Using Stereo XPTV to determine Cylindrical Particle Distribution and Velocity in a Binary Fluidized Bed

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Keywords
binary fluidized bed, cylindrical particle fluidization, particle velocity, X-ray particle tracking velocimetry

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Using Stereo XPTV to determine Cylindrical Particle Distribution and Velocity in a Binary Fluidized Bed

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Abstract

Non-spherical particles are commonly found when processing biomass or municipal solid waste (MSW). In this study, cylindrical particles are used as generic non-spherical particles and are co-fluidized with small spherical particles. X-ray particle tracking velocimetry (XPTV) is used to track the 3D particle position and velocity of a single tagged cylindrical particle over a long time period in the binary fluidized bed. The effects of superficial gas velocity ($u_f$), cylindrical particle mass fraction ($\alpha$), particle sphericity ($\Phi$), and bed material size on the cylindrical tracer particle location and velocity are investigated.

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Keywords

binary fluidized bed, cylindrical particle fluidization, particle velocity, X-ray particle tracking velocimetry
Introduction

Using municipal solid waste (MSW) and biomass as alternative energy sources to produce electricity and heat has drawn great attention around the world, as it not only reduces the consumption of fossil fuels, but it also offers a solution for solid waste disposal. Combustion, pyrolysis, and gasification are the most common ways of producing electricity from MSW and biomass, and fluidized beds are ideal equipment to carry out these thermochemical processes because of its efficient heat and mass transfer, uniform temperature distribution, and sufficient gas-solid contacting. The shape and size of MSW or biomass particles may vary significantly, making them difficult to be fluidized. To improve the fluidization of this non-spherical material, inert bed material like silica sand, must be added to assist the fluidization process. Thus, fluidization of MSW and biomass in a fluidized bed is significantly different from conventional fluidization because it is multi-component fluidization involving non-spherical particles of various size, shape, and density, which result in phenomena such as segregation, bridging, and tunneling, and complicates the fluidization process.

Several experimental and numerical simulation studies have explored the co-fluidization of non-spherical particles with inert bed material, especially for the cylindrical particles, which is a common shape of biomass and preprocessed MSW. Most of these studies focused on equipment scale fluidization properties such as bed pressure drop \(^1\), minimum fluidization velocity \(u_{mf}\)\(^2-7\), flow regimes identification and transition \(^5,8-11\), and segregation and mixing \(^9,12,13\). However, non-spherical particle scale behavior such as particle distribution, velocity, and orientation still remain incomplete because of the lack of experimental data. Vollmari et al. \(^14\) performed an experimental and numerical investigation on cylindrical particle fluidization. Particle tracking velocimetry (PTV) and digital image analysis (DIA) methods were used to
obtain the position and orientation of cylindrical particles in the near-wall region. Simulation and experimental results showed good agreement at lower superficial gas velocities, but poor agreement at high superficial gas velocities.

Similar research was carried out by Cai et al. \cite{15,16} in a CFB (Circulating Fluidized Bed), where the cylindrical particle distribution and orientation was roughly estimated from visual observations and then compared with the detailed particle motion behavior obtained from numerical simulations. Results indicate that the orientation and position of the cylindrical particles was highly correlated. Fotovat et al. \cite{17,18} investigated the co-fluidization of cylindrical and spherical particles with inert bed material using a radioactive particle tracking (RPT) method. Particle distribution and circulation behavior were obtained and discussed. Oschmann et al. \cite{19} studied the mixing and orientation of different non-spherical particles in a fluidized bed using CFD-DEM numerical simulations. They investigated the dependency of mixing and particle orientation on particle properties, operational parameters, and simulation parameters. Results showed that increasing the superficial gas velocity enhanced the mixing of non-spherical particles. Soria-Verdugo et al. \cite{20,21} studied the circulation behavior of a large object immersed in a pseudo 2D bubbling bed and developed a circulation time model to estimate particle residence time in different bed regions. Tavassoli et al. \cite{22} studied the particle-fluid heat transfer for non-spherical particles using DNS (Direct Numerical Simulation) and IBM (Immersed Boundary Method) method. Results indicated that the heat-transfer correlation of spherical particles could be used for spherocylinders particles under some circumstances. Gui et al. \cite{23} studied the particle-particle collision in swirling jets with DNS and DEM (Discrete Element Method) method and they found out that collision probability was positively correlated to the normal components of Reynolds stress tensor. Das et al. \cite{24} studied
flow and heat transfer of randomly packed spherical particles in slender fixed-bed with DNS method, and models of pressure drop and Nusselt for wall-to-bed heat transfer were proposed. Zgheib et al. studied the cylindrical particle-laden gravity currents using DNS method. The results showed that vortex structures have a strong influence on wall shear-stress and deposition pattern. Besides, Vincent et al., Bordoloi and Variano, and Wang et al. carried out similar research with numerical simulation method as well. Chan et al. studied the particle velocity and residence time distribution in CFB and standpipes using PEPT (Positron Emission Particle Tracking) method. Models of residence time and particle velocity were proposed based on the experimental data. Mahmoudi et al. carried out research on overall solids residence time and its distribution in the riser of a CFB with RPT method. Slip factors for the different flow modes are determined. Balakin et al. performed investigation on the particle motion in hydrocyclone and pneumatic system using PEPT method. Experimental results indicated that PEPT could visualize the behavior of both the carrier and dispersed phase in the two equipments. Most of the studies on particle scale behavior are based on numerical simulation, visual observation in the multi-component fluidized bed near-wall region, or pseudo 2D fluidized bed visualization because it is too difficult to quantify the particle motion in the opaque interior region of the binary fluidized bed. RPT and PEPT can overcome the problem, but both the two methods are expensive and need radioactive material.

X-ray based multiphase flow measurement methods are more promising compared to conventional visible light based methods because X-rays can easily penetrate the dense fluidized bed region. Using X-ray stereography imaging, it is possible to reconstruct the three dimensional position, orientation, and velocity of a tagged tracer particle. In this study, cylindrical particles are treated as a typical
non-spherical particle and co-fluidized with spherical particles. A single cylindrical particle is removed from the bed, tagged so it can be identified with X-rays, and then reinserted into the binary bed. Then, X-ray Particle Tracking Velocimetry (XPTV), following the method reported by Heindel et al.\textsuperscript{34,35} is used to determine the 3D position and velocity of the tagged tracer particles across the whole bed. Effects of operating conditions (e.g., superficial gas velocity ($u_f$), particle sphericity ($\Phi$), particle mass fraction ($\alpha$), and bed material particle size) on 3D position and velocity are reported.

**Experimental Procedures**

**Experimental setup**

Figure 1a depicts a schematic representation of the experimental setup. The two-component fluidization tests are carried out in a transparent cylindrical fluidized bed made of Plexiglas with internal diameter $D = 100$ mm and total height $H = 900$ mm. Henceforth, the two-component fluidized bed is referred to as a binary fluidized bed. A perforated plate type distributor is installed at the bottom of the bed, above the gas chamber that is connected to the air supply system. The distributor, with total open area of 0.62\%, contains 62 holes of 1 mm diameter, and a mesh screen is attached to the distributor to prevent bed material from blocking the aeration holes. The distributor pressure drop is much higher than the bed pressure drop so that the plenum-bed coupling can be avoided. The gas chamber beneath the distributor is filled with marbles to provide a uniform gas pressure before entering the bed. The compressed air comes from the laboratory air supply system and is regulated by an automatic mass flow control system, which can measure, adjust, and record the gas flow rate. The air is humidified by a water tank before entering the fluidized bed to prevent static electricity buildup. Bed pressure drop is measured with a wall-mounted
pressure transducer located above the distributor. The signals obtained from the pressure transducer are processed and stored by a computer controlled data acquisition system. In order to reduce the bed pressure error caused by pressure fluctuations, the pressure signal is sampled at a frequency of 1000 Hz for 32 seconds and then averaged as the final measurement. Both the original data and averaged data are recorded for further analysis. More details about the fluidized bed system are available in Escudero and Heindel

The X-ray stereography imaging system, shown in Figure 1b, consists of two identical X-ray source/detector pairs, which can capture the X-ray radiographic projections simultaneously from two perpendicular directions. The 3D position \( P(x, y, z) \) of the tracer particle can be determined by solving the function set (1) based on the geometrical model in Figure 1b. The function set is overdetermined, thus it cannot be solved directly, but the least-squares solution of the function set can be obtained using a pseudo inverse matrix. More details of the position/velocity measurement and the fabrication of tracer particle can be found in the supplemental material and literature

\[
\begin{bmatrix}
b - a & -x_i & 0 \\
z_1 & 0 & -x_i \\
0 & z_1 & a - b \\
y_2 & a - b & 0 \\
z_2 & 0 & a - b \\
\end{bmatrix}
\begin{bmatrix}
x \\
y \\
z \\
\end{bmatrix}
=
\begin{bmatrix}
-ax_i \\
0 \\
a z_1 \\
y_2 \\
0 \\
\end{bmatrix}
\]

(1)

Particle and bed material properties

Two different lengths of cylindrical particles made of ABS plastic are used in this work. These two cylindrical particles have different diameters but the same volume equivalent diameter \( d_v = 6 \) mm; hence, the particles have different sphericities \( \Phi \), but the shape factor is univariate. In order to evaluate the effect of cylindrical particle
anisotropic hydrodynamic characteristic on particle behavior, a spherical particle with
the same volume and material is used as well. More information about the cylindrical
and spherical particles can be found in Table 1, where Asr is the particle aspect ratio
($l/l_d$) and $\varepsilon_0$ is the particle packing voidage. One may note that the densities of
cylindrical particles and spherical particle are slightly different. This is due to the
small material variation of the cylindrical and spherical particles, but their densities
are still comparable with each other.

Two different diameters of glass beads are used as the spherical inert bed material
in this work and detailed properties of each glass bead system are found in Table 2.
Note that the minimum fluidization velocity ($u_{mf}$) reported in Table 2 was
experimentally determined for the respective single component fluidized bed.

Results and Discussion

Cylindrical or spherical particles and bed material are premixed in desired mass
fractions before being poured into the fluidized bed. In order to evaluate the effect of
cylindrical/spherical particle mass fraction $\alpha$ on fluidization, mixtures of different
mass fractions but with the same static bed height $H_0$ (i.e., the same apparent volume)
are used, and the mass of the cylindrical/spherical particles and bed material for
different mixtures are calculated according to non-spherical particle packing studies
37-42. Each cylindrical/spherical particle type (CA, CB, and SA) are co-fluidized with
one of two types of bed material (BG300-500 and BG800-1000, respectively) with
particle mass fractions $\alpha = 0\%, 10\%, 30\%$ and $50\%$ at static bed heights of $H_0 = 100$
mm ($H_0/D = 1$); thus, a total of 24 operating conditions are investigated. The range of
superficial gas velocity $u_f$ tested is $0\text{-}1.2$ m/s, which is well beyond the minimum
fluidization velocity of the glass bead material (Table 2). Many investigations
1,2,5,9,43,44 have proven that for binary and multi-component mixture fluidization, the
fluidization characteristics (e.g., flow regime transition, bed pressure drop, and minimum fluidization velocity $u_{mf}$) are different when increasing or decreasing the fluidization velocity, and the results from decreasing the velocity are more repeatable \textsuperscript{1,9,10}. Therefore, all experiments in this research are carried as the fluidization velocity decreases, which means that $u_f$ is first rapidly increased to 1.2 m/s, allowed to operating at this velocity for 15 minutes, and then $u_f$ was set to the next testing velocity and kept for 15 minutes. After this, the fluidization became stable and the pressure data was measured and recorded for 32 s. The process was repeated until $u_f = 0$. Basic fluidization experiments focused on minimum fluidization velocity $u_{mf}$ and bed pressure drop were carried out first, and these results can be found in the supplemental material.

Distribution of cylindrical particles

Knowing the cylindrical particle distribution is vital for the efficient operation of the binary fluidized bed because the heat and mass transfer properties and product composition can vary considerably with particle distribution. Several studies have assessed particle distribution using a tracer particle method \textsuperscript{15,18,20,21,45,46}. In this paper, X-ray particle tracing velocimetry (XPTV) is used to investigate the cylindrical particle distribution characteristics during its co-fluidization with bed material.

The ideal tracer method would be to label every cylindrical particle and track all of them at the same time. However, according to the measuring principle described in previous section, it is impossible to determine the accurate 3D position of multiple tracer particles if two or more tracer particles overlapped in each X-ray projection. Thus, only one cylindrical particle in the binary fluidized bed is tagged as the tracer particle. In this way, the vital problem of determining particle distribution is how to assess the distribution of all cylindrical particles by observing the motion of only one.
cylindrical tracer particle. Previous researchers have solved this problem by using the concept of ergodicity.\textsuperscript{18,21,45}

According to the concept of ergodicity, if the binary fluidized bed is observed for a sufficiently long time, the time average motion properties of one particle can represent the population average of all similar particles. In other words, if the position of only one cylindrical tracer particle among the entire cylindrical particle population is observed and recorded for a sufficiently long time during stable fluidization, its time averaged position distribution is comparable to the instantaneous position distribution of all cylindrical particles. Fotovat et al.\textsuperscript{18} and Buist et al.\textsuperscript{45,46} applied this theory using radioactivity particle tracking (RPT) and magnetic particle tracking (MPT) in a non-spherical particle fluidization. The challenge with the ergodicity assumption is deciding the total number of observations $N$ that the particle must be observed to define a long enough time. The contour map in Figure 2 shows the tracer particle time averaged position distribution derived from different total number of observations $N$. The horizontal axis is the dimensionless radial coordinate $r/R$, where $r$ is the horizontal distance between the tracer particle centroid and bed central axis and the $R$ is the bed radius. The vertical axis is the tracer particle height $h$ from the surface of the gas distributor. The colors in the contour plots represent the distribution probability of the tracer particle in the corresponding position, and it is calculated by dividing the number of observations that the tracer particle appears in the ring determined by the given $r$ and $h$, by the total number of observations $N$.

Curve in Figure 2 shows the effect of the total number of observations $N$ on the sum of the squared error (SSE) at every position for a binary bed composed of the CB particles and GB300-500 bed material, with the cylindrical particle mass fraction $\alpha = 10\%$ and fluidization gas velocity $u_f = 0.65$ m/s. The position of the tracer particle is
captured for \( N = 100, 500, 1000, 2000, 3000, 4000, 5000, 6000, 7000, 8000, 9000 \) and 10000 observations. The vertical axis represents the SSE between each observation window and the horizontal axis identifies the given observation windows. For example, \( N_{1000-2000} \) compares the data between \( N = 1000 \) and \( N = 2000 \) and the corresponding vertical axis value of 0.4 is the SSE between the respective data sets by the following equation

\[
\text{SSE} = \sum_{i=1}^{K} (C_{i,N=1000} - C_{i,N=2000})^2,
\]

where \( i \) is the identifier of position on the distribution contour plots, \( K \) is the total number of positions, which is 20000 in this study (100 in radial direction and 200 in height, \( K = 100 \times 200 \)). \( C_{i,N=1000} \) and \( C_{i,N=2000} \) is the distribution probability of the tracer particle on position \( i \) when \( N = 1000 \) and \( N = 2000 \), respectively. Figure 2 shows that the total number of observations \( N \) greatly affects the distribution results, but as \( N \) increases, the distribution results become similar (e.g., when \( N > 10000 \), the difference in SSE is very small). All subsequent tests for all other conditions were carried out with \( N = 11000 \) (about 10 minutes of data collection) to ensure a converged result.

Figure 3 shows the CA and CB particle distribution contour maps and associated flow regime photos when using bed material GB300-500 with \( \alpha = 10\% \). The gas velocity is \( u_f = 0.16 \text{ m/s}, 0.40 \text{ m/s}, \) and 0.65 m/s. If the minimum fluidization gas velocity \( u_{mf} \) of the bed material GB300-500 is used as the reference, the corresponding fluidization numbers are \( u_f/u_{mf} = 1.5, 3.6, \) and 5.9. Overall, the concentration of cylindrical particles is higher near the bed wall than in the center region. This phenomenon indicates two things. First, at any instant the cylindrical particles tend to accumulate in the near-wall region. This may create non-uniform heat and mass transfer between the bed material and cylindrical particles in this region. Second, the
resident time of the cylindrical particles in the near-wall is higher than in the central region. This is caused by fast moving large bubbles in the central region quickly transporting the cylindrical particles through this region and creating a slow moving solids packed region near the wall. Similar large scale motion was found for single component fluidized beds using X-ray computed tomography. 

When \( u_f \) is relatively low (e.g., \( u_f = 1.5u_{mf} \) (0.16 m/s)), CA particles mainly spread vertically between the range \( h = 10 - 125 \) mm, while CB particles distribute \( h = 25 - 110 \) mm. This indicates that when \( u_f \) is relatively low, cylindrical particles are prone to distribute in the upper part of the bed and the maximum attainable depth is lower (i.e., only a small fraction of the cylindrical particles reach the bottom of the bed). The horizontal distributions of cylindrical particles are non-uniform as well, especially for the distribution of CB particles which formed two vertical bands near the wall. It is necessary to clarify that at low \( u_f \), a high concentration of cylindrical particles is visually observed near the bed bottom, which is a static packing layer and captured in the accompanying flow regime photo. This was also observed by Chen et al. in a similar binary system.

When \( u_f \) is relatively high (e.g., \( u_f = 3.6u_{mf} \) (0.40 m/s)), CA particles mainly distribute vertically between the range \( h = 0 - 175 \) mm, while CB particles distribute \( h = 0 - 150 \) mm. When \( u_f \) increases further to \( u_f = 5.9u_{mf} \) (0.65 m/s), CA particles mainly distribute vertically between the range \( h = 0 - 200 \) mm, while CB particles distribute \( h = 0 - 175 \) mm. This indicates that increasing \( u_f \) increases the range in which cylindrical particles are found. The maximum attainable depth of cylindrical particles is positively related to \( u_f \). The horizontal distribution of cylindrical particles also becomes less concentrated in the near-wall region as \( u_f \) increases. Note that in Figure 3b, a cone-shaped region with low particle concentration can be observed near
the distributor. Compared to the photograph of the same operating condition, and 
based on visual observations during bed operation, a static region of CB particles are 
observed in this cone-shape region.

According to previous research, particles in multi-component fluidized beds can 
be categorized into jetsam or flotsam. Jetsam usually tends to sink and 
accumulate at the bottom of the bed while flotsam tends to float on the upper layer of 
the bed. Furthermore, the mixing conditions of jetsam and flotsam can be classified 
into three types: complete mixing, complete segregation, and partial mixing. In 
complete mixing, jetsam and flotsam are uniformly mixed with each other across the 
bed. In complete segregation, jetsam totally accumulates at the bottom of the bed and 
flotsam is fluidized above the jetsam. In partial mixing, part of the jetsam accumulates 
at the bottom of the bed and forms a static layer, other jetsam is fluidized and 
uniformly mixed with flotsam above the static layer. Particles with greater density 
usually behave like jetsam and lighter particles behave like flotsam. In the current 
research, CA particle fluidization show characteristics of complete mixing for all 
conditions in Figures 3a, 3c, and 3e. However, for CB particle fluidization, both 
operating conditions in Figure 3b and 3d reveal characteristics of partial mixing. 
When $u_f$ increases further to $u_f = 5.9u_{mf}$ (0.65 m/s) in Figure 3f, the mixing condition 
improves, implying an increase of $u_f$ can enhance the mixing of jetsam and flotsam. 
Although the density of the cylindrical particles is less than that of the bed material, 
CB particles obviously behave like jetsam, which is caused by its large size and 
anisotropic shape, similar to the classic segregation problem called “the Brazil nut 
effect”. The formation mechanism of the cone-shaped packing layer in Figure 3b 
will be discussed in the next section.

Figure 4 shows the cylindrical particle distribution of CB particles ($\Phi = 0.69$) at
(a) $\alpha = 30\%$ and (b) $\alpha = 50\%$ with $u_f = 0.65$ m/s ($5.9u_{mf}$) with bed material GB300-500. The effect of cylindrical particle mass fraction $\alpha$ on particle distribution characteristics can be obtained by comparing Figure 4 with Figure 3f. For $\alpha = 10\%$ and 30\%, no particle packing is observed in the associated flow regime photographs and particle distribution results, which implies that the mixture is in a state of complete mixing. Comparing the particle distribution maps in Figure 4a and 3f, the horizontal distribution of the cylindrical particles is similar. In the vertical direction, however, the cylindrical particle distribution for $\alpha = 30\%$ at the bottom of the bed is greater than that of $\alpha = 10\%$, which implies that increasing $\alpha$ encourages cylindrical particle accumulate in the bottom bed region. When $\alpha = 50\%$ (Figure 4b), a static packing layer is observed both in the flow regime photograph and the particle distribution contour map, indicating that the mixing state is in the partial mixing regime. Thus, increasing cylindrical particle mass fraction leads to a change in mixing state from complete mixing to partial mixing. What needs to be explained is that some high concentration spots can be found in Figure 4b, which seems to indicate that sometimes tracer particle got stack. However, the analysis of tracer particle trajectory showed that tracer particle hardly got stack in stable fluidization as long as it did not fall down into the bottom part (i.e. the static layer) during the $u_f$ decreasing process. The velocity of the tracer particle may become slower and its trajectory became twisted when touching the bottom part (i.e. the surface of the static packing layer), but it hardly stopped moving and would eventually move back to the fluidization part. Tracer particle also behaved in the same way in Figure 8.

Figure 5 shows the particle distribution of SA particles ($\Phi = 1.00$) with $\alpha = 10\%$, $u_f = 0.40$ m/s ($3.6u_{mf}$), and bed material GB300-500. The effect of particle shape on particle distribution characteristics can be determined by comparing Figure 5 with
Figures 3c and 3d. All three conditions are operating in the state of complete mixing. The horizontal distributions of all the three materials are similar. The only difference is that the CB particles have a slightly higher concentration in the near-wall region, which implies that particles migrate to the bed wall as $\Phi$ decreases. In the vertical direction, cylindrical particle concentration near the bed bottom increases as $\Phi$ decreases; this is emphasized by the observed small dead zones by the distributor plate for the CB particle system. The results in Figure 3 agree well with the numerical simulation reported by Hernandez-Jimenez et al.\textsuperscript{52}, which indicates this research may be interesting for the numerical simulation community.

Figures 6a and 6b show the distribution of CA and CB cylindrical particles with GB800-1000, $u_f = 0.65$ m/s ($1.2u_{mf}$), and $\alpha = 10\%$, respectively. It is necessary to point out that the fluidization number $1.2u_{mf}$ here is determined based on the $u_{mf}$ of the GB800-1000 bed material. Compared with the particle distribution in Figures 3e and 3f at the same $u_f$ and $\alpha$ but with smaller bed material GB300-500, a static packing layer is not observed in any of the four operating conditions, which means all the beds are in the state of complete mixing. For the vertical distribution, the cylindrical particles with GB300-500 cover a larger region than the GB800-1000 systems. The reason for this is that the larger bed material particles require a higher superficial gas velocity to fluidize the bed due to its greater inertia, thus the fluidization intensity of the cylindrical particles with GB800-1000 is lower if $u_f$ is the same for both bed materials. For the horizontal distribution, results with GB300-500 and GB800-1000 both show similar profiles, but the cylindrical particle distribution gradient from the bed wall to the center with GB800-1000 is greater than that of GB300-500. This phenomenon indicates that the increase of the bed material size will inhibit the dispersion of cylindrical particles when $u_f$ remains fixed. This may also result from a
decrease in the fluidization intensity. However, the situation changes when the fluidization number is used as a reference. Figures 6c and 6d show the distribution of CA and CB cylindrical particles with GB800-1000 at \( u_f = 0.90 \) m/s \((1.7 u_{mf})\) and \( \alpha = 10\% \), respectively. These figures are compared to the particle distributions in Figures 3a and 3b with GB300-500 for the same \( \alpha \) and similar fluidization numbers. It is obvious that the increase in bed material size can remarkably enhance particle mixing and cylindrical particle dispersion if the fluidization numbers are similar.

**Velocity of cylindrical particles**

In addition to particle location, XPTV can provide particle motion information if particle locations can be identified in successive frames with a known time interval. In this section, the motion characteristics of cylindrical particles are discussed. The cylindrical particle trajectory properties are elucidated first, then the cylindrical particle velocity distribution is discussed.

Figure 7 shows the typical trajectory of a single CA cylindrical particle during binary particle fluidization. The particle pathline shown in the figure encompasses 96 location samples for a 5.3 sec time sequence. It is clear that the motion of a single cylindrical particle forms a complete circulation route inside the binary bed. The cylindrical particle first rises quickly along the central axis of the bed and then begins to move radially towards the bed wall. After it reaches the bed wall, the cylindrical particle begins to migrate down the bed wall. When the cylindrical particle approaches the gas distributor, it gradually moves toward the bed center region and the next circulation begins. This phenomenon has been reported by others \(^{20,21,53,54}\). The pathline becomes twisted and “jerky” in some places as shown by the inset; this is because in this region, the upward and downward motion of cylindrical particle is intermittent. For example, sometimes the particle motion nearly stops and then begins.
to jump or jerk around its position \(^{20,21}\). This phenomenon may be caused by the interaction between a large bubble and a group of cylindrical particles. The histograms in Figure 8a show the raw measurements data distribution of 3 data points randomly picked from the velocity curve in Figure 8a. The data normality is tested with Kolmogorov-Smirnov method and it is confirmed that velocity follow a normal distribution with significance level \(\alpha = 0.05\). However, the standard deviation \(\sigma\) is relatively great, which can be attributed to the rapid particle jerking motion shown in Figure 7. It is commonly accepted that the total velocity \(v_{\text{total}}\) of cylindrical particle consists of two parts: the slower mainstream velocity \(v_s\) caused by the internal circulation, and the rapid and high frequency jerking velocity \(v_f\) caused by passing bubbles, i.e. \(v_{\text{total}} = v_s + v_f\). Some researchers argue that although \(v_f\) is very fast, but it is in fact a kind of vibration and adds up to no net motion in any direction \(^{21}\), and our experimental investigation also supports this opinion. Therefore, \(v_f\) makes the velocity standard deviation \(\sigma\) remarkably great, but as \(v_f\) is vibration motion velocity, which means the distribution of \(v_f\) is symmetric with respect to \(v_f = 0\) and results in normal distribution of \(v_{\text{total}}\). Positive and negative \(v_f\) counteracts with each other when calculating the mean value, thus the velocity mean value is representative when characterizing the circulation motion of cylindrical particles, and it eliminates that bias caused by particle jerking as well.

The trajectory of the cylindrical particle implies the vertical and horizontal velocity components vary in different ways, depending on particle position. Therefore, it is reasonable to discuss the distribution of the average vertical \((u_V)\) and horizontal \((u_H)\) velocity components separately. Figure 8a, 8b, 8d, and 8e show the \(u_V\) and \(u_H\) velocity components for both CA and CB cylindrical particles and \(u_f = 0.65\) m/s \((5.9u_{mf})\), \(\alpha = 10\%\), and GB300-500 bed material. The horizontal axis is the
dimensionless radial coordinate \( r/R \). The vertical axis is the \( u_V \) or \( u_H \) value, with \( u_V \) positive moving away from the gas distributor (up) and negative moving towards the gas distributor (down). A positive \( u_H \) means the particle is moving towards the wall and a negative \( u_H \) means particle is moving away from the wall. Different curves represent different height ranges \( (h) \) in which the particle is located. The individual data points represent the mean velocity value of all particles located in the given \( r/R \) and \( h \) ranges.

Figures 8a and 8b display the general characteristics of the \( u_V \) distribution. When \( 0<h<1.2H_0 \), \( u_V \) decreases from a positive velocity to a negative velocity with increasing \( r/R \). When \( r/R \) is in the bed centerline or wall region, \( u_V \) is upward or downward, respectively, which agrees well with the trajectory profile. The upward velocity in the centerline region is generally higher than the downward velocity in near-wall region, which explains why the particle resident time in the centerline region is less than in the near-wall region. Figures 8a and 8b also show when \( u_V = 0 \), the inflection point where the particle velocity changes from upward to downward, and reveals that the radial position of the inflection point moves toward the bed wall as \( h \) increases. Hence, a funnel-shaped region, as schematically shown in Figure 10c with light blue curve, is identified where \( u_V \) is upward inside the funnel and downward outside the funnel. Figure 10c also depicts that the decrease of particle sphericity \( \Phi \) enlarges the upper part of the funnel-shaped region. When \( 0<h<1.2H_0 \), the absolute value of \( u_V \) generally increases with \( h \), and may be attributed to the growing bubbles as \( h \) increases. When \( h>1.2H_0 \), \( u_V \) varies considerably, particularly in Figure 8b. According to the distribution characteristics described previously, this region is in fact the freeboard region where the cylindrical particles are mostly influenced by bubble agitation and rupture, which is an extremely unstable and
stochastic process. However, $u_V$ still decreases with increasing $r/R$ when $r/R > 0.5$, which implies that wall-induced drag is affecting the particles in this region.

Figures 8d and 8e display the overall characteristics of the $u_H$ distribution. The absolute value of $u_H$ first increases and then decreases with increasing $r/R$. For CA particles ($\phi = 0.85$) in Figure 8d, $u_H$ is positive when $h > 1.2H_0$. But for CB particles in Figure 8e, $u_H$ is positive when $h > 0.8H_0$. Considering the circulation route described in the previous section, this phenomenon indicates that the decrease of $\Phi$ causes the particles to move towards the wall at a lower bed region. Additionally, the two $u_H$ curves when $0 < h < 0.8H_0$ and $1.2H_0 < h < 2.0H_0$ are similar for both CA and CB systems, which indicates that $u_H$ is insensitive to the variation of $h$ in these two ranges. The curve when $0.8H_0 < h < 1.2H_0$ varies between the CA and CB system because this is the boundary region between positive and negative $u_H$.

The cylindrical particle vertical velocities were averaged vertically ($u_{V,va}$) for a fixed $r/R$ locations (Figures 9a and 9b), as well as horizontally ($u_{V,ha}$) for a fixed $h$ value (Figures 9c and 9d), for the different CA and CB particles with $\alpha = 10\%$ and GB300-500 bed material. Figures 9a and 9b illustrate the relationship between $u_{V,va}$ and $r/R$ for different $u_f$. The $u_{V,va} = 0$ curve is also marked in the figures with a dashed horizontal line, and the intersection points between the $u_{V,va} = 0$ curve and the $u_{V,va}$ curves can be regarded as the averaged $u_V$ direction inflection points under different $u_f$. Figures 9a and 9b show that the radial position of the inflection point gets closer to the bed wall as $u_f$ increases, which means that the diameter of the center region where the cylindrical particles move upward increases with increasing $u_f$. When $u_{V,ha} > 0$, $u_{V,ha}$ increase as $u_f$ increases, however, the when $u_{V,ha} < 0$, $u_{V,ha}$ does not vary significantly as $u_f$ increases except for $u_f = 0.16$ m/s ($1.5u_{mf}$). Figures 9c and 9d show the relationship between $u_{V,ha}$ and $h$ for different $u_f$. When $0 < h < H_0$ ($= 100$ mm), $u_{V,ha}$
is nearly zero, which can be attributed to the balance between the upward and downward particle streams. When \( h > H_0 \), \( u_{v,ha} \) increases with \( h \), which is cause by the particle being propelled upward by fast-rising large bubbles and then being thrown towards the wall where they interact with other particles or the wall itself, slowing the downward particle velocity.

Similar vertical and horizontal averages for the horizontal velocity \( u_H \) were also generated; Figure 10 displays these results for different CA and CB particles with \( \alpha = 30\% \) and GB300-500 bed material. Figures 10a and 10b illustrate the variation of vertical averaged \( u_H \) (written as \( u_{H,va} \)) at fixed \( r/R \) values under different \( u_f \) conditions. Both \( u_{H,va} \) in Figures 10a and 10b are nearly zero and independent of \( u_f \) over the entire range of \( r/R \) values. This is can be attributed to the balance between the toward-wall and away-from-wall particle motions during the circulation. However, if the direction of \( u_H \) is neglected and the average absolute value of \( u_H \) is considered (written as \( |u_{H,va}| \)), Figure 10c and 10d are obtained. These figures show that the \( |u_{H,va}| \) profile increases monotonically with increasing \( u_f \) except in the near-wall region, which means the horizontal motion of the cylindrical particles, and the accompanying particle circulation, becomes stronger as \( u_f \) increases. Figures 10e and 10f show the horizontal averaged \( u_H \) (written as \( u_{H,ha} \)) with \( h \) under different \( u_f \). The \( u_{H,ha} = 0 \) curve is also marked in the figures with a dashed horizontal line, and the intersection points between the \( u_{H,ha} = 0 \) curve and the \( u_{H,ha} \) curves can be regarded as the averaged \( u_H \) direction inflection points under different \( u_f \). From the figures, it is obvious that the average \( u_H \) direction inflection points becomes higher as \( u_f \) increases, which means the cylindrical particle gross circulation region become larger with increasing \( u_f \). Considering the increase of the center region where cylindrical particles move upward when \( u_f \) increases, we can deduce that the center of the particle circulation route
shown in Figure 7 will get higher and closer to the wall as $u_f$ increases.

Figures 11a-c show the effect of cylindrical particle mass fraction $\alpha$ on $u_{V,va}$. When $u_f = 0.16$ m/s ($1.5u_{mf}$), $u_{V,va}$ is nearly zero for all three mass fractions. Increasing $u_f$ produces a small effect of $\alpha$, but the trends are inconsistent with $\alpha$ and particle type. Figures 11d-f manifest the effect of $\alpha$ on the $u_{H,ha}$. It is obvious that in the freeboard region ($h > H_0$), $u_{H,ha}$ is highest when $\alpha = 10\%$ for all $u_f$ values investigated, but there is no consistent trend in the bed bulk ($h < H_0$).

Figures 12a-c display the effect of particle sphericity $\Phi$ on $u_{V,va}$. When the mass fraction of cylindrical particle is relatively low (e.g., $\alpha = 10\%$ in Figure 12a), the effect of $\Phi$ is negligible. When $\alpha$ is much higher (e.g., $\alpha = 30\%$ and $50\%$ in Figures 12b and 12c), a small effect of $\Phi$ becomes evident where $u_{V,va}$ first increases and then decreases with the increase of $\Phi$ when $0.69 \leq \Phi \leq 1$. Figures 12d-f exhibit a negligible effect of particle sphericity $\Phi$ on the $u_{H,ha}$. Figures 13a-f manifest the effect of bed material size on $u_{V,va}$ and $u_{H,ha}$. Both $u_{V,va}$ and $u_{H,ha}$ decrease as the bed material size increases, for all conditions in this study.

Conclusions

The fluidization of a binary fluidized bed composed of larger cylindrical particles and smaller bed material was experimentally studied using an X-ray stereography XPTV system. The fabrication method of the X-ray tracer particle and its three dimensional position determination were outlined. The effects of superficial gas velocity $u_f$, cylindrical particle mass fraction $\alpha$, particle sphericity $\Phi$, and bed material size on the cylindrical particle distribution and velocity were investigated and discussed in detail.

Overall, the cylindrical particles were found in the near-wall region more often than in the bed centerline region. Increasing the superficial gas velocity $u_f$ provided a
slight improvement in the uniformity of the vertical and horizontal distributions. Increasing the cylindrical particle mass fraction $\alpha$ caused the bed mixing conditions to transition from complete mixed into partial mixed, increased the particle penetration into the bottom of the bed (particularly along the bed walls), and had no effect on the horizontal particle distribution. Under partial bed mixing, a cone-shaped cylindrical particle packing layer formed at the bottom of the bed. The accumulation of cylindrical particles in the bed bottom was more pronounced when the cylindrical particle sphericity was decreased while the volume equivalent diameter remained constant. For a fixed superficial gas velocity $u_f$, the increase in bed material size suppressed the dispersion of the cylindrical particles in the bed. However, when the fluidization number $u_f/u_{mf}$ was constant, larger bed material enhanced the cylindrical particle distribution.

The overall motion of the cylindrical particles formed an internal circulation route inside the bed consisting of a fast moving upward path in the center and a slower moving downward path near the bed wall. The radial coordinate where the cylindrical particle vertical velocity $u_V$ changed from up to down increased as $h$ increased, creating a funnel-like structure of upward moving particles. As $u_f$ increased, the region of upward moving particles increased, and the downward moving particle region near the wall became thinner. When the particle vertical position $h$ and static bed height $H_0$ satisfied $0<h<1.2H_0$, the absolute value of $u_V$ first increased and then decreased from the central bed region to the wall region, and increased overall with $h$. When $h>1.2H_0$, $u_V$ became random and hard to predict. The absolute value of cylindrical particle horizontal velocity $u_H$ first increased and then decreased with increasing $r/R$. Additionally, $u_H$ was insensitive to the variation of $h$ when $0<h<0.8H_0$ and $1.2H_0<h<2.0H_0$. The cylindrical particles circulation region became larger with
increasing $u_f$.

For the vertical averaged vertical velocity $u_{v,va}$ and horizontal averaged horizontal velocity $u_{h,ha}$, when $\alpha \leq 30\%$, $u_{v,va}$ was insensitive to the variation of $\alpha$. When $\alpha \geq 30\%$, both $u_{v,va}$ and $u_{h,ha}$ decreased with increasing $\alpha$. When $\alpha$ was relatively low, $u_{v,va}$ was independent of $\Phi$. When $\alpha$ was large, $u_{v,va}$ first increased and then decreased with the increase of $\Phi$. The effect of $\Phi$ on $u_{h,ha}$ is nearly negligible. Both $u_{v,va}$ and $u_{h,ha}$ decreased as bed material size increased.

This study provided unique data using XPTV, and increased the understanding of binary fluidization of non-spherical particles. Additionally, the data can be used to validate numerical simulations.

**Acknowledgments**

Financial supports from the Major Program of NSFC (Grant No. 51390492) and General Program of NSFC (Grant No. 51576045) are sincerely acknowledged. The X-ray facility used in this work was developed with support from the US National Science Foundation under Grant No. CTS-0216367 and Iowa State University.

**Notation**

- $a$: distance between X-ray source and the coordinate origin O (mm)
- $Asr$: cylindrical particle aspect ratio, i.e. $l/d_i$
- $b$: distance between X-ray detector surface and the coordinate origin O (mm)
- $C_{i,N=1000}$: distribution probability of the tracer particle on position $i$ when $N=1000$
- $C_{i,N=2000}$: distribution probability of the tracer particle on position $i$ when $N=2000$
- $d_i$: diameter of cylinder particle (mm)
- $d_{imax}$: maximum diameter of bed material particle (mm)
- $d_{imin}$: minimum diameter of bed material particle (mm)
- $d_j$: average diameter of bed material particle (mm)
\( d_v \)  cylindrical particle volume equivalent diameter; diameter of Sphere A (mm)

\( D \)  fluidized bed diameter (mm)

\( h \)  particle height from the distributor plate (mm)

\( H \)  bed height in fluidized bed from the distributor plate (mm)

\( H_0 \)  static bed height (mm)

\( i \)  identifier of positions on the distribution contour plots

\( K \)  total number of positions on the distribution contour plots

\( l \)  length of cylindrical particle (mm)

\( m_i \)  mass of a cylindrical particle (kg)

\( N \)  total number of observations

\( O, O_1, O_2 \)  coordinate origin, center of the imaging plane on X-ray detector 1 and 2

\( P, P_1, P_2 \)  tracer center 3D position and its projection onto X-ray detector 1 and 2

\( S_1, S_2 \)  position where X-ray source 1 and 2 are emitted

\( r \)  the horizontal distance between the tracer particle centroid and bed central axis (mm)

\( R \)  the fluidized bed radius (mm)

\( \text{SSE} \)  sum of the squared error

\( u \)  tracer particle velocity vector (m/s)

\( u_f \)  superficial gas velocity (m/s)

\( u_{H} \)  the horizontal component of cylindrical particle velocity (m/s)

\( u_{H,va} \), \( u_{H,ha} \)  the vertical or horizontal averaged value of \( u_{H} \) (m/s)

\( |u_{H,va}| \)  the absolute value of the horizontal averaged \( u_{H} \) (m/s)

\( u_{mf} \)  minimum fluidization gas velocity (m/s)

\( u_v \)  the vertical component of cylindrical particle velocity (m/s)
\( u_{V,\text{va}}, u_{V,\text{ha}} \) the vertical or horizontal averaged value of \( u_V \) (m/s)

\( x, y, z \) tracer particle center 3D position \( x, y, \) and \( z \) coordinate (mm)

\( x_1, z_1 \) \( x \) and \( z \) coordinate of projection center position of tracer particle on detector 1 (mm)

\( y_2, z_2 \) \( y \) and \( z \) coordinate of projection center position of tracer particle on detector 2 (mm)

Greek letters

\( \alpha \) cylindrical/large spherical particle mass fraction

\( \varepsilon_0 \) cylindrical/spherical particle static packing voidage

\( \rho_i \) cylindrical particle density (kg/m\(^3\))

\( \rho_j \) bed material particle density (kg/m\(^3\))

\( \Phi \) particle sphericity

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Table 1. Large Particle Properties

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<th>Particle type</th>
<th>Material</th>
<th>$d_i$ (mm)</th>
<th>$l$ (mm)</th>
<th>$d_v$ (mm)</th>
<th>$\rho_i$ (kg/m$^3$)</th>
<th>$\Phi$</th>
<th>$\text{Asr}$ ($l/d_i$)</th>
<th>$\varepsilon_0$</th>
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<td>Cylinder A</td>
<td>ABS</td>
<td>4.5±0.1</td>
<td>7.1±0.5</td>
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<td>Cylinder B</td>
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<td>5.33</td>
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<tr>
<td>Sphere A</td>
<td>ABS</td>
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<td>6.0</td>
<td>998.8±2.1</td>
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<td>-</td>
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Table 2. Bed Material Properties

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<th>Bed material type</th>
<th>Material</th>
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<th>$d_{j\text{min}}$ (μm)</th>
<th>$\rho_j$ (kg/m$^3$)</th>
<th>$\varepsilon_0$</th>
<th>$u_{mf}$ (m/s)</th>
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<td>BG300-500</td>
<td>Glass bead</td>
<td>300</td>
<td>500</td>
<td>2500.6±7.8</td>
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<td>0.11</td>
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<td>BG800-1000</td>
<td>Glass bead</td>
<td>800</td>
<td>1000</td>
<td>2537.4±2.7</td>
<td>0.367</td>
<td>0.54</td>
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Figure List

Figure 1. Schematic diagram of the fluidized bed experimental system (a) and the stereo X-ray stereography system (b): (1) data processing computer, (2) pressure sensor, (3) fluidized bed, (4) gas chamber, (5) valve controller, (6) mass flow rate control valve, (7) water tank, (8) X-ray source 1, (9) X-ray source 2, (10) X-ray detector 1, (11) X-ray detector 2.

Figure 2. Cylindrical particle distribution results with different number of observations $N$ and the variation of resulting SSE with $N$.

Figure 3. Photographs and cylindrical particle distribution contour maps with bed material GB300-500 at $\alpha = 10\%$: (a), (c), and (e) for CA ($\Phi = 0.85$) with $u_f = 0.16$ m/s, 0.40 m/s and 0.65 m/s; (b), (d), and (f) for CB ($\Phi = 0.69$) with the same $u_f$ conditions.

Figure 4. Cylindrical particle distribution of CB particles ($\Phi = 0.69$) at (a) $\alpha = 30\%$ and (b) $\alpha = 50\%$ with $u_f = 0.65$ m/s ($5.9u_{mf}$) and bed material GB300-500.

Figure 5. Particle distribution of SA particles ($\Phi = 1.00$) with $\alpha = 10\%$, $u_f = 0.40$ m/s ($3.6u_{mf}$), and bed material GB300-500.

Figure 6. Photographs and cylindrical particle distribution contour maps of CA particles ($\Phi = 0.85$) with $u_f = 0.65$ m/s ($1.2u_{mf}$) (a) and $u_f = 0.90$ m/s ($1.7u_{mf}$) (c); CB particles ($\Phi = 0.69$) with $u_f = 0.65$ m/s ($1.2u_{mf}$) (b) and $u_f = 0.90$ m/s ($1.7u_{mf}$) (d), all with $\alpha = 10\%$ and bed material GB800-1000.

Figure 7. Brief circulation trajectory of a single cylindrical particle in a binary fluidized bed.

Figure 8. The vertical ($u_V$) and horizontal ($u_H$) velocity components for CA ($\Phi = 0.85$) (a) and (d), and CB ($\Phi = 0.69$) particles (b) and (e), both with $\alpha =$
10%, $u_f = 0.65$ m/s ($5.9u_{mf}$) and bed material GB300-500, together with
the particle distribution sketch(c) and raw measurement data distribution
histograms (a).

Figure 9. The effects of superficial gas velocity $u_f$ on $u_{V,va}$ and $u_{V,ha}$ for CA particles ($\Phi = 0.85$) (a) and (c), and CB particles ($\Phi = 0.69$) (b) and (d), both with $\alpha = 10\%$ and bed material GB300-500.

Figure 10. The effects of superficial gas velocity $u_f$ on $u_{H,va}$, $|u_{H,va}|$, and $u_{H,ha}$ for CA particles ($\Phi = 0.85$) (a), (c), and (e), and CB particles ($\Phi = 0.69$) (b), (d), and (f), both with $\alpha = 30\%$ and bed material GB300-500.

Figure 11. The effects of particle mass fraction $\alpha$ on $u_{V,va}$ and $u_{H,ha}$ for CA particles ($\Phi = 0.85$) (a) and (d), CB particles ($\Phi = 0.69$) (b) and (e), and SA particles ($\Phi = 1.00$) (c) and (f), all with bed material GB300-500.

Figure 12. The effects of particle sphericity $\Phi$ on $u_{V,va}$ and $u_{H,ha}$ when $\alpha = 10\%$ (a) and (d), $\alpha = 30\%$ (b) and (e), and $\alpha = 50\%$ (c) and (f), all with bed material GB300-500.

Figure 13. The effects of bed material particle size on $u_{V,va}$ and $u_{H,ha}$ for CA particles ($\Phi = 0.85$) (a) and (d), CB particles ($\Phi = 0.69$) (b) and (e), and SA particles ($\Phi = 1.00$) (c) and (f), all with $u_f = 0.65$ m/s.
Trajectory overlapping caused by particle jerking

Projection 1

Projection 2

Tracer particle

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