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Zhiguo Wang* C. Jim Lim†

Xiaotao T. Bi‡

*University of British Columbia

†The University of British Columbia, cjlim@chml.ubc.ca

‡University of British Columbia, xbi@chml.ubc.ca

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EXPERIMENTAL AND COMPUTATIONAL STUDIES OF GAS MIXING IN CONICAL SPOUTED BEDS

Zhiguo Wang, C. Jim Lim and Hsiaotao T. Bi
Fluidization Research Centre, University of British Columbia, Vancouver, Canada

ABSTRACT

The residence time distribution data in a conical spouted bed, obtained both from the bed bottom and the bed surface at different radial positions, were analyzed to obtain the mean residence time and the Peclet number. In parallel, local flow structures of a bed with the same dimensions and operating conditions as in the experiment were generated from the computational fluid dynamics (CFD) simulation using the FLUENT codes, and then were used for the simulation of gas dispersion. The results show that CFD simulations agree reasonably well with experiments. The radial distribution of the Peclet number is quite complex, with a maximum value at r=0.135 m under three operating conditions investigated.

INTRODUCTION

The gas residence time distribution is of considerable importance in predicting the conversion and selectivity for various catalytic reactions, and backmixing is undesirable as it may lead to increased by-products. Both vertical and horizontal mixing/dispersion can be studied using steady and unsteady state tracer techniques. In the steady state tracer experiment, a steady flow of tracer gas is introduced into the spouted bed at a certain location, and the tracer concentration is measured either downstream or upstream of the injection point. Ideally, the injection rate should be adjusted to match the local gas velocity in the bed to achieve an isokinetic injection ((1). Based on the tracer concentration measured upstream of the injection point, axial backmixing coefficient can be derived (2). On the other hand, the radial dispersion coefficient is obtained by analyzing radial profiles of tracer concentrations measured downstream of the injection point (1). The overall or effective axial dispersion coefficient over the entire bed could be derived using the unsteady state tracer technique.

For gas-solid multiphase systems, there have been a large number of researches on gas backmixing and/or radial dispersion in bubbling fluidized beds, circulating fluidized beds/risers and downers, (e.g. 3-7). However, there have been only a few studies on gas mixing in cylindrical spouted beds and conical spouted beds (e.g. 8-13), and almost no reports on the residence time distribution (RTD) simulation based on computational fluid dynamics (CFD) for spouted beds. In this study, gas dispersion in conical spouted beds was investigated both experimentally and analytically based on tracer techniques and CFD simulations, respectively.

EXPERIMENTS

The conical spouted bed (full column) used in this study is made of Plexiglas with an included angle γ of 45°. The diameter at the conical base D_i is 0.038 m, the diameter of the nozzle D_0 is 0.019 m, and the diameter of the upper cylindrical section D_c is 0.45 m. For more details, please see Wang et al. (14). Glass beads of 1.16 mm in mean diameter with a narrow size distribution were used as the bed material, compressed air at the ambient temperature was used as the spouting gas. The static bed height employed in the experiment is 0.40 m, and spouting velocities of 23.5 and 17.0 m/s based on the bottom inlet were used in the experiment with the higher one for stable spouting and the lower one for internal/partial spouting.

During experiments, helium was used as the gas tracer. For RTD measurements, the tracer was introduced as a step function into the spouting air far away from the bottom of the conical spouted bed by a solenoid valve, and the unsteady state response was measured by a thermal conductivity detection (TCD) system. Two sampling probes of 1 mm ID were connected separately to two TCDs to measure the tracer concentration, with one located just below the gas inlet and the other just above the bed surface, with a separation distance of 0.4 m. Both probes can be radially traversed to measure the tracer concentration at different radial positions. Output signals from TCDs were amplified and collected via a data acquisition system.

ESTIMATION OF THE GAS MIXING BEHAVIOUR

For a negative step input function, the response curve from each sampling probe can be converted to the cumulative distribution function *F*(*t*) by

$$F(t) = 1 - \frac{V(t) - V_\infty}{V_0 - V_\infty} \tag{1}$$

where V_0 is the average voltage signal corresponding to $C_{He}=C_0$, V_∞ is the average value corresponding to $C_{He}=0$, and V(t) is the transient value.

To convert the cumulative distribution function into RTD function, the cumulative distribution function *F*(*t*) was fitted first by using Levenberg-Marquardt method, and the RTD function *E(t)* was obtained by differentiating the fitted *F*(*t*) curve.

$$E(t) = \frac{dF(t)}{dt} \tag{2}$$

For the two-probe measurement system, the mean residence time and the dispersion Peclet number of the dense section can be derived from the difference between measured RTD functions below the gas inlet and above the bed surface by assuming that spouted bed behaves like a linear system, as shown in Figure 1.

For an open-closed system, the axial Peclet number *Pe* can be related to the variance for a flow system with small backmixing by (15)

$$\sigma^2 = \frac{2}{Pe} + 3\left(\frac{1}{Pe}\right)^2 \tag{3}$$

Wang et al.: Studies of Gas Mixing in Conical Spouted Beds

where $\sigma^2 = \Delta\sigma_t^2 / (\Delta\hat{t})^2$ is the variance, Pe is the Peclet number, $Pe = u_g L / D$, u_g is the interstitial gas velocity, L is the distance between two sampling points, D is the dispersion coefficient.

EXPERIMENTAL RESULTS AND DISCUSSIONS

Figure 2 shows some experimental raw signal V, calculated F functions and E functions at the inlet as well as at the bed surface based on negative step tracer experiments with response time lags included. It can be seen from Figure 2 that the response at the gas inlet was not a perfect step function, which could mean either that there exists gas backmixing between the tracer injection point and TCD 1 or that the tracer injection was not a perfect step function.

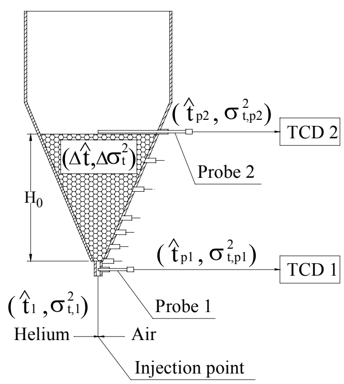


Fig. 1. Definition of the mean residence time and corresponding variance for different sections.

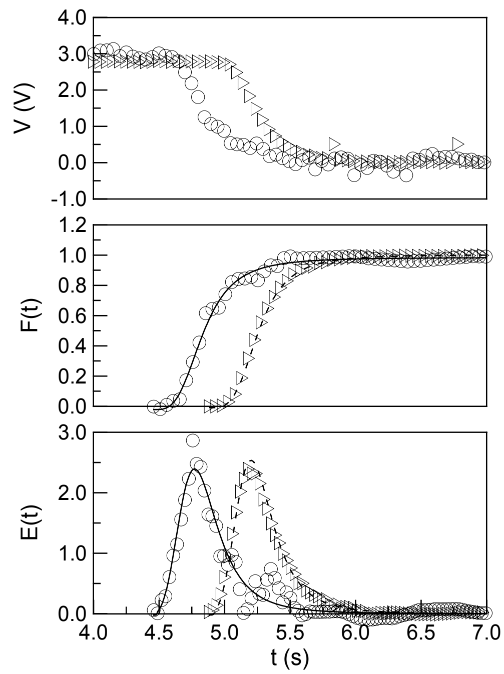
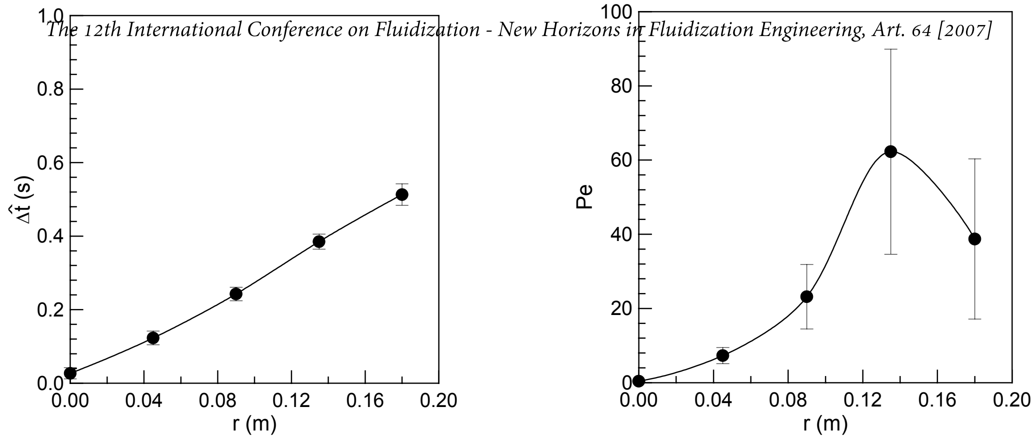


Fig. 2. Experimental raw data V, calculated F functions and E functions at the inlet (O) and the bed surface (▷) with the probe located at the axis. (Stable spouting, U_i=23.5 m/s, full column)

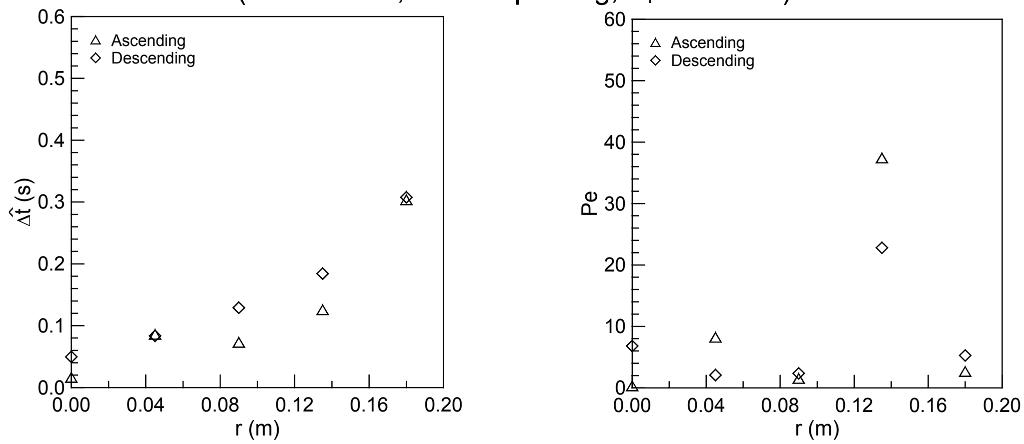
Figures 3 to 4 show the radial distribution of the mean residence time and the Peclet number at different operating velocities or states, one in stable spouting state and other two at partial spouting states in the gas velocity ascending and descending processes, respectively. Figure 3(a) shows that the mean residence time increases with increasing radial distance from the centre of the column toward the wall, meaning that gas velocity inside the conical spouted bed has a radial distribution, higher in the centre and lower near the wall. Figure 3(b) shows that the radial distribution of the Peclet number is quite complex, with a maximum value at r=0.135 m. This trend is commonly observed at different operating velocities or states as shown in Figure 4. Meanwhile, the radial distribution of gas velocity in the ascending process is different from that in the descending process, so is the gas mixing behaviour, even though operating velocities are almost the same.

The 12th International Conference on Fluidization - New Horizons in Fluidization Engineering, Art. 64 [2007]



(a) (b)

Fig. 3. Radial distribution of the mean residence time and the Peclet number. (Full column, stable spouting, U_i=23.5 m/s)



(a) (b)

Fig. 4. Radial distribution of the mean residence time and the Peclet number. (Full column, partial spouting, $U_{i,d}$=17.05 m/s, Z_d=0.216 m or $U_{i,a}$=16.95 m/s, Z_a=0.131 m)

SIMULATION OF GAS MIXING IN A CONICAL SPOUTED BED

General gas mixing model

From the analysis of a small control volume in the vertical direction with the assumptions that (a) the dispersion coefficient is constant within the bed and (b) gas density remains constant within the bed, the following expression can be derived:

$$\frac{d(\varepsilon_g \cdot \rho_g \cdot X_a)}{dt} + \frac{d(\varepsilon_g \cdot \rho_g \cdot v_{g,z} \cdot X_a)}{dZ} - \frac{d}{dZ}(\varepsilon_g \cdot \rho_g \cdot D \cdot \frac{dX_a}{dZ}) = 0 \tag{4}$$

In X and Y directions, similar expressions can be obtained, and the general three-dimensional equation in Cartesian coordinates can be written as

$$\frac{\partial(\varepsilon_g \cdot \rho_g \cdot X_a)}{\partial t} + \nabla \cdot (\varepsilon_g \cdot \rho_g \cdot \overrightarrow{v_g} \cdot X_a) - \nabla \cdot [\varepsilon_g \cdot (\rho_g \cdot D) \cdot \nabla \cdot (X_a)] = 0 \tag{5}$$

Because the flow rate of the tracer is very small with a maximum helium volume fraction of 0.3 % during experiments, the volume fraction of the helium is assumed to be proportional to the mass fraction of the helium.

Wang et al.: Studies of Gas Mixing in Conical Spouted Beds

$$X_a = C_{He} \cdot \frac{\rho_{He}}{\rho_g} \qquad (6)$$

where C_{He} is the volume fraction of helium, ρ_{He} is the density of helium.

As discussed before, using the negative step tracer input, the cumulative distribution function *F(t)* can be written as Equation (1). Considering the linear characteristics of Equations (1) and (6), as well as the assumed linear characteristics of the sampling probes, for convenience, a pseudo positive step function curve derived from experiments was used as the tracer input in the current simulation with following boundary conditions:
When $t-t_i<0$, $X_a=0$;
When $t-t_i \geq 0$, $X_a=F(t-t_i)$ at the tracer inlet (Pseudo positive step function)
$\frac{\partial X_a}{\partial z}=0$ at the outlet, and $\frac{\partial X_a}{\partial r}=0$ at the wall

In order to simulate gas mixing behaviour in a conical spouted bed, gas velocity field as well as the distribution of the voidage need to be calculated first. Thus, a conical spouted bed with the same geometrical dimensions and operating conditions as those in the current experiment was simulated first using the FLUENT simulation software, with details given in Wang et al. (14).

In the simulation, once stable spouting has been reached and the average gas velocity field and voidage distribution were calculated, a DEFINE_ON_DEMAND function was activated to pass the averaged gas velocity field and the voidage distribution to three User Defined Memories (UDMs) for further simulation on gas mixing behaviour. At the same time, the current time t_i was obtained. After changing t_i to the exact value just obtained, the User Defined Function was activated again. To achieve the positive step injection, the whole column should be patched with 0.0 for the User Defined Scalar (UDS) ϕ after it had been defined (including defining UDS and the corresponding dispersion coefficient), and corresponding boundary conditions should be defined too. To simulate gas mixing at the stable spouting state and to save computation time, all equations were turned off except the newly defined UDS equation.

Figures 5 and 6 show the comparison between experimental results and CFD simulation results on the mean residence time and Peclet number. It can be seen that the dispersion coefficient affects simulation results significantly, with better agreement between experimental and simulated results at D=0.0002 m^2/s for the central spout region. Near the wall, simulation results underestimate the Peclet number greatly and overestimate the mean residence time significantly for all values of the dispersion coefficient investigated. With a small dispersion coefficient, CFD simulation gives a similar radial distribution curve on the Peclet number as that from the experiment. The existence of a peak Peclet number in the radial profile can thus be attributed to the compromise of the decreasing gas velocity and the increasing gas streamline length as the radial location changes from the centre to the wall. Figure 6 suggests that the difference between the CFD simulation and the experiment still cannot be resolved even using different values of the dispersion coefficient for the spout and the annulus. Moreover, neglecting the dispersion (D=0.0), the difference between the CFD simulation and the experiment near the wall (r=0.180 m) still exists, suggesting that gas convection may be the dominant factor near the wall. On the other hand, that the simulated gas velocity is lower near the wall (or higher in the spout) than from the experiment suggest that the mismatch between the simulated and measured Pe may result from the mismatch of the simulated and measured radial velocity profiles.

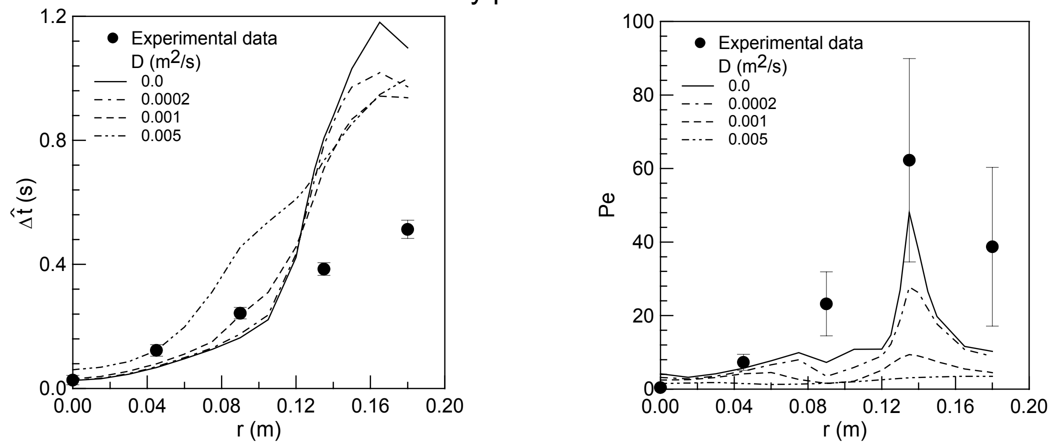


Fig. 5. Comparison between the experiments (symbols) and CFD simulations (lines). (Full column, stable spouting, U_i=23.5 m/s.)

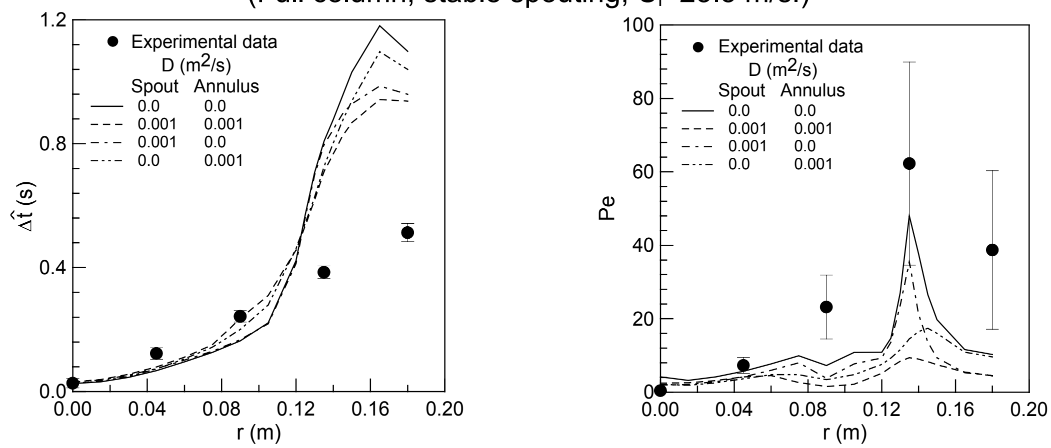


Fig. 6. Comparison between the experiments (symbols) and CFD simulations (lines). (Full column, stable spouting, U_i=23.5 m/s.)

CONCLUSIONS

Helium tracer experiments clearly show that there are radial distributions of the gas velocity inside a conical spouted bed with higher velocity in the centre and lower near the wall. There exists a maximum value of Peclet number at r=0.135 m, resulting from the balance of decreasing gas velocity and increasing gas streamline length with increasing the radial distance from the center to the wall. Meanwhile, the gas mixing in the ascending process is different from that in the descending process, implying that the radial distribution of gas velocity is different too, even though operating velocities are identical. The gas mixing was also simulated using a CFD model. It was found that, with a small dispersion coefficient, CFD simulation gives a similar radial distribution curve on the Peclet number as that from the experiment. The difference between the CFD simulation and the experiment cannot be eliminated using different values of the dispersion coefficient for the spout and the annulus, likely due to the mismatch of simulated and measured radial velocity profiles.

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NOTATION

C_0	Initial volume fraction of helium in air, (%v/v)
C_{He}	Volume fraction of helium in air, (%v/v)
D	Dispersion coefficient, (m^2/s)
D_0	Gas inlet diameter, (m)
D_c	Diameter of the cylindrical section, (m)
D_i	Diameter of the bed bottom, (m)
E(t)	RTD function
F(t)	Cumulative distribution function
H_0	Static bed height, (m)
L	Distance between two sampling points, (m)
Pe	Peclet number, $=(u_g L)/D$, (-)
r	Radial coordinate, (m)
t	Time, (s)
t_i	Time starting the injection of tracer gas, (s)
$\hat{t}_1$	Mean residence time between the tracer injection point and the tip of the probe 1, (s)
$\hat{t}_{p1}$	Mean residence time for the electric signal from probe 1, (s)
$\hat{t}_{p2}$	Mean residence time for the electric signal from probe 2, (s)
u_g	Interstitial gas velocities, (m/s)
U_i	Superficial gas velocity based on D_i, (m/s)
$U_{i,a}$	Superficial gas velocity based on D_i during ascending process, (m/s)
$U_{i,d}$	Superficial gas velocity based on D_i during descending process, (m/s)
$\vec{v}_g$	Velocity vector of the gas phase, (m/s)
$v_{g,z}$	Local axial interstitial gas velocity, (m/s)
V	Magnitude of the measured electrical signal, (V)
V_0	Average value of the electrical signal corresponding to $C_{He}=C_0$, (V)
V_∞	Average value of the electrical signal corresponding to $C_{He}=0$, (V)
X_a	Mass fraction of the tracer gas, (-)
Z	Axial coordinate, (m)
Z_a	Height of the internal spout in the ascending process, (m)
Z_d	Height of the internal spout in the descending process, (m)

Greek letters

$\Delta\hat{t}$	Mean residence time inside the conical spouted bed, (s)
$\Delta\sigma_t^2$	Variance corresponds to $\Delta\hat{t}$, (s^2)
ε_g	Voidage, (-)

The 12th International Conference on Fluidization - New Horizons in Fluidization Engineering, Art. 64 [2007]

ϕ	Arbitrary scalar in the gas phase, (-)
γ	Cone angle, (°)
ρ_{He}	Density of helium, (kg/m^3)
ρ_g	Density of the gas phase, (kg/m^3)
σ^2	Variance, $\sigma^2 = \Delta\sigma_t^2 / (\Delta\hat{t})^2$ (-)
$\sigma_{t,1}^2$	Variance corresponding to $\hat{t}_1$, (s^2)
$\sigma_{t,p1}^2$	Variance corresponding to $\hat{t}_{p1}$, (s^2)
$\sigma_{t,p2}^2$	Variance corresponding to $\hat{t}_{p2}$, (s^2)

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