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Radial Gas Mixing Characteristics in a Downer Reactor

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Abstract–In a downer reactor (0.1 m-I.D.×3.5 m-high), the effects of gas velocity (1.6-4.5 m/s), solids circulation rate (0-40 kg/m^{2}s) and particle size (84, 164 μm) on the gas mixing coefficient have been determined. The radial dispersion coefficient (D_r) decreases and the radial Peclet number (Pe_r) increases as gas velocity increases. At lower gas velocities, D_r in the bed of particles is lower than that of gas flow only, but the reverse trend is observed at higher gas velocities. Gas mixing in the reactor of smaller particle size varies significantly with gas velocity, whereas gas mixing varies smoothly in the reactor of larger particle size. At lower gas velocities, D_r increases with increasing solids circulation rate (G_s), however, D_r decreases with increasing G_s at higher gas velocities. Based on the obtained D_r values, the downer reactor is found to be a good gas-solids contacting reactor having good radial gas mixing.

Key words : Fluidization, Downer, Circulating Fluidized Bed, Radial Gas Mixing, Dispersion Model

INTRODUCTION

It has been known that circulating fluidized bed (riser) reactors offer significant advantages over conventional bubbling fluidized bed reactors [Cho et al., 1996; Han et al., 1985]. However, the riser has non-uniform flow structure such as the core-annulus flow structure, and less product selectivity [Wang et al., 1992; Zhu et al., 1995]. These disadvantages of the riser reactor may be caused by flow of gas and solids against gravity. Therefore, a down-flow circulating fluidized bed reactor (simply downer reactor) has been devised recently. In a downer reactor, the axial solids dispersion and non-uniformity of radial gas and solids flow can be significantly reduced by the same directions of gas and solid flow for gravity [Bai et al., 1992; Wei et al., 1995]. Although a comprehensive understanding of the gas mixing characteristics in a downer reactor is very important for design purposes, not enough studies of gas mixing in downer reactors have been done compared with those in riser reactors [Wei et al., 1995]. Moreover, the obtained gas mixing data in the riser reactors in the literature vary greatly due to the difference in particle properties, reactor geometry, and operating variables [Amos et al., 1993]. One study on the gas mixing characteristics in a downer reactor (0.14 m-I.D.) is reported by Wei et al. [1995]. Certainly, further studies are needed on the gas mixing characteristics in downer reactors to provide basic design data for downer reactors. Therefore, the objective of this study is to determine the effects of gas velocity (U_g=1.6-4.5 m/s), solids circulation rate (G_s=0-40 kg/m^{2}s) and particle size (d_p=84, 164 μm) on the gas mixing coefficient in a downer reactor which was determined by using a dispersion model.

EXPERIMENTAL

Experiments were carried out in a downer (0.1 m I.D.×3.5 m high) and a riser (0.078 m I.D.×7.5 m high) as shown schematically in Fig. 1. It consists of a riser, two cyclones, fluidized bed feeder, downer, gas-solid separator and loop-seal. The bed materials used were silica sand with a particle density of 3,120 kg/m^3 and mean diameter of 84, 164 μm. The gas velocity and the solids circulation rate were varied in the range of 1.6-4.5 m/

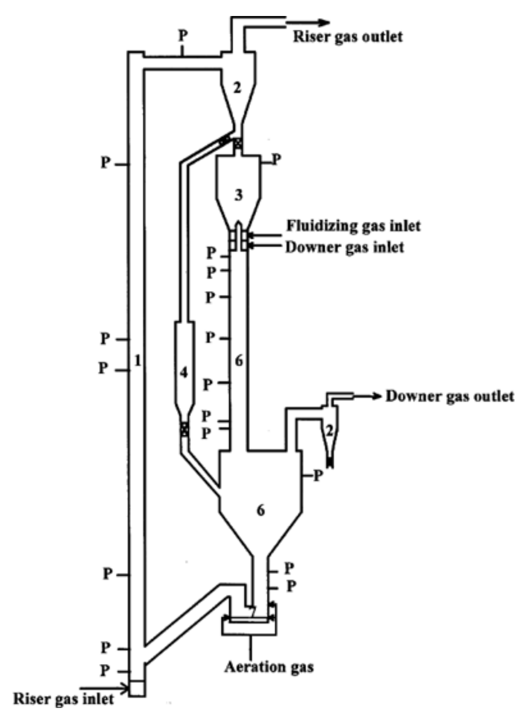

Fig. 1. Schematic flow diagram of the experimental apparatus.

1. Upflow riser	5. Downer
2. Cyclone	6. Separator and solid hopper
3. Fluidized bed feeder	7. Loopseal
4. Measuring tank	

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s and 0-40 kg/m^{2}s, respectively. Pressure taps were mounted flush with the wall of the column and connected to pressure transducers. The resulting pressure drop signals were stored in a computer through a data acquisition system.

In the constant velocity section defined by Wang et al. [1992], ΔP_a=0 so that the particle suspension density in this study was determined from the time averaged pressure drop gradient in the constant velocity section. The axial profiles of pressure and pressure gradient indicate that length of constant velocity section is 15 to 20 times the bed diameter from the entrance of the downer. Solids circulation rate was measured by using a measuring tank. Since solid holdup (ε_s) does not vary appreciably in the constant velocity section, the location of tracer gas (He) injection was selected at 2.0 m from the entrance of the downer.

The detailed scheme of the gas mixing experiment is shown in Fig. 2. Tracer gas was injected at steady state through a down-facing injection tube (0.95 mm-I.D) at the center of the downer with the injection velocity as of the superficial gas velocity in the downer. Tracer gas in the downer was sampled at 0.15 m below the tracer gas injection point. The sampling probe (6 mm-I.D.) was covered with a mesh screen to prevent the entrance of solid particles from the bed. As solids fall in the downer, the sampling probe was faced downward to avoid retention of solids on the probe.

THEORY

In general, the gas mixing coefficient was determined by using the dispersion model [Yang et al., 1984; Bader et al., 1988; Lee and Kim, 1989; Zheng et al., 1992; Namkung and Kim, 1998]. A mass balance on a differential volume element ($2\pi r dr dx$) of the downer under the conditions of steady state leads to Eq. (1).

$$U_g\frac{\partial C}{\partial x}=D_a\frac{\partial^2 C}{\partial x^2}+D_r\frac{1}{r}\frac{\partial}{\partial r}\left(r\frac{\partial C}{\partial r}\right) \tag{1}$$

where x is the distance between the injection and sampling planes, which is positive in the downward direction. The radial distance from the centerline of the downer is given by r and C is the tracer gas concentration. Assuming that the axial dispersion can be neglected without gas backmixing in the downer [Wei et al., 1995], Eq. (1) becomes

$$U_g\frac{\partial C}{\partial x}=D_r\frac{1}{r}\frac{\partial}{\partial r}\left(r\frac{\partial C}{\partial r}\right) \tag{2}$$

with the following boundary conditions:

$$x=-\infty,\ C\to 0$$
$$r=R,\ \partial C/\partial r=0 \tag{3}$$

Then, Eq. (1) has a solution of the form [Klinkenberg et al., 1953]

$$\frac{C}{C_o}=1+\sum_{i=1}^{\infty}\frac{J_0(a_i r^*)}{J_0^2(a_i)}\exp\left(-\frac{2a_i^2 x^*}{Pe_r}\right) \tag{4}$$

where $Pe_r=U_g D_t/D_r$, $r^*=r/R$, $x^*=x/R$, $J_1(\alpha_i)=0$, J_i is the Bessel function of first kind of i-th order, and α_i is the positive roots in ascending order of Bessel function J_1.

The average concentration C_o in this study is calculated from

$$C_0=\frac{1}{\pi R^2}\int_0^R 2\pi r C dr \tag{5}$$

Peclet number (Pe_r) in Eq. (4) can be determined by the optimization technique with the measured values of C/C_o at r^* and x^*. From the obtained values of C/C_o vs. r are well fitted to Eq. (4) from which D_r value was derived.

RESULTS AND DISCUSSION

To prevent stagnation of solids on the sampling probe in the downer, a down-facing sampling probe should be employed. However, we have tested the tracer gas concentration profiles

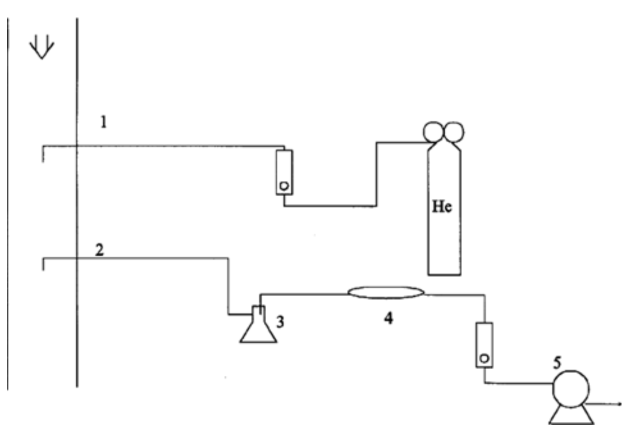

Fig. 2. Injection and sampling systems for the gas mixing experiment.
1. Injection probe
2. Sampling probe
3. Flask
4. Gas sampling column
5. Pump

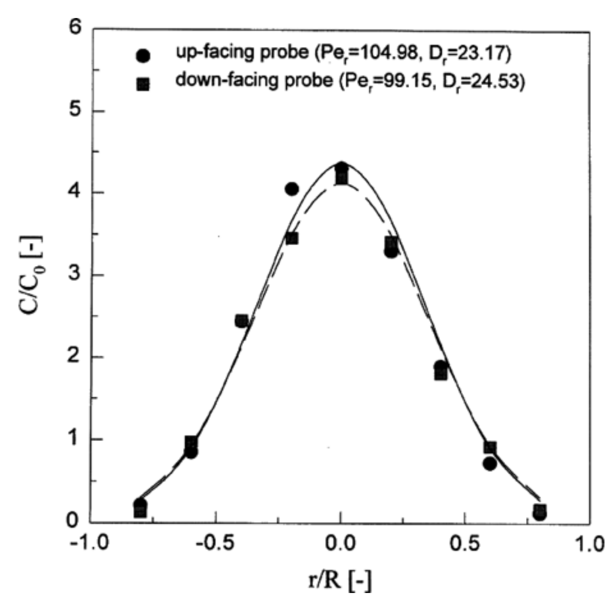

Fig. 3. Comparison of the measured concentration profile between the up- and down-facing probe in the downer of gas alone (U_g=2.43 m/s).
[Symbol : experimental data; Line: calculated values by the Eq. (4)]

between the down- and up-facing sampling probes as shown in Fig. 3. As can be seen, the difference in tracer gas concentration profiles is found to be small so that a down-facing sampling probe was used in this study. An analytical solution of Klinkenberg [1953] seems to fit well with the measured tracer gas concentration profiles. Therefore, the radial Peclet number (Pe_r) can be determined by using the solution of Klinkenberg [Eq. (4)].

The effect of gas velocity (U_g) on D_r and Pe_r in the downer of gas flow alone is shown in Fig. 4. As can be seen, D_r increases linearly, and D_r value is higher than that of the Fickian gas diffusion coefficient ($D_{Air\text{-}He}$=0.71 cm^2/s) [Hirshfelder et al., 1974] due to the enhancement of turbulent intensity with gas flow. Pe_r is nearly constant in the range of 98-116 with increasing gas velocity as reported previously [Sherwood et al., 1975; Werther et al., 1992]. The resulting mean value of Pe_r is 107, which is somewhat lower than the expected one since the measured gas velocity profiles by an air-velocity-meter are flatter than

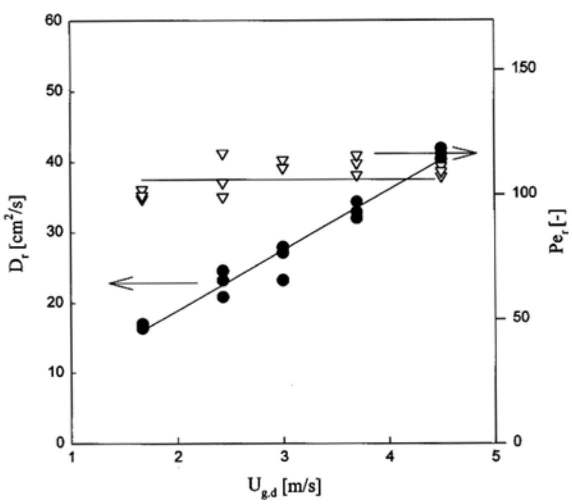

Fig. 4. Effect of gas velocity on D_r and Pe_r in the downer of gas flow alone.

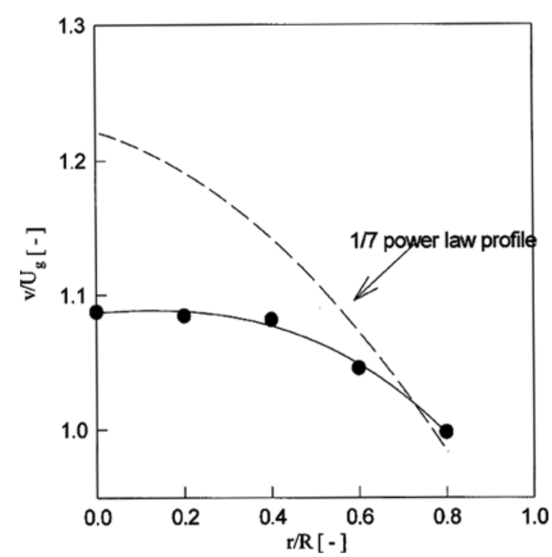

Fig. 5. Measured gas velocity profile in the downer of gas flow alone (U_g=2.43 m/s).

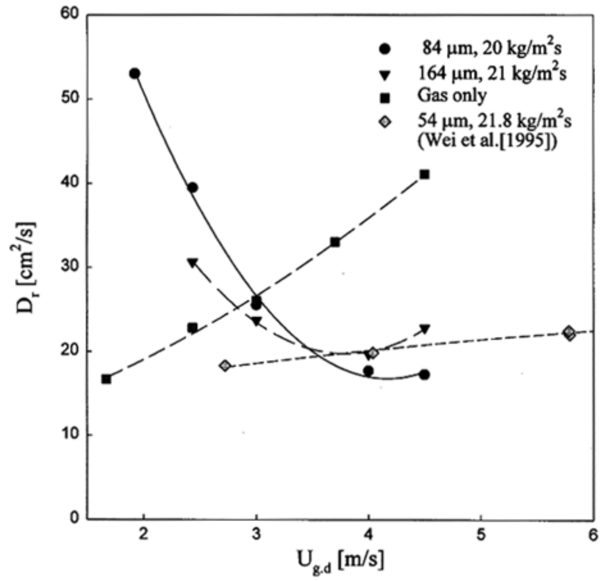

Fig. 6. Effect of gas velocity on D_r.

the turbulent velocity profile as shown in Fig. 5 [Amos et al., 1993].

The effect of U_g on D_r in the presence of solid particles in the downer reactor is shown in Fig. 6 where D_r decreases with increasing U_g. At lower U_g, D_r in the presence of solid particles is higher than that of gas flow alone but a reverse trend was found at higher U_g. These gas mixing characteristics can be explained by cluster formation in the downer at lower U_g. Aubert et al. [1994] reported that particle terminal velocity increases with the formation of clusters at lower U_g in a downer that may have higher positive slip velocities (U_{sl}=U_p–U_g) than terminal velocities because clusters behave like larger particles. In this study, we have observed higher slip velocities than terminal velocities at lower U_g as shown in Fig. 7. Since gas dispersion is a function of turbulent intensity [Adams, 1988; Namkung and Kim, 1998], the larger particles (0.5, 1 and 3 mm) produce more turbulence than the smaller ones (200 mm) as reported by Tsuji et al. [1984]. At lower U_g, D_r is higher than that of gas flow alone since clusters behave like larger particles in

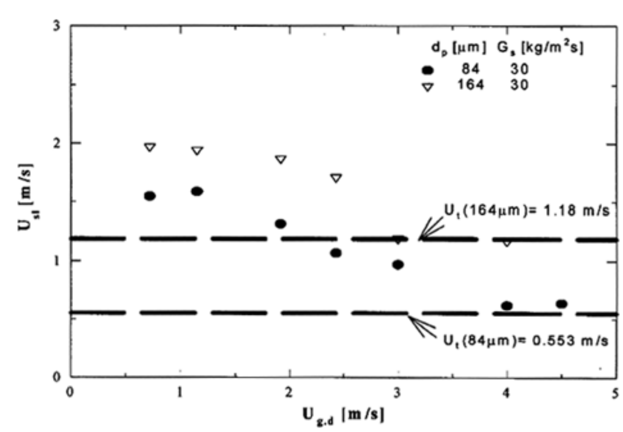

Fig. 7. Effect of gas velocity on slip velocity with the variation of particle size.
(G_s=30 kg/m^{2}s)

the bed [Aubert et al., 1994]. Also, variation of D_r with different particle size can be seen in Fig. 6 where D_r varies significantly with small particles but it varies marginally with larger particles with increasing U_g.

The effect of solid circulation rate (G_s) on D_r is shown in Fig. 8. At lower gas velocities (U_g<3.0 m/s), D_r increases, whereas it decreases with increasing G_s at higher gas velocities (U_g>3.0 m/s) since clusters behave like larger particles at lower U_g that provide higher turbulent intensity in the downer [Tsuji et al., 1984]. The influence of suspension density (ρ_{sus}) on D_r is shown in Fig. 9. As can be seen, the influence of suspension density on D_r is very similar to that of G_s on D_r.

To compare D_r values in the downer and the riser reactors, the obtained D_r values and details of the experimental and operating conditions of the present and previous studies are summarized in Table 1. D_r in the downer is higher than that in the riser at similar operating conditions. Therefore, it can be claimed that gas mixing in the downer is superior to that in the riser.

Recently, Wei et al. [1995] proposed a correlation of the radial gas mixing in the downer as:

$$Pe_r = 3.14\times10^{-3} Re^{0.95} \varepsilon^{-7.34} \tag{6}$$

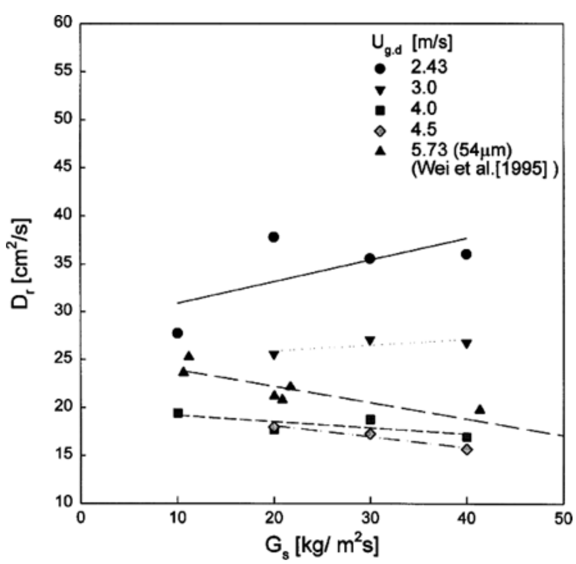

Fig. 8. Effect of solids circulation rate on D_r (d_p=84 μm).

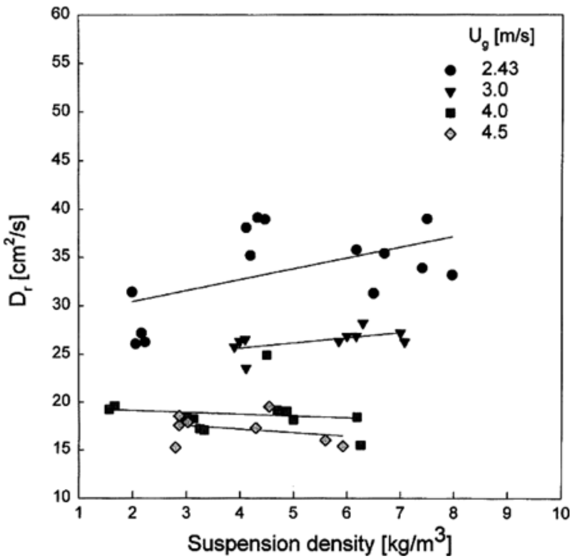

Fig. 9. Effect of suspension density (ρ_{sus}) on D_r (d_p=84 μm).

As can be seen in Fig. 10, the agreement between D_r values of the present study and from Eq. (6) is found to be very poor. In this study, the radial gas mixing characteristics have been studied at lower range of gas velocities compared to the study of Wei et al. [1995]. The radial gas mixing coefficients in terms of Pe_r of the present and previous study [Wei et al., 1995] of downer reactors have been correlated as a function of physical properties of gas and solid (ρ_s, ρ_g, d_p) phases, reactor diameter (D_t) and operating variables (U_g, G_s) as:

$$Pe_r = 1.513\times10^{-6}\left(\frac{G_s}{\rho_g U_g}\right)^{0.167} Re^{1.671}\left(\frac{d_p}{D_t}\right)^{0.103}\left(\frac{\rho_s-\rho_g}{\rho_g}\right)^{0.294} \tag{7}$$

This correlation covers the ranges of variables $3.487\le G_s/(\rho_g U_g)\le 19.166$, $22857\le Re\le 72240$, $3.857\times10^{-4}\le d_p/D_t\le 16.4\times10^{-4}$ and $1285\le(\rho_s-\rho_g)/\rho_g\le 2869$ with a correlation coefficient of 0.960. The goodness of fit between experimental and calculated values of the radial Peclet number is shown in Fig. 11.

CONCLUSIONS

The radial gas dispersion coefficients in a downer reactor have been determined by using the dispersion model. The ra-

Table 1. Details of the experimental and operating conditions of the present and previous studies

Authors	Column size	Particles	Particle size [μm] Density [kg/m³]	Tracer gas	U_g [m/s] G_s [kg/m²s]	D_r [cm²/s]
Yang et al. [1984]	0.115 m-ID, 8 m high	Silica-gel	220/730	He	2.8-5.5 32-160	2.2-7.6
Zheng et al. [1992]	0.102 m-ID, 5.25 m high	Resins	701/1392	CO_2	0-6.2 0-30	5.8-18.3
Namkung et al. [1998]	0.10 m-ID, 5.3 m high	Silica sand	125/3055	CO_2	2.5-4.5 0-53	7.0-17.2
Wei et al. [1995]*	0.14 m-ID, 7.6 m high (5.8 m)	FCC particles Al_2O_3	54, 1398 54, 1710	H_2	2.0-9.0 0-70	14.8-31.2
This study*	0.10 m-ID, 3.6 m high	Silica sand	84, 104, 164/3120	He	1.9-4.5 0-40	17.3-53.0

*: downer

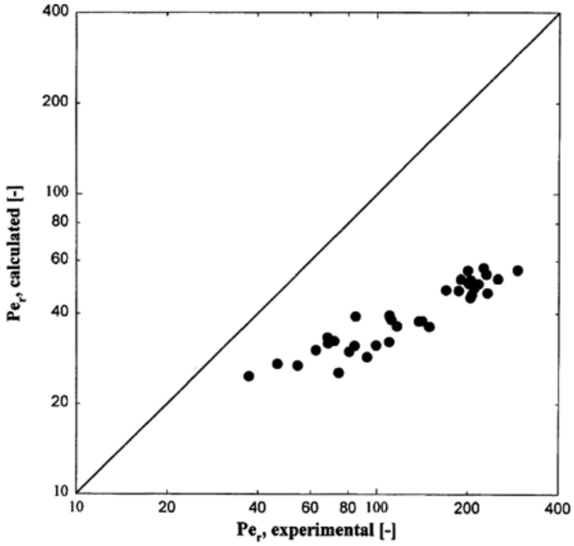

Fig. 10. Comparison of Pe_r between the experimental and calculated values from the correlation of Wei et al. [1995] [Eq. (6)].

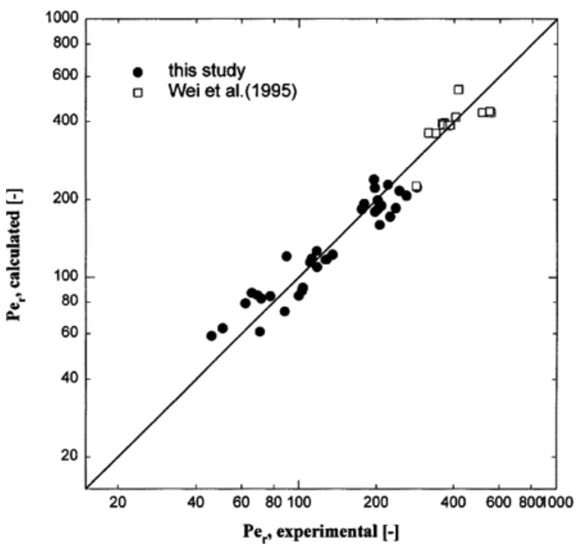

Fig. 11. Comparison of Pe_r between the experimental and calculates values from the proposed correlation [Eq. (7)].

dial dispersion coefficient (D_r) decreases and the radial Peclet number (Pe_r) increases as the gas velocity increases. At lower gas velocities, D_r in the presence of particles is lower than that of gas flow alone but the trend is reversed at higher gas velocities. The radial dispersion coefficient (D_r) varies significantly with smaller particles, but the variation is dampened with larger particles with increasing gas velocity. At lower gas velocities, D_r increases with increasing solids circulation rate (G_s). However, D_r decreases with G_s at higher gas velocities. Based on the obtained D_r values in the downer reactor, it can be concluded that the downer reactor is a good gas-solids contacting reactor having good radial gas mixing.

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NOMENCLATURE

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_p : particle diameter [m]
D_r : radial gas dispersion coefficient [m^2/s]
D_t : diameter of column [m]
G_s : solids circulation rate [kg/m^{2}s]
J_i : Bessel function of first kind of i-th order
Pe_r : Peclet number, U_gD_t/D_r [-]
R : riser radius [m]
Re : Reynolds number, $\rho_gU_gD_t/\mu$ [-]
r : radial position [m]
r^* : dimensionless radial distance, r/R [-]
U_g : superficial gas velocity [m/s]
U_p : particle velocity [m/s]
U_{sl} : slip velocity ($=U_p-U_g$) [m/s]
v : local gas velocity [m/s]
x : vertical distance from tracer gas injection level [m]
x^* : dimensionless axial distance, x/R [-]

Greek Letters

ΔP_a : pressure drop due to acceleration [pa]
ε_s : solid holdup [-]
ε : voidage [-]
ρ_g : gas density [kg/m^3]
ρ_s : particle density [kg/m^3]
ρ_{sus} : suspension density [kg/m^3]
μ : gas viscosity [kg/ms]

REFERENCES

Adams, C. K., "Gas Mixing in Fast Fluidized Beds," Circulating Fluidized Bed Technology II, Basu, P. and Large, J. F., eds., 299 (1988).

Amos, G., Rhodes, M. J. and Mineo, H., "Gas Mixing in Gas-Solids Risers," *Chem. Eng. Sci.*, **48**, 943 (1993).

Aubert, E., Barreteau, D., Gsuthier, T. and Pontier. R., "Pressure Profiles and Slip Velocities in a Co-Current Downflow Fluidized Bed Reactor," Circulating Fluidized Bed Technology IV, Adidan, A. A., eds., 403 (1994).

Bader, R., Findlay, J. and Knowlton, T. M., "Gas/Solid Flow Patterns in a 30.5 cm-Diameter CFB," Circulating Fluidized Bed Technology II, Basu, P. and Large, J. F., eds., 123 (1988).

Bai, D.-R., Jin, Y., Yu, Z.-Q. and Zhu, Z.-X., "The Axial Distribution of the Cross-Sectionally Averaged Voidage in Fast Fluidized Beds," *Powder Technol.*, **71**, 51 (1992).

Cho, Y. J., Kim, S. D. and Han, G. Y., "Bed-to-Wall Heat Transfer Characteristics in a Circulating Fluidized Bed," *Korean J. Chem. Eng.*, **13**, 627 (1996).

Gayan, P., de Diego, L. F. and Adanez, J., "Radial Gas Mixing in a

Fast Fluidized Bed," *Powder Technol.*, **94**, 163 (1997).

Han, G. Y., Lee, G. S. and Kim, S. D., "Hydrodynamic Characteristics of a Circulating Fluidized Bed," *Korean J. Chem. Eng.*, **2**, 141 (1985).

Hirschfelder, J. O., Curtis, C. F. and Bird, R. B., "Molecular Theory of Gases and Liquids," John Wiley & Sons, New York (1964).

Klinkenberg, A., Krajenbrink, H. J. and Lauwerier, H. A., "Diffusion in a Fluid Moving at Uniform Velocity in a Tube," *Ind. Eng. Chem.*, **45**, 1202 (1953).

Lee, G. S. and Kim, S. D., "Gas Mixing in Slugging and Turbulent Fluidized Beds," *Chem. Eng. Comm.*, **86**, 91 (1989).

Namkung, W. and Kim, S. D., "Gas Mixing Characteristics in a Fast Fluidized Bed," *HWAHAK KONGHAK*, **36**, 797 (1998).

Sherwood, T. K., Pigford, R. L. and Wilke, C. R., "Mass Transfer," McGraw-Hill, New York, 112 (1975).

Tsuji, Y., Morikawa, Y. and Shiomi, M., "LDV Measurement of an Air-Solid Two-Phase Flow in a Vertical Pipe," *J. Fluid Mech.*, **139**, 417 (1984).

Wang, Z., Bai, D.-R. and Jin, Y., "Hydrodynamics of Cocurrent Downflow Circulating Fluidized Bed (CDCFB)," *Powder Technol.*, **70**, 271 (1992).

Wei, F., Jin, Y., Yu, Z. and Liu, J., "Gas Mixing in the Cocurrent Downflow Circulating Fluidized Bed," *Chem. Eng. Technol.*, **18**, 59 (1995).

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).

Yang, G., Zhe, H. and Zhao, L., "Radial Gas Mixing in a Fast Fluidized Bed," Fluidization IV, Kunii, D. and Toei, R., eds., Engineering Foundation, 145 (1984).

Zheng, Q., Wei, X. and Fei, L., "Experimental Study on Radial Gas Dispersion and its Enhancement in Circulating Fluidized Beds," Fluidization VII, Potter, O. E. and Nicklin, D. J., eds., Engineering Foundation, 285 (1992).

Zhu, J.-X., Yu, X.-Q., Jin, Y., Grace, J. R. and. Issangya, A., "Cocurrnet Downflow Circulating Fluidized Bed (Downer) Reactors–A State of the Art Review," *Can. J. Chem. Eng.*, **73**, 662 (1995).