^2 C_w}{dx^2} + k(C_w - C_d) = 0 \tag{6}$$

If the axial distance (x) is far below the tracer injection point, the tracer concentrations in the crowd and dispersed regions approach zero. At the tracer injection plane, the sum of tracer gas concentration in the crowd (C_w) and the dispersed regions (C_d) equals the concentration of tracer gas at injection level (C_o). Then, the boundary conditions are as follows :

$$\text{B.C.: } x = -\infty, \quad C_w = C_d = 0 \tag{7}$$

$$x = 0, \quad \left(\frac{V_w}{V_T}\right)\frac{C_w}{C_o} + \left(\frac{V_d}{V_T}\right)\frac{C_d}{C_o} = 1 \tag{8}$$

where, x=0 is the tracer injection point and (V_w/V_T) is the ratio of the volume of crowd region to that of the total.

Eqs. (5) and (6) are solved by using the boundary conditions of Eqs. (7) and (8). The obtained solutions are :

$$\frac{C_w}{C_o} = \frac{(k + \varepsilon_d U_d r_1)\exp(r_1 x)}{k + \varepsilon_d U_d r_1 \left(\frac{V_w}{V_T}\right)} \tag{9}$$

$$\frac{C_d}{C_o} = \frac{k \exp(r_1 x)}{k + \varepsilon_d U_d r_1 \left(\frac{V_w}{V_T}\right)} \tag{10}$$

where

$$r_1 = \frac{\frac{U_w}{D_{bc}} - \frac{k}{\varepsilon_d U_d} + \sqrt{\left(\frac{U_w}{D_{bc}}\right)^2 + \frac{2kU_w}{\varepsilon_d U_d D_{bc}} + \left(\frac{k}{\varepsilon_d U_d}\right)^2 + \frac{4k}{\varepsilon_w D_{bc}}}}{2} \tag{11}$$

The values of (V_w/V_T), ε_d, C_w and C_d are needed to determine k and D_{bc}. However, it is impossible to measure exact values of ε_d, C_w and C_d. Therefore, ε_d value is calculated from the proposed correlation of Zhang et al. [1991] at r=0 ($\varepsilon=\varepsilon^{0.191}$). As can be seen in Fig. 2, the tracer concentration at the center region

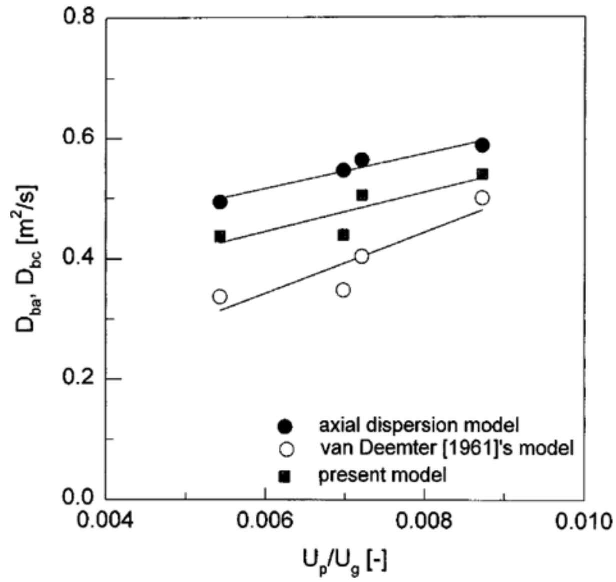

Fig. 6. Effect of U_p/U_g on D_b in the dense region of the CFB.

($|r/R| \le 0.55$) is low and nearly constant irrespective of the radial distance. Therefore, the values of C_w ($|r/R| > 0.55$) and C_d are used as the tracer gas concentration at the wall and center regions of the bed, respectively. If ε_d and ε are known, then (V_w/V_T) can be determined from the following equation.

$$\frac{V_w}{V_T} = \frac{\varepsilon - \varepsilon_d}{\varepsilon_w - \varepsilon_d} \tag{12}$$

To obtain the analytical solutions of D_{bc} and k from Eqs. (9) and (10), the mean values of C_w and C_d along the radial direction at each axial position are calculated from the experimental data. Next, using the boundary conditions of C_w/C_o and C_d/C_o at x=0 (Eq. (8)), the gradient (r_1) of C_w/C_o and C_d/C_o versus the axial distance (x) in the semi-log plot is adjusted until the gradient (r_1) becomes identical. Then, D_{bc} and k can be calculated from the gradient and intersection of the profile of C_w/C_o and C_d/C_o versus x in the semi-log plot.

The calculated values of the gas backmixing coefficients (D_{ba}, D_{bc}) from the axial dispersion and the two-phase models are shown in Fig. 6. As can be seen, the gas backmixing coefficient increases with increasing U_p/U_g since downflow of solids near the wall region is more pronounced with increasing U_p/U_g. The obtained D_{bc} values from the two-phase model are lower than those from the axial dispersion model. This difference may come from the assumption of gas backmixing in the dispersed region being absent in the two-phase model. However, a little amount of backmixed tracer gas in the dispersed region may exist from mass transfer between the crowd and dispersed regions and the local downflow of solids in the dispersed region. The amount of backmixed tracer gas in the dispersed region of the dense region is higher than that of the backmixed tracer gas in the core region of the dilute region. Therefore, the difference in gas backmixing coefficients (D_{ba}, D_{bc}) from the axial dispersion and the two-phase models in the dense region is higher than those in the dilute region [Namkung and Kim, 1998a]. As can be seen, the calculated values of D_{bc} from the model of van Deemter [1961] in the bubbling bed have the lowest values. Owing to the different gas backmixing characteristics of the crowd and dispersed phases in the dense region (Fig. 2), it can be claimed that the obtained gas backmixing coefficient from the two-phase model better describes gas flow patterns in the bed compared to the axial dispersion model.

The mass transfer coefficient (k) increases with increasing U_p/U_g because the upward gas velocity and downflow of solids increase with increasing U_p/U_g. As expected, k values in the dense region (6.535-15.215 m/s) are much higher than those in the dilute region (0.109-0.277 m/s) of CFB [Namkung and Kim, 1998a].

CONCLUSIONS

The gas backmixing coefficient (D_{ba}) in the lower fluidizing gas velocity ranges increases with increasing gas velocity. The Peclet number (Pe_b) decreases significantly with increasing U_g in the lower fluidizing gas velocities; however, as U_g is increased above 0.5 m/s, Pe_b is not affected largely by the increasing of U_g. In the dense region of the CFB, a considerable amount of backmixed tracer gas near the wall region of the bed is observed. However, at the center region of the bed, gas backmixing is negligibly small. The two-phase model in the dense region of CFB is proposed to calculate D_{bc} and mass transfer coefficient (k) between crowd and dispersed regions. The gas backmixing coefficient and the mass transfer coefficient between two-phase increase with increasing the ratio of average particle to gas velocities (U_p/U_g).

NOMENCLATURE

Ar : Archimedes number [-]
C : concentration of tracer gas [mol/m^3]
C_o : concentration of tracer gas at injection level, mean mixed vol%, tracer gas volumetric flow rate/U_g multiplied by the bed area [mol/m^3]
D_{ba} : gas backmixing coefficient from the axial dispersion model [m^2/s]
D_{bc} : gas backmixing coefficient from the two-phase model [m^2/s]
D_t : column diameter [m]
G_s : solids circulation rate [kg/m^{2}s]
k : mass transfer coefficients between crowd and dispersed region [m/s]
Pe_b : Peclet number ($U_g D_t/D_{ba}$)
Re : Reynolds number ($U_g d_p \rho_g/\mu$) [-]
U_g : superficial gas velocity [m/s]
U_p : particle velocity (G_s/ρ_s) [m/s]
U_t : particle terminal velocity [m/s]
x : vertical distance from tracer gas injection level [m]
z : distance from the distributor [m]

Greek Letters

ε : gas volume fraction [-]
ε_{mf} : gas volume fraction in the minimum fluidization conditions [-]
ρ_s : particle density [kg/m^3]

Subscripts

c : crowd region
d : dispersed region

REFERENCES

Amos, G., Rhodes, M. J. and Mineo, H., "Gas Mixing in Gas-Solid Risers," *Chem. Eng. Sci.*, **48**, 943 (1993).
Arena, U., "Circulating Fluidized Bed," Editors J. R. Grace, A. A. Avidan and T. M. Knowlton, Blackie A & P, London, chap 3 (1997).
Brereton, C. M. H., Grace, J. R. and Yu, J., "Axial Gas Mixing in a CFB," CFB Technol. II, Editors P. Basu and J. F. Large, Pergamon Press, New York, 307 (1988).
Cankurt, N. T. and Yerushalmi, J., "Gas Backmixing in High Velocity Fluidized Beds," Fluidization, Editors J. F. Davidson and D. L. Kearins, Cambridge Univ. Press, London, 387 (1978).
Haider, A. and Levenspiel, O., "Drag Coefficient and Terminal Velocity of Spherical and Nonspherical Particles," *Powder Technol.*, **58**, 63 (1989).
Ishii, H. and Horio, M., "The Flow Structure of a Circulating Fluidized Bed," *Adv. Powder Technol.*, **2**, 25 (1991).
Latham, R. and Potter, O. E., "Backmixing of Gas in a 6-in Diameter Fluidized Bed," *Chem. Eng. J.*, **1**, 152 (1970).
Lee, J. S. and Kim, S. D., "The Vertical Pneumatic Transport of Cement Raw Meal," *HWAHAK KONGHAK*, **20**, 207 (1982).
Lee, G. S. and Kim, S. D., "Gas Mixing in Slugging and Turbulent Fluidized Beds," *Chem. Eng. Comm.*, **86**, 91 (1989).
Li, Y. and Wu, P., "A Study on Axial Gas Mixing in a Fast Fluidized Bed," CFB Technol. III, Editors P. Basu, M. Horio, M. Hasatani, Pergamon Press, New York, 581 (1991).
Li, J. and Weinstein, H., "An Experimental Comparison of Gas Backmixing in Fluidized Beds Across the Regime Spectrum," *Chem. Eng. Sci.*, **44**, 1697 (1989).
Mason, E. A., Sc. D. Thesis, MIT, Cambridge, USA (1950) (cited in van Deemter (1961)).
Namkung, W. and Kim, S. D., "Gas Mixing in the Upper Dilute Region of a Circulating Fluidized Bed," CFB Technol. V, Editors M. Kwauk and J. Li, Science Press, Beijing, 134 (1997).
Namkung, W. and Kim, S. D., "Gas Backmixing in a Circulating Fluidized Bed," *Powder Technol.*, **99**(1), 70 (1998a).
Namkung, W. and Kim, S. D., "Gas Mixing Characteristics in a Fast Fluidized Bed," *HWAHAK KONGHAK*, **36**, 797 (1998b).
Namkung, W., Kim, S. W. and Kim, S. D., "Flow Regimes and Axial Pressure Profiles in a Circulating Fluidized Bed," *Chem. Eng. J.*, in press (1999).
Nguyen, N. V., Whitehead, A. B. and Potter, O. E., "Gas Backmixing Solid Movement, and Bubble Activities in Large Scale Fluidized Beds," *AIChE J.*, **23**, 913 (1977).
Nguyen, H. V., Potter, O. E., Dent, D. C. and Whitehead, A. B., "Gas Backmixing in Large Scale Fluidized Beds Containing Tube Assemblies," *AIChE J.*, **27**, 509 (1981).
Overchasier, R. H., Todd, D. B. and Olney, R. B., "Some Effects of Baffles on a Fluidized System," *AIChE J.*, **5**, 54 (1957).
van Deemter, J. J., "Mixing and Contacting in Gas-Solid Fluidized Beds," *Chem. Eng. Sci.*, **13**, 143 (1961).
Van Deemter, J. J., "Fluidization," Editors J. F. Davidson, D. Harrison and R. Clift, Academic Press, London, Chap. 7 (1985).

Werther, J., Hartge, E. U. and Kruse, M., "Radial Gas Mixing in the Upper Dilute Core of a Circulating Fluidized Bed," *Powder Technol.*, **70**, 293 (1992).

Wippen, D., Wittemann, K., Kuhne, J., Helmrich, H. and Schugerl, K., "Characterization of Fluidized Bed Reactors with Gas Tracer Measurements," *Chem. Eng. Comm.*, **10**, 307 (1981).

Yerushalmi, J., "Gas Mixing in High Velocity Fluidized Bed," Gas Fluidization Technology, Eds. D. Geldart, Wiley, Chichester, U.K., Chap. 7 (1986).

Zhang, W., Tung, Y. and Johnsson, F., "Radial Voidage Profiles in Fast Fluidized Beds of Different Diameters," *Chem. Eng. Sci.*, **46**, 3045 (1991).