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Effects of Mixing Using Side Port Air Injection on a Biomass Fluidized Bed

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Abstract
Fluidized beds are being used in practice to gasify biomass to create producer gas, a flammable gas that can be used for process heating. However, recent literature has identified the need to better understand and characterize biomass fluidization hydrodynamics, and has motivated the combined experimental-numerical effort in this work. A cylindrical reactor is considered and a side port is introduced to inject air and promote mixing within the bed. Comparisons between the computational fluid dynamics (CFD) simulations with experiments indicate that three-dimensional simulations are necessary to capture the fluidization behavior of the more complex geometry. This paper considers the effects of increasing side port air flow on the homogeneity of the bed material in a 10.2 cm diameter fluidized bed filled with 500-600 μm ground walnut shell particles. The use of two air injection ports diametrically opposed to each other is also modeled using CFD to determine their effects on fluidization hydrodynamics. Whenever possible, the simulations are compared to experimental data of time-average local gas holdup obtained using X-ray computed tomography. This study will show that increasing the fluidization and side port air flows contribute to a more homogeneous bed. Furthermore, the introduction of two side ports results in a more symmetric gas-solid distribution.

Disciplines
Chemical Engineering | Mechanical Engineering

Comments
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Fluidized beds are being used in practice to gasify biomass to create producer gas, a flammable gas that can be used for process heating. However, recent literature has identified the need to better understand and characterize biomass fluidization hydrodynamics, and has motivated the combined experimental-numerical effort in this work. A cylindrical reactor is considered and a side port is introduced to inject air and promote mixing within the bed. Comparisons between the computational fluid dynamics (CFD) simulations with experiments indicate that three-dimensional simulations are necessary to capture the fluidization behavior of the more complex geometry. This paper considers the effects of increasing side port air flow on the homogeneity of the bed material in a 10.2 cm diameter fluidized bed filled with 500-600 μm ground walnut shell particles. The use of two air injection ports diametrically opposed to each other is also modeled using CFD to determine their effects on fluidization hydrodynamics. Whenever possible, the simulations are compared to experimental data of time-average local gas holdup obtained using X-ray computed tomography. This study will show that increasing the fluidization and side port air flows contribute to a more homogeneous bed. Furthermore, the introduction of two side ports results in a more symmetric gas-solid distribution.

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1 Introduction

Fluidized bed gasifiers are found in many industrial processes to convert feedstocks with low-carbon content into valuable products such as fuels, basic chemicals, and hydrogen. Some advantages of fluidized bed operation include nearly isothermal conditions due to rapid mixing of particles, high heat and mass transfer rates, and the ability to work with particles of various sizes [1,2]. The use of biomass in fluidized beds is of current interest because biomass is considered a renewable alternative energy resource that can potentially provide low cost power production or process heating needs. Although biomass gasifiers are being built and used in biorefineries, there are problems with fluidizing the media. Biomass particles are typically difficult to fluidize due to their peculiar shape and a second inert material, such as sand, alumina, or calcite, is typically added to the bed. The large differences in size and density between the biomass and inert particles lead to non-uniform distribution of the biomass within the fluidized bed, and particle interactions and mixing become major issues. Therefore, the fluidization characteristics of biomass particles are of critical importance because of known problems such as particle agglomeration, defluidization, elutriation, and segregation [3-8].

One method to enhance and promote mixing in a fluidized bed is with the inclusion of a side port injection, where either additional gas, biomass or some combination is fed into the reactor bed. There have been studies on the effects of injecting gas through a side port and the influence on fluidization and gasification. Experimental studies of Rajan et al. [9] found that side air injection influenced the circulation pattern of the fluidized bed. Low jet flow promoted mixing and enhanced solid circulation, while high flowrates showed a tendency to increase elutriation, and in small diameter beds, caused slugging. Chyang et al. [10] experimentally studied the modes of gas discharge from a single jet in a two-dimensional (2D) fluidized bed. They identified three regimes for either a bubbling, transition, or jetting flow by comparing a modified Froude number and the ratio of nozzle-to-particle diameter. Chen and Weinstein [11] numerically and experimentally investigated a 2D fluidized bed with horizontal jet injection. They compared a solid’s volume fraction in the jet-influenced area and found three different regions: coherent voids, bubble trains and a zone of surrounding compaction. Another important contribution was made by Xuereb et al. [12] who experimentally determined the jet penetration length and expansion angle as well as the effects of inlet and jet velocity and particle diameter. They confirmed that close to the injection point, there is a dragging zone of particles from the dense phase into the jet.

Earlier work on horizontal injection mainly focused on finding empirical correlations to measure the jet penetration length [13-15]. Geometric parameters that characterize the injection port (e.g., shape, diameter, and location) and flow conditions such as fluidizing gas velocity and side air injection velocity, have been considered to determine their effects on jet penetration. Hong et al. [16] proposed a correlation for inclined jets, based on Merry’s correlation [15] for horizontal jets and validated it with experimental data and numerical simulations using a two-fluid model for fluidization. The influences of gas velocity of the jet, nozzle diameter, inclination, and location were studied in detail. Numerical simulations conducted by Tyler and Mees [17] compared three discretization schemes (hybrid, minmod, and Superbee), and found that simulating the bed with the Superbee scheme resulted in bubble and jet behavior, shape, and formation consistent with what was observed in experiments. From their preliminary study with qualitative comparisons, they concluded that three-dimensional (3D) simulations were in best agreement with experiments. More numerical simulations were performed by Li et al. [18] using a scaled Gilbaro drag model in a 3D cylindrical reactor to avoid overprediction of bed expansion and agglomeration of particles in the bed. Li et al. [19] also investigated the effects of single and multiple jets on the hydrodynamics of a
rectangular fluidized bed. They concluded that multiple jets do not influence each other significantly until they start to overlap. Another conclusion was that gas injection strongly affects the fluid behavior of the bed that is above the injection port when the injection flow rate is relatively high, and side effects are negligible in the part below the injection port. It was also shown that deep penetration of the jet enhanced solid circulation in the core; however, deeper penetration could lead to a slugging bed flow.

Initially, the use of 2D versus 3D simulations will be compared with experiments to determine the best approach to model and capture the bed hydrodynamics adequately. From a computational resource point of view, 2D simulations are easier to perform than 3D simulations, but they may not capture the proper physics. Previous work of Xie et al. [20] and Deza et al. [21,22] have shown very good agreement using a 2D approach for a cylinder reactor with no side air injection when the flow is limited to the bubbling regime for Geldart B particles. To examine the influence of side port air injection, glass beads were used for the bed material in the preliminary studies by Deza et al. [23] and Min et al. [24] because the properties of glass are well-characterized. The simulations for both 2D and 3D representations of the reactor compared well with the experiments. In this work, simulations of a fluidized bed reactor with side port using a medium of ground walnut shell, a Geldart B particle, will be studied to determine if 2D simulations are appropriate to properly model the fluidization of biomass. This paper considers the effects of increasing fluidization air flow and sideport air flow on the homogeneity of the bed material in a 10.2 cm diameter fluidized bed. Two air injection ports diametrically opposite to each other are also considered using CFD to determine how the ports affect the fluidization hydrodynamics. Whenever possible, the simulations are compared to experimental data of time-average local gas holdup obtained using X-ray computed tomography.

2 Experimental Setup

2.1 Fluidized Bed Reactor. A fluidized bed reactor with a side injection port and an internal diameter of 10.2 cm is used in the experiments. The three main components of the reactor are the top chamber (freeboard), bed chamber, and plenum as illustrated in Fig. 1(a). The material fluidizes in the bed chamber, which includes a single injection port on the sidewall 2.8 cm from the bottom of the bed chamber. A 1.1 cm ID polypropylene tube is attached to the injection port to supply additional air to improve mixing. The distributor located below the bed is comprised of approximately 1.3 cm hole spacing. A 45 mesh screen is attached to the distributor plate to avoid particles from accumulating in the holes. Under the distributor plate is the plenum filled with glass marbles to evenly distribute the fluidizing gas. Above the chamber is a 61 cm tall freeboard region to avoid elutriation of bed material.

Pressure is measured with a 0 to 34.5 kPa pressure transducer. The pressure transducer is located in the air inlet plate and has a maximum error of ± 86 Pa, which corresponds to ± 0.25% of the full scale. Fluidization and side injection air, supplied by the laboratory compressed air system, are controlled using ball valves, pressure regulators, and flow meters. The error of the flow meters is less than ±2% of the full scale reading.

2.2 X-Ray System. Iowa State University’s XFloViz facility was used to image the fluidized bed in this study and has been described in detail in the literature [25]. Consequently, only a brief outline will be presented here. Two LORAD LPX200 portable X-ray tubes provide the X-ray energy. Current and voltage can be adjusted from 0.1 to 10.0 mA and 10 to 200 kV, respectively, with a maximum power of 900 W. Low energy radiation is suppressed by 1 mm thick copper and aluminum filters. In this study, only X-ray computed tomography (CT) imaging is considered. For this purpose, located opposite one of the X-ray sources is a CCD camera connected to a square 44 x 44 cm cesium-iodide phosphor screen which transforms radiation into visible light. A 50 mm Nikon lens captures images which are digitized by an Apogee Alta U9 CCD system. This system has 3072 x 2048 pixels and is thermoelectrically cooled to allow long exposure times. Usually, an exposure time of 1 s with 4 x 4 binning is chosen to minimize acquisition time while maintaining the signal strength. The camera system and X-ray source are located on a 1.0 m ID rotation ring that can rotate 360° around the fluidized bed. CT data are acquired using software developed by Iowa State University’s Center for Nondestructive Evaluation (CNDE) and a computer-control data acquisition system. The software allows for control of the camera/detector pair, as well as motion control for the rotation ring. Volumetric reconstruction of the CT images is performed using CNDE’s 64-node LINUX cluster.

2.3 CT Images. In X-ray computed tomography with a conical X-ray beam, a series of 2D projections are captured at various rotation angles and reconstructed into a 3D volumetric image. Since multiple images must be acquired for one CT, the resulting 3D image is necessarily time-averaged. This image is a map of CT intensity values which are proportional to X-ray attenuation, which in turn is proportional to density. In this study, only CT images of the fluidized bed filled with ground walnut shell were acquired; these consisted of recording 360 projections, one at every degree, at a power setting of 130 kV and 4.2 mA. For all tests, the exposure time was 1 s at 4 x 4 binning per degree, and each test took approximately 45 mins. A total of 300 vertical slices were captured. To minimize image acquisition noise, the CCD camera was cooled to 0°C using the thermoelectric cooler. A pixel non-uniformity calibration was employed [25] to adjust individual pixels to respond identically to incident X-ray energy. After calibration, the 2D projections were reconstructed into 3D images using CNDE software.

2.4 Gas Holdup. In order to quantify the CT data, time-average local gas holdup (void fraction) was calculated for each flow condition. The time-average local gas holdup, e_v, can be determined by knowing the local X-ray attenuation for the flow (\(I_f\)), the particle (\(I_p\)), and the gas (\(I_g\)). Since the attenuation is proportional to the CT intensity (\(CTI\)), the time-average local gas holdup can be calculated by knowing time-average local CT intensity data for the flow, the particle (\(CTI_f\)), and the gas (\(CTI_g\)). Therefore, the time-average local gas holdup is defined as:

![Fig. 1 Schematic of the (a) fluidized bed used in the experiment and (b) bed chamber used in the simulation including the side port injector](image-url)
\[ \epsilon_g = \frac{\lambda - \lambda_p}{\lambda_c - \lambda_p} = \frac{CTI - CTIp}{CTI_c - CTIp} \]

(1)

It is difficult to determine the CT intensity for a single particle due to its small size; however, the local CT intensity for a static (bulk) bed of particles (CTIb) can be used. From Eq. (1), the void fraction for the bulk material can be calculated using local CT intensities for the bed, where:

\[ \epsilon_{g,b} = \frac{CTI_b - CTIp}{CTI_c - CTIp} \]

(2)

For a granular material, the void fraction of the bulk material \( \epsilon_{g,b} \) is defined as:

\[ \epsilon_{g,b} = 1 - \frac{\rho_b}{\rho} = \text{Constant} \]

(3)

The bed material bulk density \( \rho_b \) and particle density \( \rho \) can be found experimentally and in property tables, respectively. Substituting CTI from Eq. (2) into Eq. (1) and rearranging yields an equation to determine the local gas holdup based on CTs for the flow condition, the gas, and the bulk material:

\[ \epsilon_g(i,j,k) - CTI_b(i,j,k) - \frac{\epsilon_{g,b}(i,j,k)\epsilon_g(i,j,k)}{CTI_g(i,j,k) - CTI_b(i,j,k)} = \text{Constant} \]

(4)

and \( i,j,k \) represent the locations of individual voxels in the 3D volume, where a voxel is a 3D pixel. In this study, CT data were acquired for a bed of static bulk material and the empty reactor (air only) at identical power settings used to capture fluidization acquired for a bed of static bulk material and the empty reactor volume, where a voxel is a 3D pixel. In this study, CT data were intensities for the bed, where:

\[ \epsilon_g(i,j,k) - CTI_b(i,j,k) - \frac{\epsilon_{g,b}(i,j,k)\epsilon_g(i,j,k)}{CTI_g(i,j,k) - CTI_b(i,j,k)} = \text{Constant} \]

and \( i,j,k \) represent the locations of individual voxels in the 3D volume, where a voxel is a 3D pixel. In this study, CT data were acquired for a bed of static bulk material and the empty reactor (air only) at identical power settings used to capture fluidization (flow) CT data. Using Eq. (4), each flow file was converted to show time-average local gas holdup and a smoothing method was employed to reduce noise. The resulting time-average gas holdup values are determined on a 3D grid with an approximate voxel size of 0.6 mm \( \times \) 0.6 mm \( \times \) 0.6 mm. Estimated absolute uncertainty in the local gas holdup is \( \pm 0.04 \), which is a worst-case estimate with most data falling within an absolute gas holdup error of \( \pm 0.02 \).

Three-dimensional images were viewed using internally developed visualization software, which allowed viewing of the volumetric images at any location within the imaging volume, and to adjust color mapping schemes. Since volume files contain information outside the cylindrical region of interest, a clipping feature was also used to isolate the fluidized bed. Once isolated, the spatial range was modified to show the vertical \( y-z \) plane (x-slice) and the vertical \( x-z \) plane (y-slice) through the column center, as well as horizontal \( x-y \) planes (z-slices) at heights of 3.2 cm and 9.0 cm from the distributor plate.

3 Two-Fluid Model

3.1 Governing Equations. A multifluid Eulerian-Eulerian model is employed in Multiphase Flow with Interphase eXchanges (MFIX) [26] and assumes that each phase behaves as interpenetrating continua with its own physical properties. The instantaneous variables are averaged over a region that is larger than the particle spacing but smaller than the flow domain. Volume fractions are introduced to track the fraction each phase occupies in the averaging volume, where \( \epsilon_g \) is the gas phase volume fraction (also referred to as the void fraction) and \( \epsilon_s \) is the solid phase volume fraction. The solid phase is described with an effective particle diameter \( d_p \) and characteristic material properties, and solved using conservation equations for the solid phase. The effective particle diameter is \( d_p = \psi \bar{d}_p \), where \( \bar{d}_p \) is the mean diameter and \( \psi \) is the estimated sphericity of the actual particles.

The continuity equations for the gas phase (g) and the solid phase (s), respectively, are:

\[ \frac{\partial}{\partial t} \left( \epsilon_g \rho_g \right) + \nabla \cdot \left( \epsilon_g \rho_g \mathbf{u}_g \right) = 0 \]

(5)

\[ \frac{\partial}{\partial t} \left( \epsilon_s \rho_s \right) + \nabla \cdot \left( \epsilon_s \rho_s \mathbf{u}_s \right) = 0 \]

(6)

with density \( \rho \) and velocity vector \( \mathbf{u} \).

The momentum equations for the gas and solid phases have the form:

\[ \frac{\partial}{\partial t} \left( \epsilon_g \rho_g \mathbf{u}_g \right) + \nabla \cdot \left( \epsilon_g \rho_g \mathbf{u}_g \mathbf{u}_g \right) = -\nabla \cdot \mathbf{P}_g + \nabla \cdot \mathbf{\sigma}_g + \mathbf{I}_g + \epsilon_g \rho_g \mathbf{g} \]

(7)

\[ \frac{\partial}{\partial t} \left( \epsilon_s \rho_s \mathbf{u}_s \right) + \nabla \cdot \left( \epsilon_s \rho_s \mathbf{u}_s \mathbf{u}_s \right) = -\nabla \cdot \mathbf{P}_g + \nabla \cdot \mathbf{\sigma}_s - \mathbf{I}_g + \epsilon_s \rho_s \mathbf{g} \]

(8)

where \( \mathbf{\sigma} \) is the stress tensors, \( \mathbf{g} \) is gravity, and \( \mathbf{I} \) the interaction forces accounting for the momentum transfer between the gas and solid phases.

The granular temperature \( \theta \) for the solid phase can be related to the granular energy, defined as the specific kinetic energy of the random fluctuating component of the particle velocity. The resulting transport equation for the granular energy [27] is:

\[ \frac{3}{2} \left( \frac{\partial}{\partial t} \left( \epsilon_g \rho_g \theta \right) + \nabla \cdot \left( \epsilon_g \rho_g \theta \mathbf{u}_g \right) \right) = \mathbf{\sigma}_s \cdot \mathbf{u}_s - \nabla \cdot \mathbf{q}_\theta - \gamma_\theta + \phi_\theta \]

(9)

where \( \mathbf{q}_\theta \) is the diffusive flux of granular energy, \( \gamma_\theta \) is the rate of granular energy dissipation due to inelastic collisions [28], and \( \phi_\theta \) is the transfer of granular energy between the gas phase and solid phase.

Kinetic theory for granular flow is used to calculate the solid stress tensor and solid-solid interaction force in the rapid granular flow regime [26]. There are two distinct flow regimes in granular flow: a viscous or rapidly shearing regime in which stresses arise due to collisional or translational momentum transfer, and a plastic or slowly shearing regime in which stresses arise due to Coulomb friction between solids in close contact. A blending function to provide a smooth transition between each regime is employed [20]. Further details related to the constitutive relations in Eqs. (7)–(9) can be found in the MFIX theory guide [26].

The interaction force accounts for the gas-solid momentum transfer, which is expressed as the product of the coefficient for the interphase drag force between the gas and solid phases and the slip velocity. The Gidaspow model [29] is used to represent the interphase drag force coefficient based on the Ergun and Wen-Yu equations, with a blending function to avoid a discontinuity between the use of the equations. Previous studies by the authors have shown the validity of using the model for glass beads and ground walnut shell [30].

3.2 Solution Methodology. To discretize the governing equations in MFIX, a finite volume approach for a staggered grid is used to reduce numerical instabilities [31]. Velocities are stored at the cell surfaces, and scalars, such as void fraction and pressure, are stored at the center of the cell. Discretization of time derivatives are first-order and discretization of spatial derivatives are second-order. An important feature is the use of a second-order discretization scheme for the convective terms, known as the Superbee method [32], which improves convergence and accuracy of the solution. A modification of the SIMPLE algorithm is used.
to solve the governing equations [31]. The first modification uses an equation for the solid volume fraction that includes the effect of the solids pressure to help facilitate convergence for both loosely and densely packed regions. The second modification uses a variable time-stepping scheme to improve convergence and execution speeds.

4 Results and Discussion

4.1 Problem Description. The schematic of the fluidized bed reactor used in the experiments is shown in Fig. 1(a); the simulations model the bed chamber shown in Fig. 1(b). For all simulations, air is uniformly provided at the bottom of the domain equal to the superficial gas velocity as a simplification of the flow across the distributor plate in the experiments. The side port injection is also modeled with uniform air velocity at the inlet. The no-slip condition is used to model the gas-wall interactions and a partial-slip condition is used for the particle-wall interactions [33]. Table 1 summarizes the ground walnut shell particle properties and flow conditions. To account for the non-spherical nature of the ground walnut shell (it is more chunk-like), the sphericity and coefficient of restitution were numerically estimated based on previous work by Deza et al. [30], whereas the other properties were provided from the experiments. Two inlet gas velocities are examined; the lower velocity of \( U_g = 1.5 U_{mf} \) represents a mild bubbling bed and the higher velocity of \( U_g = 3.0 U_{mf} \) represents a moderate industrial reactor flow rate [34], where \( U_{mf} \) is the minimum fluidization velocity. A base case with no side port air injection (\( Q_{side} = 0 \)) and two additional cases of \( Q_{side} = 5\% \) and 10\% \( Q_{mf} \) are studied, where \( Q_{mf} \) is the minimum fluidization volumetric flow rate based on the bed inlet characteristics. Finally, a comparison using two ports diametrically opposite to each other with 5\%\( Q_{mf} \) air is studied; however, this particular case is only explored numerically to demonstrate the effects of multiple ports in enhanced mixing.

The grid resolution study by Deza et al. [30] identified a sufficient number of cells that would produce an estimated numerical error less than 1\%. The study was for a 2D domain, where a total of 2400 grid cells provided adequate resolution of the domain. The work herein uses a resolution for the 3D domain with 40 x 60 cells in the radial and axial directions and 16 cells in the azimuthal direction that form parallelepiped cells due to the circular cross-section of the domain. Although the grid resolution may seem coarse, Table 2 compares the computational time required for simulations performed on an AMD Opteron cluster (dual processor, dual core 2.4 GHz AMD 280 Opteron). The time step used by MFIX automatically adjusts to help the simulation converge. The mean time step for a 3D simulation with 3.0\( U_{mf} \cdot 10\% Q_{mf} \) was on the order of 0.00024 s. The simulations are time-averaged from 5 to 65 s (which represents the average of 6000 time-steps).

4.2 Two- and Three-Dimensional Simulations. The pressure drop across the ground walnut shell-filled fluidized bed versus the superficial gas inlet velocity when \( Q_{side} = 0 \) (base case) is shown in Fig. 2. The results compare the experimental measurements to that predicted using MFIX. Error bars are shown for both pressure and velocity on a sample of data to maintain clarity of the data presented. Once the bed is fluidized at \( U_{mf} = 18.4 \) cm/s, the measured pressure drop is approximately constant at 470 ± 86 Pa [35] whereas the predicted pressure drop is approximately 560 Pa. CFD modeling predicts the same pressure drop through the bed for both two- and three-dimensions. It should be noted that the slight discrepancy of CFD predictions with experiments may be due to errors associated with the irregular particle sizes for the ground walnut shell. Furthermore, the simulations modeled a single particle diameter of 550 \( \mu \)m, whereas the experiments used ground walnut shell particles with diameters ranging from 500 – 600 \( \mu \)m. It is particularly encouraging that for the base case, the 2D and 3D simulations are almost identical.

Figures 3(a)–3(c) present contours of the void fraction for the fluidized bed at \( U_g = 1.5 U_{mf} \) and \( Q_i = 10\% Q_{mf} \). The 2D and 3D simulations, Figs. 3(a) and 3(c), respectively, are compared to the experiments (Fig. 3(b)) using an interrogation region up to \( z = 15 \) cm and the bed diameter. The contours for the experiment and 3D simulation correspond to the \( x - z \) plane (see Fig. 1(b)) through the bed centerline and injection port. Also shown is void fraction averaged along horizontal planes, which produce a spatial average that can vary in the axial direction. Therefore, the curve in Fig. 3(d) shows the horizontal-average void fraction versus axial direction, and identifies that the bed expands to approximately 11 cm after fluidization. For the 2D simulation, time-average void fraction was horizontally-averaged across the bed width, while for the experiment and 3D simulation, horizontal averaging was performed for the \( x - y \) plane (circular cross-section). These side-by-side comparisons help elucidate the hydrodynamic features between both the 2D and 3D simulations and their agreement with the experiments. The side port air injection tends to cause a slight non-uniformity of the fluidized media near the port, which is accompanied by higher void fraction (i.e., more gas). Overall, the

<table>
<thead>
<tr>
<th>Table 1 Properties and Flow Characteristics for Walnut Shell</th>
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<tbody>
<tr>
<td>Property</td>
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<tr>
<td>Particle diameter, ( d ) (cm)</td>
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<tr>
<td>Particle density, ( \rho_s ) (g/cm(^3))</td>
</tr>
<tr>
<td>Bulk density, ( \rho_b ) (g/cm(^3))</td>
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<tr>
<td>Sphericity, ( \psi ) (-)</td>
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<td>Coefficient of restitution, ( e ) (-)</td>
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<tr>
<td>Initial void fraction, ( \varepsilon_i ) (-)</td>
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<td>Minimum fluidization velocity, ( U_{mf} ) (cm/s)</td>
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<table>
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<tr>
<th>Table 2 Central Processing Unit Information</th>
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<td></td>
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<tr>
<td>CPU time (s)</td>
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<tr>
<td># Time-steps</td>
</tr>
<tr>
<td>Average ( \Delta t ) (s)</td>
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<tr>
<td># Cells</td>
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<td>( \mu ) time-step/cell</td>
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Fig. 2 Pressure drop versus superficial gas velocity comparing experiments and simulations for the fluidized bed with no side port (\( Q_{side} = 0 \))
bed uniformly fluidizes; this feature is observed in the experiment, and the 3D simulation also predicts the same fluidization hydrodynamics. The 2D simulation predicts relatively uniform fluidization; however, there is more gas present within the center of the bed, displacing the solid particles. Figure 3(d) confirms that the 2D simulation predicts higher void fraction, which increases from 0.6 to 0.7 within the fluidizing bed. The 3D simulation predicts slightly lower void fraction, but it is very uniform, as indicated by

Fig. 3 Time-average void fraction of the fluidized bed at \( U_g = 1.5U_{mf} \) and \( Q_s = 10\%Q_{mf} \) for the (a) 2D simulation, (b) experiment, (c) 3D simulation, and (d) horizontal averages across the reactor diameter versus axial direction.

Fig. 4 Time-average void fraction of the fluidized bed at \( U_g = 3.0U_{mf} \) and \( Q_s = 10\%Q_{mf} \) for the (a) 2D simulation, (b) experiment, (c) 3D simulation, and (d) horizontal averages across the reactor diameter versus axial direction.

Fig. 5 Time-average void fraction profiles of the fluidized bed at \( U_g = 3.0U_{mf} \) and \( Q_s = 10\%Q_{mf} \) at (a) \( z = 3.2 \) cm and (b) \( z = 9.0 \) cm. Experimental data shown as symbols and simulations are shown as lines.
the constant value of 0.6; the average void fraction measured in the experiment is 0.62. The error bars on the experimental data represent an absolute gas holdup error of ±0.02, which is typical of most data.

The case for \( U_g = 3.0U_{mf} \) and \( Q_s = 10\%Q_{mf} \) is shown in Fig. 4. Examining the 2D void fraction contours (Fig. 4(a)), the non-uniformity of the fluidization is very apparent near the bed expansion height of 15 cm. The 3D simulation (Fig. 4(c)) compares remarkably well with the experiment (Fig. 4(b)), which is very encouraging because the inlet gas velocity \( U_g \) is large. The larger inlet gas velocity in combination with the side air injection suggests improved mixing throughout the bed, with a mean void fraction of 0.7. The 3D simulation slightly underpredicts the hydrodynamics; however, these discrepancies are most likely attributed to the single particle size used in the computational modeling or the estimate for the particle sphericity.

In an effort to further quantify and contrast the simulations with the experiments, time-average void fraction profiles at two axial locations, \( z = 3.2 \) and \( 9.0 \) cm, are shown in Fig. 5 at \( 3.0U_{mf} \) for \( 10\%Q_{mf} \). The experimental data and 3D simulations are local time-average values along a ray that passes through the centerline of the bed and side injection port at the given \( z \) height. The variations in the experimental data are attributed to the non-uniform inlet conditions that result from the 62 discrete air inlet holes of the distributor plate, and similar discrepancies have been shown by others [36–39]. The 3D prediction compares very well with the experiment, whereas the 2D simulation significantly overpredicts the presence of gas near the lower region of the bed (Fig. 5(a)).

Figures 3–5 elucidate the importance of modeling a 3D domain to capture the hydrodynamics for fluidizing biomass with side air injection. While 2D modeling is reasonable for mildly bubbling beds (e.g., \( 1.5U_{mf} \)), it is not sufficient when simulating higher flow rates (e.g., \( 3U_{mf} \)), especially for reactors with side port injection. The most likely reason is the lack of freedom for the particles to move azimuthally; thus limiting the validity for using a two-dimensional domain. Therefore, the remainder of the discussion presents numerical results based on modeling the full 3D domain.
4.3 Side Injection Flow Rate. It has been established that injecting air through a horizontal port promotes mixing in the fluidized bed. The effects of increasing side injection rate at a moderate inlet velocity of 3.0\(U_{mf}\) will be examined next. Three side injection flowrates of 5\(\%\)\(Q_{mf}\), 10\(\%\)\(Q_{mf}\), and 20\(\%\)\(Q_{mf}\) are presented in Fig. 6 for the time-average void fraction horizontally-averaged across the reactor diameter (\(x - y\) plane). Results from the experiments are shown as symbols and lines are used for the simulation data. In general, the mean void fraction trends are very similar, irrespective of side port air flow rate. With increasing axial position, the mean void fraction is relatively uniform until 10 cm, above which the void fraction gradually increases from 0.7 to 0.9 by 15 cm. Furthermore, the comparisons between the simulations and experiments are in good agreement.

To better understand the mixing trends, simulations and experiments are shown in Figs. 7(a)–7(c) as contour plots of the void fraction for the centerplane of the cylindrical reactor through the injection port (\(x - z\) plane) as well as two circular cross-sections (\(x - y\) planes) located at heights of \(z = 3.2\) cm (lower row) and 9.0 cm (upper row). For each case, experiments are on the left and 3D simulations on the right. The gas-solid distribution throughout the centerplane does not vary significantly with increasing side injection, except for the region near the port, where higher void fractions are present for higher side port air injection rates. The circular cross-sections at \(z = 3.2\) cm show higher void fractions because additional air injected through the port is present near this height. Annular sections of higher solid volume fraction are observed at \(z = 9.0\) cm because the particles tend to move toward the wall opposite to the port. The same trend is observed in both the experiments and simulations.

4.4 One versus Two Injection Ports. Injection flowrates of 5\(\%\) and 10\(\%\)\(Q_{mf}\) through one port and 5\(\%\)\(Q_{mf}\) through two ports (for a total of 10\(\%\)\(Q_{mf}\)) have been further compared in Fig. 8 for the void fraction averaged across the reactor diameter versus axial direction. As previously mentioned, only one side port was manufactured for the reactor used in the experiments. Results from the experiments are shown as symbols (only for 5\(\%\)\(Q_{mf}\) and 10\(\%\)\(Q_{mf}\) through one side port) and lines are used for the simulation data. As was observed in Fig. 8, the simulations are in good agreement with the experiments and the void fraction is relatively uniform through the bed. When comparing spatially and temporally average values, the effects of two side injectors are not sufficient for this fluidization flow rate.

Time-average local values, however, are affected by the number of side air injection ports. Figures 9(a)–9(c) show contour plots of the void fraction for no side port (\(Q_{side} = 0\)), one side air injection port with 10\(\%\)\(Q_{mf}\), and two side air injection ports each with 5\(\%\)\(Q_{mf}\), respectively. With one injection port, the particles move toward the opposite wall. However, two side air ports diametrically opposed improve the gas-solid distribution in the bed and eliminates the asymmetry of the flow.

To further quantitatively compare 3D simulations with the experiments, time-average void fraction profiles at two axial locations, \(z = 3.2\) and 9 cm, are shown in Fig. 10 for the same cases. Overall, the 3D predictions for local void fraction profiles compare well with the experiments. The void fraction is uniform at lower axial locations (Fig. 10(a)) irrespective of the side port, but the side ports tend to promote more uniform gas-solid distribution at higher axial locations (Fig. 10(b)).

This study has demonstrated how side port injection affects the mixing characteristics of the fluidizing bed. In practice, industrial reactors rely on side ports to inject additional gas as well as biomass or other granular material like coal. However, as was shown, a single port can adversely affect the homogeneity of the fluidizing material. These results elucidate that a second port may be advantageous to ensure proper mixing, which is extremely important for gasification to yield high quality producer gas. Future work will examine gasification of biomass and the effect of side port injection.
5 Conclusions

Numerical simulations of a biomass fluidizing bed with side air injection were compared to CT data for the gas-solid distribution to demonstrate the quantitative agreement for bed fluidization. Ground walnut shells in the range size of 500–600 μm were used to represent biomass because the material fluidizes uniformly and is classified as a Geldart type B particle. An Eulerian-Eulerian multifluid model was used to simulate and analyze gas-solid hydrodynamic behavior of the fluidized bed. Two- and three-dimensional simulations were performed to determine if both modeling approaches would capture the salient bed features. The predictions for pressure drop through the biomass bed were initially validated with the experiments and were found to be in good agreement. The findings showed that 2D simulations overpredicted the fluidized bed expansion and the results did not demonstrate a uniformly fluidizing bed. The 3D simulations compared well for all cases. This study demonstrates the importance of using a 3D model for a truly 3D flow in order to capture the hydrodynamics of the fluidized bed for a complicated flow and geometry.

The effects of increasing side port air flow on the homogeneity of the bed were investigated next. It was found that increasing the side port injection flow rate up to 20% of the total injection rate at two and three ports did not significantly affect the behavior of the bed, and the simulations compared well with the experimental measurements of void fraction. However, the simulations showed that adding a second side port injector on the opposite side of the reactor improved the mixing and overall homogeneity of the fluidized material. It would be of interest to study the effects of adding additional ports along the circumference of the reactor to further study their effect on the homogeneity of the bed.

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References


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