Experimental Study of Gas Mixing in a Spout-Fluid Bed

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Gas mixing in a spout-fluid bed with a cross section of 0.3 × 0.03 m and height of 2 m was investigated by simultaneously injecting two different tracer gases. One was injected into the spouting gas flow and the other was injected into the fluidizing gas flow. Steady-state tracer gas measurements were carried out to obtained radial tracer gas concentrations at various bed elevations. Effects of two important operating parameters—spouting gas velocity and fluidizing gas flow rate—on the gas mixing were discussed with flow patterns recorded by a high-resolution digital CCD camera. The results show that increasing spouting gas velocity and fluidizing gas flow rate can both promote the gas mixing in spout-fluid beds. Increasing fluidizing gas flow rate is the more effective way, given that a satisfactory mixing condition at a relatively low bed height can be obtained by increasing the fluidizing gas flow rate. For both cases, it is difficult to obtain a good mixing condition in the wall layers. Besides, the mechanism of gas mixing was preliminarily discussed. Results indicate that gas mixing in the spout-fluid bed is caused by both convection and diffusion. Diffusion other than molecular diffusion should not be neglected, especially at high spouting gas velocity or fluidizing gas flow rate. © 2005 American Institute of Chemical Engineers *AIChE J,* 52: 924–930, 2006
Keywords: fluidization, spout-fluid bed, jetting fluidized bed, gas mixing, tracer

Introduction

Spout-fluid beds are modified from spouted beds to provide better solid–fluid contact and mixing by introducing an auxiliary flow into the annular region. Spout-fluid beds have been applied in drying, particle coating, granulation, gas cleaning, heating and cooling solids, solids mixing, shale pyrolysis, coal gasification, and combustion.[1-4] For coal gasification, the CRE (Coal Research Establishment) in Great Britain had suggested adding a spout-fluid bed partial coal gasifier to the original PFBC (pressurized fluidized bed combustion–combined cycle) system, combining gasification with combustion, to improve the circulating efficiency to about 45–47%.[5] In China, spout-fluid bed coal gasifiers have been developed for both the APFBC-CC (advanced pressurized fluidized bed combustion–combined cycle) system and the PPG-CC (pressurized partial gasification–combined cycle) system by Southeast University.[6-11]

Coal gasification in the spout-fluid bed is similar to the U-gas or Westinghouse processes. In addition to injecting a portion of the gasification agent through the central nozzle, the gasification agent is also introduced through a porous or perforated distributor surrounding the central nozzle. Combustion mainly occurs in the center-spout jet region and gasification mainly processes in the annular dense region. Good interchanges of mass, heat, and momentum between these two regions would contribute to a high degree of gasification efficiency.[5,6] Therefore, an adequate knowledge of the gas mixing behavior is helpful for understanding, evaluating, and scaling-up of spout-fluid bed gasifiers.

A number of published studies in the literature[12-19] have reported valuable results of gas mixing in dense gas–particle flow systems. However, thus far there has been little informa-

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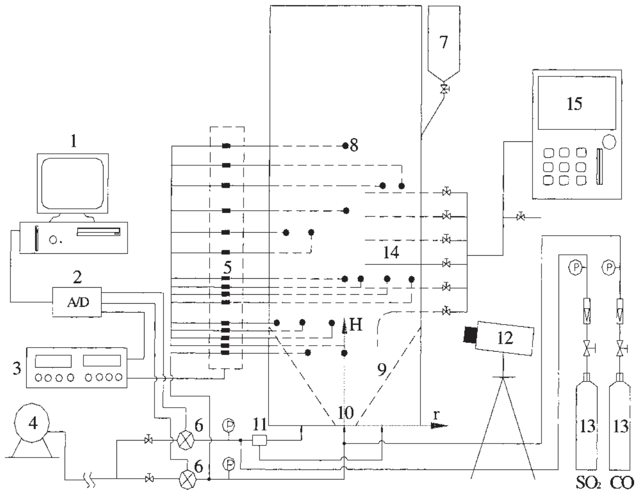

Figure 1. Spout-fluid bed experimental system for gas mixing.

1, Computer; 2, A/D converter; 3, multidifferential pressure signal transmitter; 4, roots-type blower; 5, differential pressure sensor; 6, flow meter; 7, material adding tank; 8, pressure port; 9, V-type gas distributor; 10, spouting gas nozzle; 11, fluidization flux distributor; 12, digital CCD; 13, tracer gas; 14, sampling tube; 15, gasometrical analyzer.

tion concerning the gas mixing in jetting fluidized beds and spout-fluid beds; rather, most investigations[16-19] focused on circulating fluidized beds. Xiao et al.[6] studied gas bypassing in a pressurized spout-fluid bed with a draft tube by injecting carbon dioxide into the spouting gas only. For jetting fluidized beds, Yang et al.[12,13] investigated gas interchange between the jet and the annular dense region of the semicircular jetting fluidized bed, in diameters of 3 cm and 3 m, by injecting the helium tracer into the fluidizing gas (no tracer gas was injected into the jet) or by injecting the helium tracer into the jet (no tracer gas was injected into the fluidizing gas). Gbordzoe et al.[14] studied the gas mixing in a large two-dimensional jetting fluidized bed using a V-grid air distributor and ozone as a nonreactive tracer gas. The effect of the fluidizing gas flow rate on gas transfer from the jet into the dense phase was discussed; the velocity of the jet was kept constant, whereas the fluidizing gas velocity varied from u_{mf} to $3.0u_{mf}$. Kimura et al.[15] investigated gas transfer from the jet to the annular region. Helium was used as a tracer gas and injected into the center jet nozzle. Although spout-fluid beds assemble many characteristics of jetting fluidized beds, the configurations and gas–solid interaction in the spout-fluid beds are somewhat different from those of jetting fluidized beds.[9,10] Very few data on gas mixing exist for this flow regime for the cross-flow of fluidizing gas to a submerged jet and vice versa. Thus, experimental approaches aiming at attaining more valuable information on gas mixing in spout-fluid beds are expected.

The study reported herein focuses on the experimental investigation of gas mixing in a spout-fluid bed by simultaneously injecting two different tracer gases into the bed and is centered on examining the effects of two important operating parameters—spouting gas velocity and fluidizing gas flow rate—on the gas mixing, and preliminarily discussing the gas-mixing mechanism in spout-fluid beds.

Experimental

Apparatus

The present experimental system is modified from a two-dimensional spout-fluid bed that was introduced in detail in our previous publications.[9-11] The experimental system is shown in Figure 1. The system consists of a spout-fluid bed column, a gas supply system, and measurement instruments. The column, made of 8 mm thick organic glass, has a 0.3×0.03 m rectangular cross section, height of 2 m, and width of 0.3 m. The spout nozzle is 0.03×0.03 m. A V-type gas distributor with a 60° angle of inclination was located at the bottom of the bed. The orifices in the gas distributor are 1 mm in diameter, and the total area of all orifices is 1.1% of the gas distributor.

A Roots-type blower supplied the spouting gas and the fluidizing gas. A pressure-reducing valve was installed to avoid pressure oscillations and to achieve steady flow. The gas flow rates were measured by two flow meters. The spouting gas entered the bed directly through the upward spout nozzle. The fluidizing gas, divided into two equal fluxes by a flux distributor and respectively supplied into the two gas chambers, then entered the bed by the orifices in the gas distributor.

Pressure fluctuations in the bed were obtained by a multichannel differential pressure signal sampling system. A digital camera and a digital video were used to photograph the flow patterns through transparent walls during the experiments. A pair of 500-W floodlights was used to enhance photo definition during photographing. Differential pressure signal processing and digital image analysis were used to determine the minimum spouting gas velocity u_{ms} and flow patterns.[10,11,20] The minimum spouting gas velocity is 37 m/s. Experimental conditions and particle properties studied in this work are listed in Table 1.

Tracer gas and detection system

Some previous investigations[6,12-19] injected one nonactive gas tracer (such as helium, carbon dioxide, ozone, or argon) to study the gas mixing. For jetting fluidized beds, generally, tracer gas was injected into the jet (no tracer gas was injected into the fluidizing gas) to study gas transfer from the jet to the annular dense region, or tracer gas was injected into the fluidizing gas (no tracer gas was injected into the jet) to study gas transfer from the annular dense region to the jet region.[12-15] Gas transfer from the jet to the annular dense region and that from the annular dense region to the jet cannot be detected simultaneously. Thus, the gas interchange is not very evident by injecting one tracer gas. To obtain a full knowledge of gas interchange, the above-mentioned processes might take at least double the cost of the experiments. In the present study, two different tracer gases were simultaneously injected into the bed. One was injected into the spouting gas and the other was injected into the fluidizing gas. Gas transfer from the spout jet

Table 1. Experimental Conditions and Particle Properties

Particles	d_p (mm)	SD (%)	ρ_s (kg/m^3)	ε	u_{mf} (m/s)	H_0 (m)
Mung beans	3.2	±10.3	1640	0.42	1.07	0.5

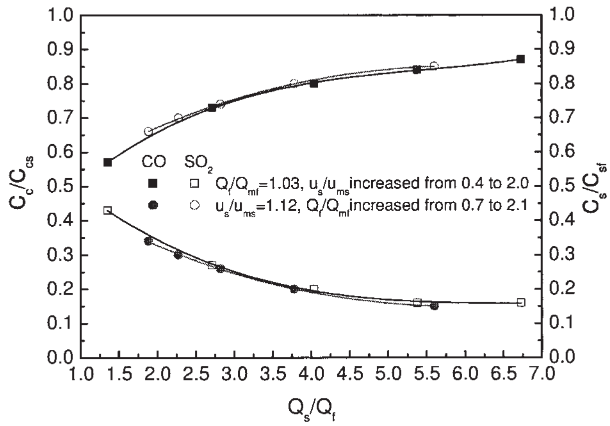

Figure 2. Dimensionless concentrations of tracer gas at complete mixing level.

to the annular dense region and that from the annular dense region to the spout jet can be realized simultaneously.

Carbon monoxide (CO) and sulfur dioxide (SO_2) were used as nonreactive tracer gases. Use of these two harmful gases is explained by the restriction of our gasometrical analysis apparatus, which is the most sensitive to these two gases. Although SO_2 has a higher molecular weight than that of either CO or air, the tracer gas profiles for the case when only SO_2 is injected into fluidizing gas are almost the same as those when only CO is injected into fluidizing gas. This implies that the molecular weight of SO_2 (64, which is about 2.5 times that of CO and air) would not influence the gas mixing. Undoubtedly, if possible, other nonreactive harmless tracer gases are more suitable. CO was injected into the spouting gas flow and SO_2 was injected into the fluidizing gas flow continuously and steadily. The tracer gas was mixed with either the spouting gas or the fluidizing gas. The concentrations of both tracer gases at the injection level are about 150×10^{-6} (150 ppm).

Steady-state tracer gas measurements were carried out to obtain radial tracer gas concentrations at various bed elevations. The tracer gases were sampled through a stainless steel tube (1 mm ID) at different elevations above the bottom of the column where several holes were drilled for measuring. The stainless steel tube port was covered with a screen to prevent the leakage of particles from the bed. At steady state, tracer gases were sampled at six horizontal locations for every elevation. The sampled tracer gases were analyzed by a gasometrical analyzer (NGA2000; Elementar, Hanau, Germany). Although six stainless steel tubes were located so as not to disturb the gas–solid flow, only one tube was inserted into the bed for each time. The CO reference concentration C_{cs} and the SO_2 reference concentration C_{sf} were obtained first by sampling the spouting gas and fluidizing gas, respectively. The dimensionless tracer gas concentrations, C_c/C_{cs} and C_s/C_{sf}, were used to analyze the gas mixing. All the data in the present work are the average values over a certain time. The time variation is 5 s. Dimensionless concentrations of tracer gas at theoretical complete mixing levels are shown in Figure 2.

Results and Discussion

The dimensionless CO concentration C_c/C_{cs} and SO_2 concentration C_s/C_{sf} in different bed locations were obtained. The mean range in the data values at a particular condition is about 7%. Because CO is only from the spouting gas flow and SO_2 is from the fluidizing gas, gas transfer from the spout region to the annular dense region and gas transfer from the annular dense region into the spout region can be shown clearly and simultaneously.

Tracer gas concentrations at different bed locations

Figure 3 plots the typical variations of dimensionless tracer gas concentration profiles measured at $u_s/u_{ms} = 1.12$ and $Q_f/Q_{mf} = 1.03$ ($Q_s/Q_f = 3.78$), where D represents the column width, which is 0.3 m. The radial distribution of C_c/C_{cs} decreases sharply, then gradually with increasing H/D, and finally reaches the average value of 0.79, which is in substantial agreement with the theoretical complete mixing value of 0.80, as presented in Figure 2. Unlike C_c/C_{cs}, the radial distribution of C_s/C_{sf} increases sharply, then gradually with increasing H/D, and finally reaches the average value of 0.21, which agrees with the theoretical value of 0.20. The mass balance closures imply that gas mixing has completed, that is, the spouting and fluidizing gas have been well mixed.

The axial distributions of C_c/C_{cs} and C_s/C_{sf} in the spout jet region ($r/R < 0.2$, determined by the snapshot of flow pattern) are different from those in the annular dense region. In the spout jet region, C_c/C_{cs} decreases with bed height, whereas C_s/C_{sf} increases. Conversely, in the annular dense region, $C_s/$

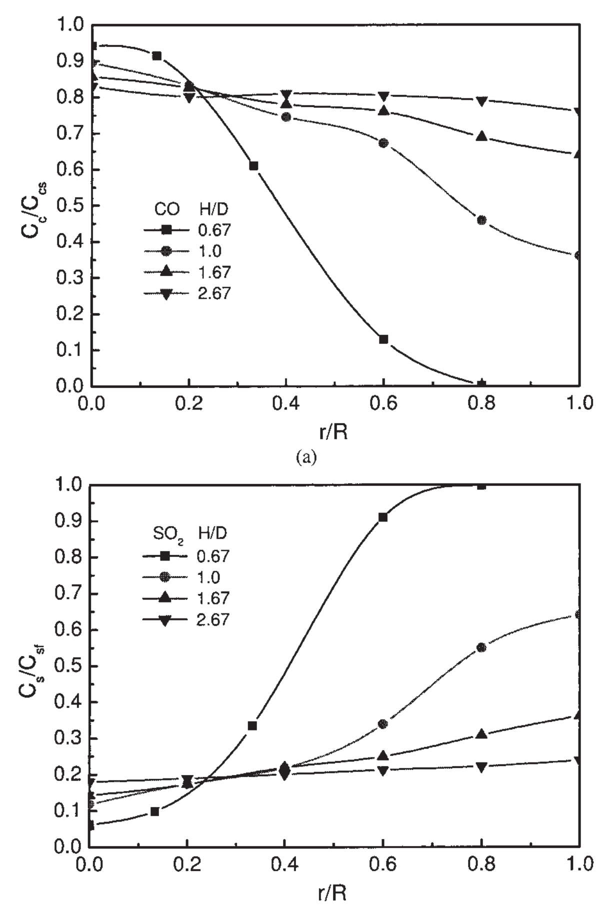

Figure 3. Varieties of tracer gas concentration with bed heights measured at u_s/u_{ms} = 1.12, Q_f/Q_{mf} = 1.03, Q_s/Q_f = 3.78.

C_{sf} increases with bed height, whereas C_c/C_{cs} decreases. The decay of C_c/C_{cs} in the spout region indicates that the fluidizing gas moves from the annular dense region into the spout jet region. This portion of fluidizing gas thus dilutes the CO concentration and densifies the SO_2 concentration in the spout jet, which can be observed by the presence of SO_2 and increase in C_c/C_{cs} in the spout jet region. On the other hand, the spouting gas also transfers from the spout jet region into the annular dense region, which can be seen by the presence of CO and the decrease in C_s/C_{sf} in the annular dense region. This portion of spouting gas dilutes the SO_2 concentration and densifies the CO concentration in the annular dense phase. The results show that the gas transfer between the central spout jet and the annular dense phase takes place synchronously.

Effect of fluidizing gas flow rate on gas mixing

Figure 4 presents the dimensionless tracer gas concentration profiles measured at different fluidizing gas flow rates when $u_s/u_{ms} = 1.12$. Taking the overview of these profiles, the radial distribution of C_c/C_{cs} decreases and C_s/C_{sf} increases with increasing fluidizing gas flow rate.

At $H/D = 0.67$, in the distributor region, C_c/C_{cs} decreases, whereas C_s/C_{sf} increases with increasing fluidizing gas flow rate. The fast decay of C_c/C_{cs} in radius indicates that a greater amount of SO_2 is carried into the bed by increasing fluidizing gas flow rate, which causes dilution of CO in both the annular region and the core region. On the other hand, when the fluidizing gas flow rate is increased from $1.37Q_{mf}$ to $2.07Q_{mf}$, the value of C_c/C_{cs} is almost zero, although C_s/C_{sf} is 1 at $r/R = 0.7$. This radial location is somewhat away from the orifices in the fluidizing gas distributor ($r/R = 0.86$). Thus, the increasing fluidizing gas flow rate might also weaken the transfer of spouting gas into the annular dense region in the distributor region. According to the flow phenomena observed in the experiments, particles in the annular region move down and are entrained into the spout region when they are in the bottom of the V-type gas distributor, then accelerate upward to form a fountain. The greater the fluidizing gas flow rate, the greater number of particles are entrained into the spout jet region, which can also be clearly seen in Figure 5. Figure 5 presents instantaneous snapshots of flow patterns obtained at $u_s/u_{ms} = 1.12$ and various fluidizing gas flow rates. It can be distinctly observed that particles mixing in the annular region and exchange between the spout region and the annular region become more and more intensive with increasing Q_f/Q_{mf}, especially in the V-type gas distributor region. The gas–particle flow becomes turbulent at a high fluidizing gas flow rate. These flow phenomena were also reported in our previous experiments using polystyrene as the bed material.[10,11] Because the movement of particles is mainly attributed to the gas flow, it can be inferred that the horizontal gas exchange is more active at a high fluidizing gas flow rate.

At $H/D = 1.0$, by increasing the fluidizing gas flow rate, the radial decrease in C_c/C_{cs} becomes sharp when $r/R < 0.6$ but gradual when $r/R > 0.6$. The radial SO_2 concentration profiles show converse trends. According to the gas–particle flow behaviors, when the fluidizing gas flow rate is $>1.3Q_{mf}$, the spout jet is submerged by the fluidizing gas, which is also shown in Figure 5. The spout jet disappears because the momentum of the spout jet is exhausted by increasing fluidizing gas; instead,

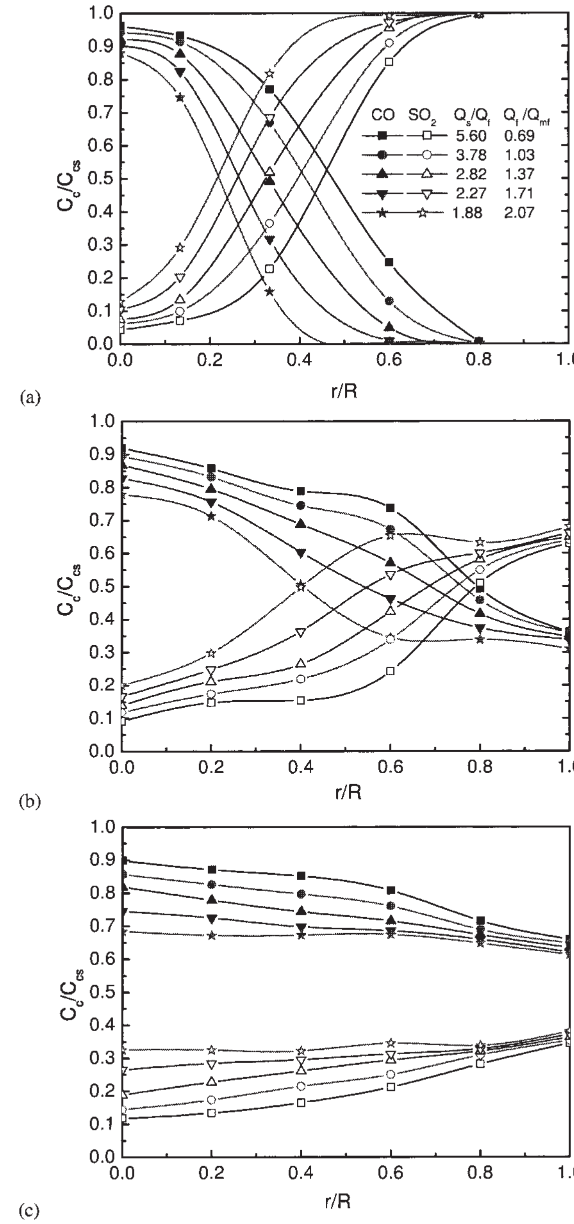

Figure 4. Tracer gas concentration profiles measured at three bed heights and different fluidizing gas flow rates.

(a) $H/D = 0.67$; (b) $H/D = 1.0$; (c) $H/D = 1.67$.

bubbles form in the central region and the annular region. In this case, spouting gas in the spout region easily diffuses into the annular region as a result of the increasing voidage in the annular region.[1-4,9-11] Bubbles that contain high SO_2 are entrained into the central region, mixing with the spouting gas. Vortices can be clearly seen in the boundary of the central dilute phase and the annular dense phase; the boundary is about at $r/R = 0.45$ for the current operating conditions. The gas exchange between these two phases becomes more intensive at a higher fluidizing gas flow rate, which leads to the rapid decay of C_c/C_{cs} and augmentation of C_s/C_{sf} at $r/R < 0.6$. For $r/R > 0.6$, almost no bubbles can be observed, especially near the wall layer and, as shown in Figure 5, particles move down

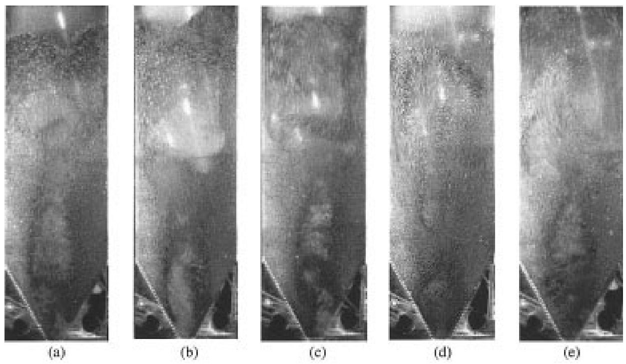

Figure 5. Instantaneous snapshots of flow patterns at u_s/u_{ms} = 1.12 and various fluidizing gas flow rates.

(a) Q_f/Q_{mf} = 0.69, Q_s/Q_f = 5.60; (b) Q_f/Q_{mf} = 1.03, Q_s/Q_f = 3.78; (c) Q_f/Q_{mf} = 1.37, Q_s/Q_f = 2.82; (d) Q_f/Q_{mf} = 1.71, Q_s/Q_f = 2.27; (e) Q_f/Q_{mf} = 2.07, Q_s/Q_f = 1.88.

steadily. In this case, a greater percentage of fluidizing gas penetrates into the annular dense region, and also a greater amount of spouting gas diffuses into this region as a result of the increasing voidage. This can be observed by comparing the tracer gas concentrations at H/D = 1.0 to that at H/D = 0.67 for the same operating condition. However, compared to the radial location at $r/R < 0.6$, the gas–particle interaction is less intensive at $r/R >$ 0.6, even if at considerable fluidizing gas flow rates. The mixing of spouting gas and fluidizing gas is less intensive at this region. Thus, the radial decrease in C_c/C_{cs} and increase in C_s/C_{sf} become slower with increasing fluidizing gas flow rate.

At H/D = 1.67, the radial C_c/C_{cs} and C_s/C_{sf} profiles become flat at large fluidizing gas flow rates. For Q_f/Q_{mf} values of 1.71 and 2.07, the radial C_c/C_{cs} reaches average values of 0.69 and 0.66, which is in good agreement with the theoretical complete mixing values of 0.70 and 0.66, respectively. At the same time, the average values of C_s/C_{sf} are also almost the same as the theoretical complete mixing values, which indicates that gas mixing has almost completed. Compared to the tracer gas profiles at a relatively low fluidizing flow rate, it is found that increasing fluidizing gas flow rate can contribute to the completion of gas mixing at a relatively low bed height. However, the C_c/C_{cs} is slightly lower and C_s/C_{sf} is slightly higher near the wall than at other radial locations, and the concentrations are somewhat higher than theoretical complete mixing values. As in the previous investigation,[17] the current results show that there is also gas transfer between the wall layer and the core region. However, there is some difficulty in achieving a satisfactory mixing condition in the wall layer even at a high fluidizing gas flow rate.

The above discussion shows that gas mixing in the spout-fluid bed can be promoted by increasing the fluidizing gas flow rate. This is attributed to the following two aspects: (1) the gas exchange with less resistance resulting from the increasing voidage in the annular region; (2) the intensive gas–particle action caused by entrainments, vortices, and bubbles. Some details in the literature confirm this aspect. Lim et al.[21] indicated heat transfer and mass transfer between gas and particles will be promoted with the supply of fluidizing gas, irrespective of whether the annular region is fluidized. Furthermore, the likelihood of particle agglomeration, dead zone, and sticking to the wall of the vessel can be effectively reduced, which suggests a better gas and particle mixing by increasing fluidizing gas flow rate. Other studies[1-4,9-11] also indicated that when the fluidizing gas flow rate increases, the voidage in the annular region increases, the spouting gas easily enters into the annular region, and exchanges of the mass and momentum between the spout region and annular region become more intensive. On the other hand, when the annular gas flow rate is beyond the minimum fluidizing gas flow rate, the excess gas forms more bubbles in the annular region.[1-4,9-11] The actions of bubbles can promote gas mixing.[15] After examining the effect of fluidizing gas flow rate on the gas mixing in a jetting fluidized bed, Gbordzoe et al.[14] indicated that when Q_f/Q_{mf} increases from 1.0 to 2.0, a profusion of fluidizing gas entered into the jet by convention, the dense phase in this case was well mixed, and bubbles appeared and were also entrained into the spout jet.

Effect of spouting gas velocity on gas mixing

The dimensionless tracer gas concentration profiles at Q_f/Q_{mf} = 1.0 and different spouting gas velocities are plotted in Figure 6. The flow patterns at these operating conditions are presented in Figure 7: the flow pattern transit from "internal jet" to "spouting" when the spouting gas velocity increases. As a whole, radial C_c/C_{cs} increases and C_s/C_{sf} decreases with increasing spouting gas velocity. However, compared to the tracer gas concentration profiles shown in Figure 4, the tracer gas concentration profiles presented in Figure 6 are obviously different.

At H/D = 0.67, in the spout jet region, C_c/C_{cs} decreases and C_s/C_{sf} increases slightly when u_s/u_{ms} is increased from 0.4 to 2.0, which implies that a greater amount of fluidizing gas is entrained into the spout jet region, thus diluting the CO concentration in this region. However, more CO is introduced into the bed with increasing spouting gas velocity, and the decay of C_c/C_{cs} and the augmentation of C_s/C_{sf} are slow, especially at high spouting gas velocities. In the annular region, C_c/C_{cs} increases and C_s/C_{sf} decreases rapidly with increasing spouting gas velocity, which indicates that gas transfer from the spout jet region increases. By increasing the spout gas velocity, the spout jet momentum flow rate increases, which leads to an increase of penetration ability of the spout jet.[10] When the spout jet ascends along the axis, more spouting gas will penetrate into the annular region.

The increasing penetration ability of spouting gas in the annular region is also the main reason that radial C_c/C_{cs} increases and C_s/C_{sf} decreases rapidly at H/D = 1.0. This also implies that a greater degree of gas mixing has taken place at this bed height in the flow pattern of spouting than internal jet. Similar to results at H/D = 1.0, C_c/C_{cs} increases and C_s/C_{sf} decreases at H/D = 1.67 with increasing spouting gas velocity. At high spouting gas velocities (u_s/u_{ms} values of 1.6 and 2.0), more particles are entrained into the central region and vortices are clearly shown in the boundary of the central dilute phase and the annular dense phase, as shown in Figure 7. The gas mixing can be promoted as a result of the entrainments and vortices. However, the tracer gas profiles obtained by increasing spouting gas velocity are not flat at H/D = 1.67, even if at high spouting gas velocities. C_c/C_{cs} is lower and C_s/C_{sf} is higher near the wall than at other radial locations, and are far from the theoretical complete mixing values, as presented in Figure 2. The results show that there is also gas transfer between the wall layer and the core region, whereas there is

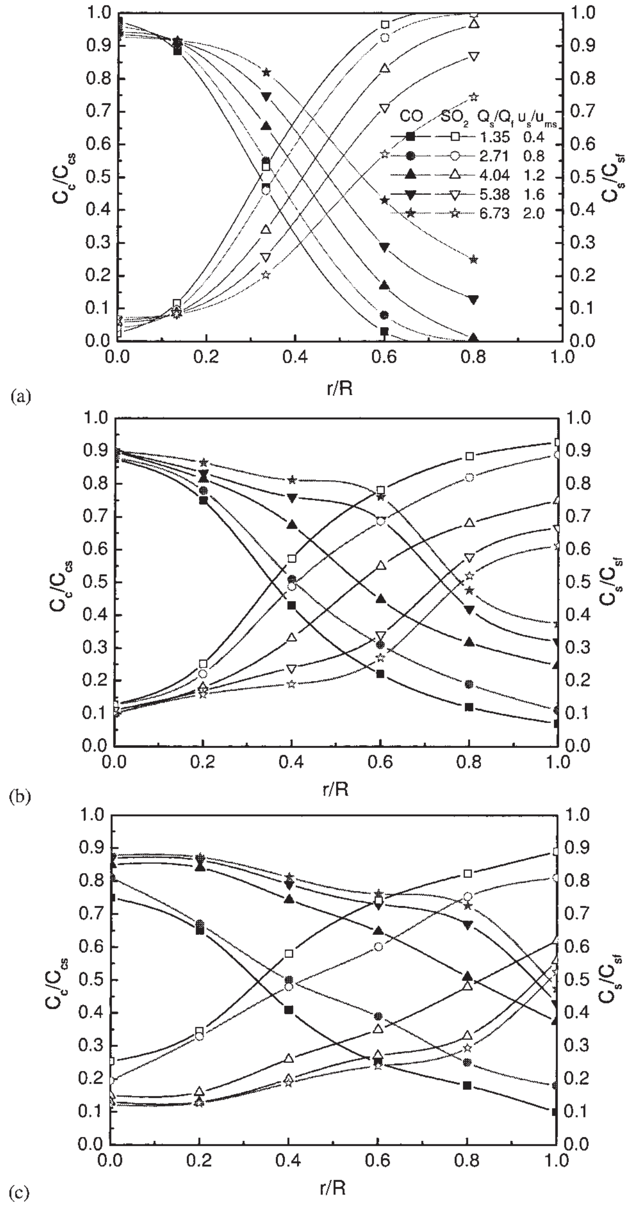

Figure 6. Tracer gas concentration profiles measured at Q_f/Q_{mf} = 1.03 and different spouting gas velocities.

(a) H/D = 0.67; (b) H/D = 1.0; (c) H/D = 1.67.

difficulty in attaining a satisfactory mixing condition in the wall layer even if at high fluidizing gas flow rates.

The above discussion shows that the increase in spouting gas velocity can promote gas mixing in spout-fluid beds, the main contributions to which are an increase in the penetration ability of spouting gas, entrainments, and vortices. For a given fluidizing gas flow rate, the penetration ability of the spout jet increases with increasing spouting gas velocity[9] and the spouting gas easily enters into the annular region when the spout jet ascends in the bed[1,2]: this might promote the gas exchange between the spout region and the annular region.[6,15] On the other hand, when the spouting gas velocity increases, the momentum of spouting gas increases and the entrainment of gas and particles from the annular region by spout over its entire height is promoted.[3,4,6] Gas–particle movements are turbulent in the boundary of the annular region and vortices are clearly seen in the spout region. These phenomena were observed in the present experiments and recorded by a high-resolution CCD camera in our previous experiments.[9-11]

However, in a comparison of the tracer gas profiles obtained by increasing spouting gas velocity to those by increasing the fluidizing gas flow rate, it is found that (1) the gas mixing in the spout-fluid bed is more effective by increasing the fluidizing gas flow rate and (2) the gas mixing could complete at a relatively low bed height by increasing fluidizing gas flow rate. Behie et al.[22] studied the gas transfer from a jet into a fluidizing bed and concluded that the mass transfer from a high-velocity jet into a fluidizing bed is not very good. He thus recommended that a jetting fluidizing bed should not be used to enhance the mass transfer by increasing the jet velocity.

Preliminary discussion of gas-mixing mechanism

The gas mixing mechanism in spout-fluid beds is complex and has not yet been fully elucidated, although the following discussion might be helpful. The previous investigation on jetting fluidized bed[14] indicated that diffusive transport of tracer gas could be considered negligible compared with convective transport. Similar conclusions have been drawn by Yang et al.[12] Yang[13] also investigated the gas interchange between the jet and the annular dense region of semicircular jetting fluidized beds in 3-cm and 3-m diameters. By numerically solving the tracer gas conservation equation, it was emphasized that the gas mixing was primarily a result of convection, and diffusion plays a negligible role. For spout-fluid beds, the flow behaviors for the cross-flow of fluidizing gas to a submerged jet and vice versa are similar to jetting fluidizing beds. Undoubtedly, convection plays an important role in the gas mixing in spout-fluid beds.

However, compared to jetting fluidizing beds, there is a stronger production of vortices and the bubbles in the fluidized part of the spout-fluid bed, thus giving rise to a diffusive transport that could not be convective. Moreover, the gas–solid motions are turbulent at high spouting gas velocity or fluidizing gas flow rate, as shown in Figures 5 and 7. More particles and

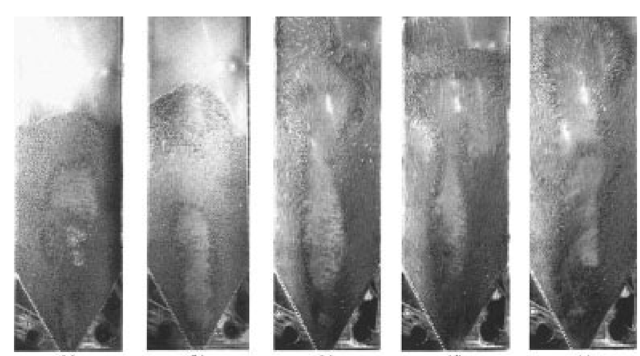

Figure 7. Instantaneous snapshots of typical flow patterns at Q_f/Q_{mf} = 1.03 and various spouting gas velocities.

(a) u_s/u_{ms} = 0.40, Q_s/Q_f = 1.35; (b) u_s/u_{ms} = 0.80, Q_s/Q_f = 2.71; (c) u_s/u_{ms} = 1.20, Q_s/Q_f = 4.04; (d) u_s/u_{ms} = 1.60, Q_s/Q_f = 5.38; (e) u_s/u_{ms} = 2.0, Q_s/Q_f = 6.73.

gas (or bubbles) are entrained into the spout jet region, and more spout jet gas transfers into the annular region at the same time. Gbordzoe et al.[14] also reported the turbulent transport phenomena and defined them as being attributed to convection. However, the turbulent transport should not be classified as a convection phenomenon, but rather as a diffusion phenomenon. This concept indicates that diffusion also contributes to the gas mixing in spout-fluid beds.

In summary, gas mixing in spout-fluid beds is attributed to both convection and diffusion. Diffusion other than molecular diffusion should not be neglected, especially at high spouting gas velocity or fluidizing gas flow rate. However, the complexity of gas mixing in spout-fluid beds clearly requires further experimental and theoretical study.

Conclusions

Gas mixing in a cold model of a spout-fluid bed coal gasifier was investigated by concurrently injecting two different tracer gases into the bed. CO was injected into the spouting gas flow and SO_2 was injected into the fluidizing gas flow. Gas transfer from the spout jet to the annular dense region and that from the annular dense region to the jet were detected simultaneously. Thus, gas interchange between the spout region and the annular region is evident by injecting two different tracer gases.

The effects of two important operating parameters—spouting gas velocity and fluidizing gas flow rate—on the gas exchange between the spout region and the annular dense region at different bed locations were discussed with flow patterns. The results show that increases in spouting gas velocity and fluidizing gas flow rate can both promote gas mixing in spout-fluid beds. Increasing the fluidizing gas flow rate is the more effective way, given that a satisfactory mixing condition at a relatively low bed height can be achieved by such an increase. For both cases, it is difficult to obtain a satisfactory mixing condition in the wall layers. Moreover, the results of gas-mixing mechanisms, preliminarily discussed, imply that gas mixing in the spout-fluid bed is caused by both convection and diffusion. Diffusion other than molecular diffusion should not be neglected, especially at high spouting gas velocity or fluidizing gas flow rate.

Acknowledgments

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Notation

C_c = local tracer gas CO concentration, 10^{-6}
C_{cs} = tracer gas CO concentration at injection level, 10^{-6}
C_s = local tracer gas SO_2 concentration, 10^{-6}
C_{sf} = tracer gas SO_2 concentration at injection level, 10^{-6}
D = bed diameter, m
d_p = particle diameter, m
f = sampling frequency, Hz
H_0 = static bed height, m
H = bed height, m
Q_f = fluidizing gas flow rate, m^3/s
Q_s = spouting gas flow rate, m^3/s
Q_{mf} = minimum fluidizing gas flow rate, m^3/s
r = local position in radius, m
R = radius of the vessel, m
t = pressure time series sampling time, s
u_s = spouting gas velocity, m/s
u_{ms} = minimum spouting gas velocity, m/s
u_{mf} = minimum fluidizing gas velocity, m/s
ε = bulk voidage of particles
ρ_s = particle density, kg/m^3

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