RADIAL SOLID MIXING CHARACTERISTICS IN SHALLOW GAS FLUIDIZED BEDS

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Introduction

Radial solid mixing is known to be at least an order of magnitude slower than the mixing in the vertical direction, and for this reason, knowledge of radial solid mixing is more important than that of axial mixing in assessing the performance of large gas-solids fluidized bed processing units[9]. Therefore, the radial dispersion coefficients of solids (D_{sr}) in fluidized beds have been determined by numerous researchers[2, 3, 4, 6, 10, 11, 12, 16, 19, 20]. However, most of the studies have been carried out in deep ($H_{mf}/D_t > 1$) two-dimensional (2-D) fluidized beds whereas, the available data of D_{sr} in shallow three-dimensional (3-D) fluidized beds are relatively sparse. Therefore, in the present study, the effects of gas velocity (0.043-0.214 m/s), static bed height (0.15-0.30 m) and particle size (322, 440, 652 μm) on D_{sr} in a shallow ($0.6 < H_{mf}/D_t < 1.2$) 3-D fluidized bed have been determined. In addition, D_{sr} in terms of Peclet number (Pe_r) of the present and previous studies[3, 10, 12] have been correlated with the relevant dimensionless groups based on the isotropic turbulence theory as: $Pe_r = 1016.25\,[(U_g - U_{mf})^2/gD_t]^{1/3}\,(H_{mf}/D_t)$.

1. Experimental

1.1 Experimental procedure

Experiments were carried out in a 0.254 m-ID × 0.9 m-high Plexiglas column as shown in **Fig. 1**. Sand particles of three different mean diameter with a density of 2,600 kg/m^3 were used as the bed materials and glass beads (ρ_s = 2,500 kg/m^3) with different sizes were used as tracer particles, respectively (**Table 1**). Initially, a partition tube (0.040 m-O.D.) was inserted at the center of the bed and the tracer particles were filled up to a desired level in the tube. Sand was then charged into the bed on a bubble cap distributor plate up to the same level as that of the tracer particles in the central partition tube. When the system reached a steady state at a given gas velocity, the partition tube containing tracer particles was quickly removed and the tracer particles were dispersed in the bed. After dispersion of tracer particles, air supply to the bed was shut-off and the mixed particles were sampled from the different radial positions (r/R = 0.0, 0.28, 0.55, 0.83) in the bed by a sampling probe connected to a vacuum pump. A weighed sample consists of sand and tracer particles was screened to separate tracer particles and determined the concentration of tracer particles.

1.2 Data analysis

The one-dimensional dispersion model has been used to determine D_{sr} in a cylindrical gas-solid fluidized bed. The governing differential equation of the model and the initial and boundary conditions are the same as those of previous study[3], and the solution is[3]

$$\frac{C}{C_\infty} = 1 + \frac{2}{r_o}\sum_{i=0}^{\infty}\frac{J_1(\alpha_i r_0)\,J_0(\alpha_i r)}{\alpha_i\,[J_o(\alpha_i R)]^2}\exp\left(-D_{sr}\,\alpha_i^2\,t\right) \qquad (1)$$

The D_{sr} values were determined by the nonlinear regression of the measured tracer concentrations along the radial positions with the calculated profiles at a given operating time.

2. Results and Discussion

It is generally known that the radial mixing of solids in gas-solid fluidized beds is primarily caused by the movement of ascending gas bubbles through the bed. It was postulated that, as a bubble rises, solids around a bubble enter its cloud and are eventually drawn into the wake. Solids are completely mixed in the wake and this gives rise to the radial mixing[16]. Also, experimental evidence shows that bubbles bursting at the upper surface of the bed induces the radial dispersion of solids[5] which contributes to the overall radial mixing of solids[11]. The factors affecting the radial dispersion of solids at the bed surface are known to be the size and density of solid particles and density and viscosity of gas phase besides the size and velocity of bubbles[20]. The gross particle circulation may also greatly affect the radial solids mixing[18].

2.1 Effect of excess gas velocity (U_g-U_{mf})

The effect of (U_g-U_{mf}) on D_{sr} is shown in **Fig. 2** where D_{sr} increases with an increase in (U_g-U_{mf}) due to the increase of bubble size and its velocity since D_{sr} is known to increase with increasing bubble size and its rising velocity[11]. Also, the rate of increase in D_{sr} decreases with an increase in (U_g-U_{mf}) since bubble growth rate may decrease due to the increase of turbulent motion of solids.

Variation of D_{sr} of the present and previous stud-

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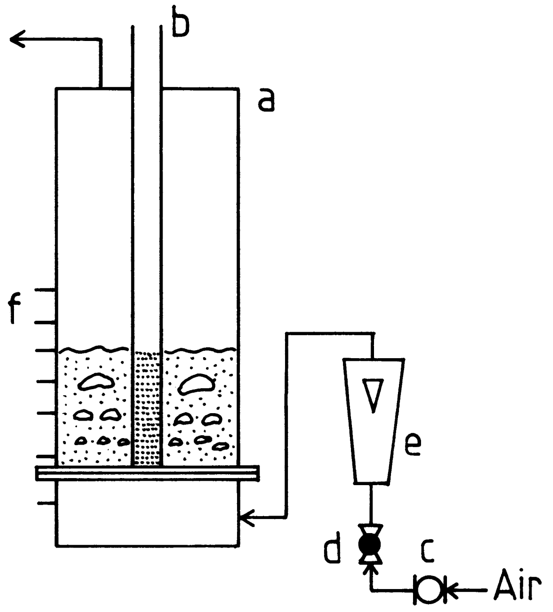

Fig. 1 Experimental apparatus
a. main column, b. partition tube, c. ball valve,
d. globe valve, e. rotameter, f. pressure taps

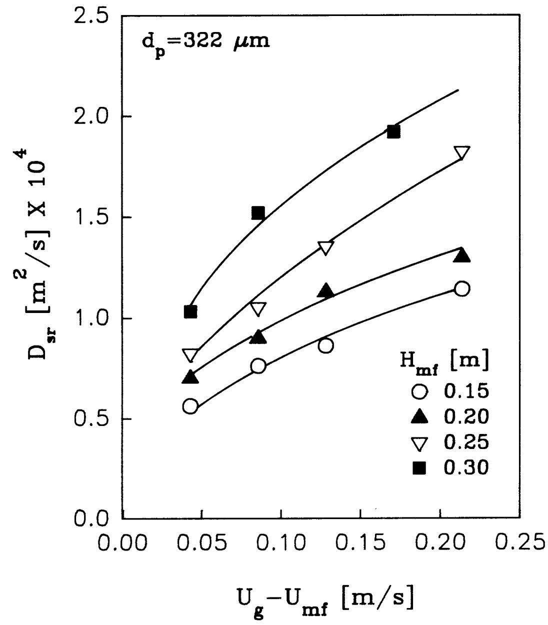

Fig. 2 Effect of excess gas velocity on the radial dispersion coefficient (D_{sr})

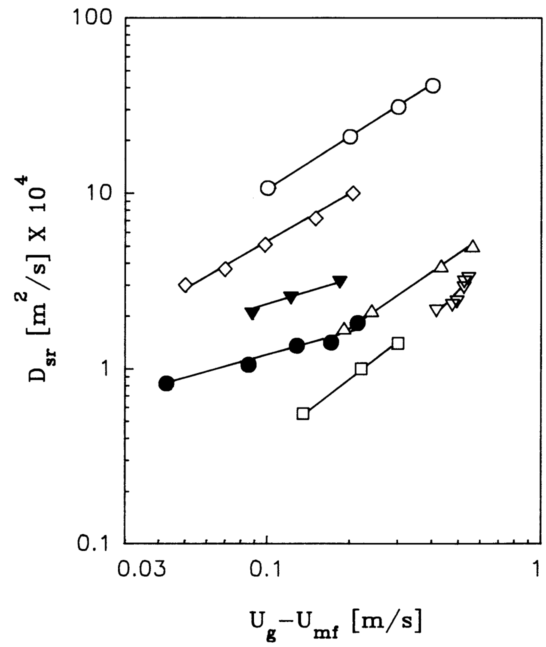

Fig. 3 Effect of excess gas velocity on D_{sr} in the present and previous studies
●, present study (3-D); ▼, Berruti *et al.*[3] (3-D);
□, Gabor[6] (2-D); ▽, Shi and Fan[20] (2-D)
◊, Hirama *et al.*[11] (2-D); △, Bellgardt and Werther[2] (2-D); ○, Mori and Nakamura[19] (2-D)

Table 1. Properties of Solid Particles

Mesh No. (K. S.)	Weight fraction (wt %) No. 1	No. 2	No. 3
−14 + 16			*
−16 + 25			0.326
−25 + 30		*	0.357
−30 + 35	*	0.203	0.277
−35 + 45	0.297	0.772	0.040
−45 + 60	0.644	0.025	
−60 + 70	0.059		
Mean Diameter [μm]	322	440	652

(*): sizes of the corresponding tracer particles

ies[3, 6, 10, 11, 20] with (U_g-U_{mf}) are shown in **Fig. 3**. As can be seen, the increasing rate of D_{sr} with (U_g-U_{mf}) are lower in 3-D beds than those in 2-D beds. Kunii and Levenspiel[16] claimed that the radial mixing of solids mainly occurs within the bubble wakes and the mean lateral displacement of solids is equal to the bubble radius. With the same increase of (U_g-U_{mf}), the mean lateral displacement of solids in 3-D beds is lower than that in 2-D beds since bubbles in 2-D beds are restrained by the walls and grow in the one dimensional lateral direction. From the Einstein random walk expression, it can be also noticed that D_{sr} for two-dimensional movement is lower than that for one-dimensional movement[15]. Moreover, the increasing rates of bubble frequency in 3-D beds are smaller than those in 2-D beds[7], which may reduce the increasing rate of D_{sr}.

2.2 Effect of static bed height (H_{mf})

The effect of H_{mf} on D_{sr} is shown in **Fig. 4** where D_{sr} increases with an increase in H_{mf} due to the increase of bubble size and rising velocity. However, in the present study, the increasing rate of D_{sr} with H_{mf} is about 0.7 which is lower than of Shi and Fan ($\cong$ 1.3)[20] since the data of the present study are obtained in a shallow fluidized bed where the radial mixing of solids may be strongly restricted by the grid surface[20].

2.3 Effect of particle size (d_p)

The effect of d_p on D_{sr} is shown in **Fig. 5** where D_{sr} decreases slightly with increasing d_p since the resistance to the particle movement increases due to the increase of inertia and gravitational forces with increasing particle size. However, the effect of d_p (322, 440, 652 μm) is less signif-

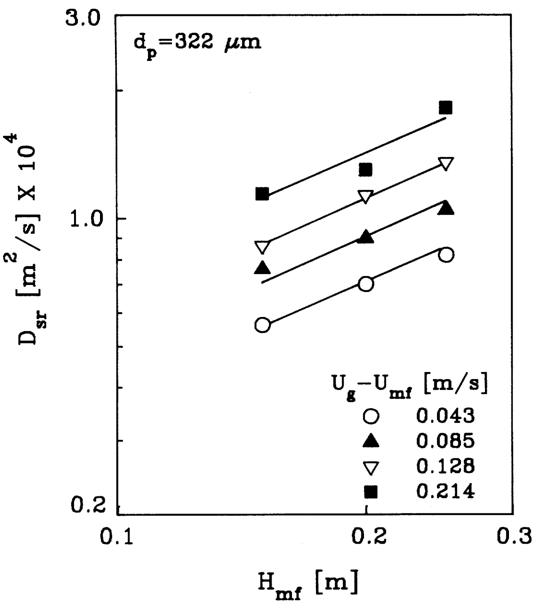

Fig. 4 Effect of static bed height on the radial dispersion coefficient (D_{sr})

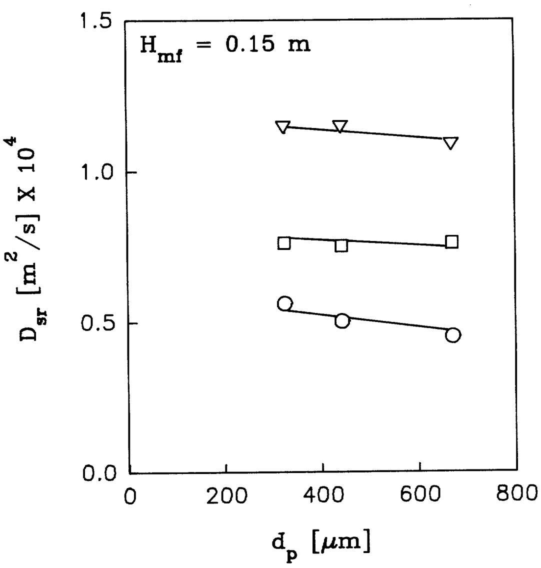

Fig. 5 Effect of particle size on the radial dispersion coefficient (D_{sr})
U_g-U_{mf} [m/s]: ○ 0.043, □ 0.085, ▽ 0.171

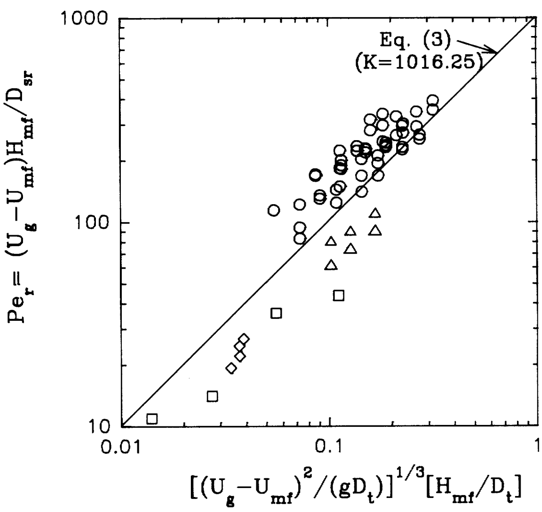

Fig. 6 Comparison of the calculated (Eq. 3) and measured Peclet numbers in the present and previous studies
○, present study; △, Berruti *et al.*[3]; □, Highley and Merrick[10] ◊, Horio *et al.*[12]

icant than those of excess gas velocity and static bed height on D_{sr} since the variation of bubble size which is one of the main factors to affect the lateral movement of solid particles is small with variation of the particle (Geldart B) size used in the present study[8].

3. Correlation

The axial dispersion coefficient (D_z) of solids in gas fluidized beds based on the isotropic turbulence theory has been proposed by Baird and Rice[1]:

$$D_z = KD_t^{4/3}\left[g\left(U_g - U_{mf}\right)\right]^{1/3} \tag{2}$$

where K is dimensionless constant.

Also, the correlations of the axial (D_z) and radial (D_r) dispersion coefficients of liquid phase in two- and three-phase fluidized beds[13, 14] and D_z of solids in turbulent fluidized beds[17] have been proposed based on the isotropic turbulence theory. In the present study, Peclet number is defined based on the static bed height (H_{mf})[3, 20] and the resulting equation can be written as:

$$Pe_r = \frac{\left(U_g - U_{mf}\right)H_{mf}}{D_{sr}} = \frac{\left(U_g - U_{mf}\right)H_{mf}}{KD_t^{4/3}\left[g\left(U_g - U_{mf}\right)\right]^{1/3}}$$

$$= \frac{1}{K}\left[\frac{\left(U_g - U_{mf}\right)^2}{gD_t}\right]^{1/3}\left(\frac{H_{mf}}{D_t}\right)$$

$$= K\left[\frac{\left(U_g - U_{mf}\right)^2}{gD_t}\right]^{1/3}\left(\frac{H_{mf}}{D_t}\right) \tag{3}$$

where dimensionless constant K value can be determined from a least squares fit of D_{sr} data obtained in the present and previous studies[3, 10, 12] from three-dimensional fluidized beds. The resulting value of K was 1016.25 with a correlation coefficient of 0.88. The range of variables covers $0.74 \times 10^{-3} < [(U_g - U_{mf})^2/gD_t] < 21.6 \times 10^{-3}$ and $0.2 < (H_{mf}/D_t) < 1.2$.

A comparison of measured and calculated (Eq. 3) Peclet numbers of the present and previous studies is shown in **Fig. 6**.

Conclusions

The radial dispersion coefficients of solids based on the one-dimensional dispersion model increases with increasing gas velocity and static bed height but, decreases slightly with an increase in particle size. The obtained radial dispersion coefficients in terms of Peclet number were correlated with the pertinent dimensionless groups based on the isotropic turbulence theory.

Nomenclature

C	=	tracer concentration	[-]

C_∞	=	equilibrium tracer concentration	[-]
d_p	=	particle diameter	[m]
D_{sr}	=	radial dispersion coefficient of solids	[m^2/s]
D_t	=	column diameter	[m]
D_z	=	axial dispersion coefficient	[m^2/s]
g	=	gravitational acceleration	[m/s^2]
h	=	height above the distributor	[m]
H_{mf}	=	static bed height	[m]
J_o, J_1	=	Bessel functions	
K	=	dimensionless constant	[-]
Pe_r	=	Peclet number based on static bed height	[-]
r	=	radial position in the reactor	[m]
r_o	=	radius of the partition tube	[m]
R	=	column radius	[m]
t	=	time after tracer injection	[s]
U_g	=	superficial gas velocity	[m/s]
U_{mf}	=	minimum fluidization velocity	[m/s]
α_i	=	constant such that $\alpha_i R$ is the ith zero of the Bessel function J_i	
μ	=	fluid viscosity	[Pa s]
ρ_g	=	fluid density	[kg/m^3]
ρ_s	=	solid density	[kg/m^3]

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