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Nathan P. Franka  
Iowa State University

Joshua B. Drake  
Iowa State University

Theodore J. Heindel  
Iowa State University, theindel@iastate.edu

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Abstract
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Keywords
fluidized bed, gasification, minimum fluidization velocity, particle injection, x-ray computed tomography

Disciplines
Acoustics, Dynamics, and Controls | Energy Systems | Fluid Dynamics

Comments
MINIMUM FLUIDIZATION VELOCITY AND GAS HOLDUP IN FLUIDIZED BEDS
WITH SIDE PORT AIR INJECTION

Nathan P. Franka, Joshua B. Drake, and Theodore J. Heindel*
Department of Mechanical Engineering
Iowa State University
Ames, Iowa 50011-2161

ABSTRACT
Fluidized beds can be used to gasify biomass in the production of producer gas, a flammable gas that can replace natural gas in process heating. Knowing how the fluidized bed hydrodynamics vary as reactor dimensions are scaled up is vital for improving reactor efficiency. This study utilizes 10.2 cm and 15.2 cm diameter fluidized beds with added side port air injection to investigate column diameter effects on fluidized bed hydrodynamics. Both inert (glass beads) and biomass (ground walnut shell and ground corncob) bed materials are used and the hydrodynamic differences with side port air injection are recorded. Minimum fluidization velocity is determined through pressure drop measurements. Time-averaged local and global gas holdup are recorded using X-ray computed tomography imaging. Results show that by varying the side port air flow rate as a percentage of the minimum fluidization flow rate, partial and complete fluidization is observed in both fluidized beds. Local gas holdup trends are also similar in both fluidized beds. These results will be used in future studies to validate computational fluid dynamics models of fluidized beds.

Keywords: fluidized bed, gasification, minimum fluidization velocity, particle injection, X-ray computed tomography

INTRODUCTION
Fluidized beds are often used in chemical, mineral, pharmaceutical, and energy industries for chemical processing and material sorting and drying. These beds are beneficial due to low pressure drops, approximately uniform temperature distributions, high heat and mass transfer rates, and the ability to fluidize many particle types of varying sizes [1-3]. In the biofuel industry, fluidized bed reactors are central components of thermochemical conversion processes such as combustion, pyrolysis, and gasification [1]. In gasification, biomass is typically injected into a heated bed of an inert catalyst, such as sand, and undergoes a thermochemical reaction to create a flammable hydrocarbon gas, called producer gas. When producer gas is cleaned, it is called synthesis gas (syngas). The inert catalyst bed provides uniform fluidization and high heat transfer rates. Gasification is becoming an important process for flammable gas production due to the large amount of biomass available and the limited production of greenhouse gases [1]. In addition, gasification may be directly applied to electricity generation. This is beneficial because quality requirements do not restrict usage of producer gas in combustion [4]. While gasification has recently gained popularity, its efficiency requires improvement, and consequently, the process must be researched in greater detail.

One difficulty in improving gasification efficiency is the inability to monitor biomass as it is injected into a fluidized bed. Problems arise with biomass reacting and being destroyed after injection into the bed. Also, since bed materials are typically opaque, efforts to optically visualize internal flow and track injected biomass particles are difficult. Shen et al. [5] simulated a biomass/sand system by injecting red wooden balls into a rectangular 2-D fluidized bed of glass beads. By using a digital image processing-based technique, it was found that the simulated biomass particles tended to move faster vertically than laterally; consequently, vertical convection was much higher than lateral convection. Additionally, as superficial air velocity increased, biomass concentrated near the surface and the
system decreased the time required to reach steady state. In that study, it was assumed that the measurement probes and injection system had no effect on biomass mixing patterns.

While some optical methods have been used to visualize fluidization behavior, these approaches cannot adequately monitor internal flow features or particle injection in larger 3-D cylindrical beds, which are more representative of real systems. Since fluidization is a dynamic process, invasive monitoring methods can influence the internal flow, thereby reducing the reliability of the measurements [6]. As a result, noninvasive monitoring techniques have been developed for use in multiphase systems. Some of these techniques include electrical impedance tomography (EIT), electrical capacitance tomography (ECT), ultrasonic computed tomography (UCT), gamma densitometry tomography (GDT), X-ray radiography/stereography (fluoroscopy), and X-ray computed tomography (CT) [6,7]. Some of these techniques have been applied to fluidized beds; however, X-ray imaging techniques have recently gained popularity since they are safer than other nuclear based techniques, have high resolution, and have controllable energy.

X-ray CT is particularly useful in visualizing fluidized beds, and can provide a three dimensional time-averaged density map of the flow structure. It is possible to calculate time-averaged local gas holdup data (or inversely solids holdup) from CT data. This method has been applied to fluidized beds in the literature but the technology is still under development. Grassler et al. [8] studied local solid distributions in circulating fluidized beds using X-ray computed tomography and was able to accurately measure solids concentrations up to 20 vol% with a spatial resolution of 0.2 mm and error range of only 5%. The study noted that X-ray CT has many advantages over capacitance probes, optical probes, and EIT because it does not affect the flow structure, is applicable at high temperatures, and can tolerate static electricity buildup. Kantzas et al. [9] used computed tomography to quantify channeling in a fluidized bed of polyethylene resin at various heights and superficial gas velocities. The study illustrated the usefulness of CT as a tool in determining gas holdup, especially in high-voidage channels. It was found that increased gas velocity increased the channeling in the resins, and a variety of channels could be formed depending on flow conditions, bed height, and resin type. Similarly, Wu et al. [10] employed CT to characterize fluidization of polyethylene resins in three fluidized beds with different diameters. Gas holdup was extracted from the CT data and it was found that fluidization hydrodynamics can be significantly affected by bed scale.

X-ray radiography/stereography and CT imaging are possible using the Iowa State University XFloViz facility. The XFloViz radiography/stereography system is capable of capturing dynamic responses in the bed due to its good temporal resolution, while the CT system captures local time-averaged three-dimensional volumetric images with excellent spatial resolution [6,11]. Previous work in the XFloViz facility, described by Franka et al. [12], used CT capabilities to image fluidized beds of glass beads, ground walnut shell, ground corn cob, and melamine plastic. While the local gas holdup was not calculated for the beds, time-averaged flow structures were found for each bed at three flow conditions, and qualitative comparisons were made. The CT imaging showed glass beads fluidize uniformly and channeling was evident in the lower density materials. CT also showed that increasing superficial gas velocity increased the fluidization uniformity.

To aid in the understanding of biomass injection, fluidized bed computational models are being developed to simulate the injection of biomass particles into fluidized beds. By experimentally tracking biomass particles, these computational fluid dynamics (CFD) simulations can be validated; the validated models can then be used to potentially enhance process efficiency, especially in gasification. Some CFD simulations exist for fluidized bed behavior; however, biomass injection has not been simulated in the literature. Deza et al. [13] modeled the fluidized bed described by Franka et al. [12] using MFIX CFD software. The CFD simulations of glass bead and walnut shell fluidization were successfully validated by the experimental X-ray data.

Scale-up is one of the main difficulties in simulating fluidized beds since wall effects become more pronounced with decreased bed diameter, and because initial predictions of the complex hydrodynamics cannot accurately be made [10,14,15]. Van Ommeren et al. [14] applied CFD simulations to 15 and 30 cm diameter fluidized beds to investigate these scaling problems associated with fluidized beds. While the simulation had some success in using various parameter sets to model fluidization, it was found that scaling laws remain insufficient in describing scale-up of fluidized beds.

One of the most important fundamental parameters for designing, analyzing, and simulating fluidized beds is the minimum fluidization velocity, \( U_{mf} \). It sets the lower boundary for fluidization and is necessary when modeling the hydrodynamics using CFD [16]. \( U_{mf} \) is a complex function of particle properties/geometry, fluid properties, and bed geometry, and may be calculated using correlations from the literature [2]. \( U_{mf} \) is generally determined experimentally since many of the parameters used in theoretical calculations can only be estimated [17]. Hilal et al. [16] analyzed the effects of bed diameter, distributor design, and inserts on \( U_{mf} \). The study found that \( U_{mf} \) decreased with increasing bed diameter, increased with decreasing distributor plate hole pitch, and decreased with increasing number of vertical inserts. Similarly, Wu et al. [10] found \( U_{mf} \) of polyethylene resins in three different fluidized beds with varying diameters. While the study was not focused on minimum fluidization research, it was found that \( U_{mf} \) increased with decreasing bed diameter. It was hypothesized that higher friction forces in small-scale beds caused an increase in the velocity.

Another important factor in fluidization behavior is the bed material. In many industrial processes, fluidized beds are composed of sand due to its uniform fluidization and heat transfer properties; however, beds made solely of biomass can also fluidize. Fluidized bed models, or cold flow beds, are generally comprised of uniform diameter and density glass beads which represent uniformly fluidizing bed material [1,3,5,8,18,19]. Glass beads are useful in achieving uniform fluidization because of their high sphericity, uniform density, and resistance to breakage.

One problem with X-ray imaging glass bead fluidized beds, however, is that X-ray penetration through the center of a
bed often saturates the column's edges, resulting in a loss of resolution. This occurs because the X-ray attenuation of glass beads is much greater than that of the surrounding air, so higher energy X-rays are required to penetrate the flow. Because attenuation increases with density, it is desirable to use bed materials that are less dense than glass to resolve the flow, while maintaining uniform fluidization. However, varying density and size can pose problems since not all materials fluidize uniformly. Fluidization hydrodynamics is governed by particle size and density according to Geldart's classification [20]. Generally, Geldart type-B particles are fluidized in gasification. Additionally, only Geldart type-B biomass satisfactorily fluidizes [21]. Ground walnut shell and ground corncob, corresponding to Geldart type-B particles, have much lower densities than glass beads but similar fluidization behavior [12]. Consequently, it is useful to image through biomass fluidized beds to gain a better understanding of fluidization behavior.

The goal of this research is to use minimum fluidization data and X-ray CT imaging to understand the effects of bed material, flow rate, and side air injection on fluidization hydrodynamics in two cold flow fluidized bed reactors. From the data, it is hoped to gain a greater understanding of the scale-up of fluidized beds with side air injection. Additionally, the study will provide experimental data for use in CFD model validation.

**EXPERIMENTAL SETUP**

**Fluidized Beds and Data Acquisition**

Two cold flow fluidized bed reactors with side ports for particle and gas injection are used in this study. The design of both reactors is nearly identical, with the exception of the diameter. The smaller reactor has an internal diameter (ID) of 10.2 cm while the larger reactor has a 15.2 cm ID. Additionally, nearly all components are fabricated from either acrylic or nylon to allow X-ray and optical penetration. Figure 1 shows a schematic of the two reactors.

The 10.2 cm ID reactor consists of three main chambers: a top chamber (a), a bed chamber (c), and a plenum (e). Bed material is fluidized in the 30.5 cm tall, 10.2 cm ID bed chamber. The bed chamber includes one injection port (k) which is located in one of the four square bosses (j) on the sidewall, 1.27 cm above the bottom of the bed chamber. This port allows particle and gas injection during fluidization. During injection experiments, a 1.91 cm UNF nylon Swagelok fitting is used as an entry port fitting.

Below the bed chamber rests the distributor plate (d). The 1.27 cm thick acrylic plate has 62, 1 mm diameter holes spaced approximately 1.27 cm in a circular grid for a total open area of 0.60%. To eliminate bed particles from lodging inside the distributor plate orifices, a 45 mesh screen with openings of 0.04 cm is attached to the plate using silicone adhesive; silicone prevents gas from escaping through the sides of the screen. The metallic screen does not affect X-ray data collection since all imaging is done above the distributor plate. The 15.2 cm tall, 10.2 cm diameter plenum is located directly under the distributor plate and is filled with glass marbles (i). The marbles allow uniform distribution of the fluidizing gas throughout the plenum before reaching the distributor plate. Gas enters the plenum through the air inlet hole (h), which is located in the center of the air inlet plate (f). A section of screen with openings of 0.127 cm is located above the air inlet hole to keep marbles from plugging the inlet hole. A 0.64 cm NPT threaded pressure tap (g) is located off-center on the air inlet plate.

Above the bed chamber is a 61 cm tall, 10.2 cm ID top chamber which prevents particle elutriation. This chamber is only necessary under high gas velocity conditions. Identical square flanges are located on the top and bottom of the plenum chamber, bed chamber, and top chamber to connect the sections together. To seal the reactor from gas leakage, four rubber gaskets (b) are placed between each flange. The reactor is assembled by bolting flanges in each section with eight 1.27 cm diameter nylon bolts and 16 nylon washers.

The 15.2 cm ID reactor is nearly identical to the 10.2 cm ID reactor, but scaled up in the radial direction. The distributor in the 15.2 cm ID reactor has an aerating plate with an open area ratio of 0.57%, compared with 0.60% for the 10.2 cm ID reactor. Additionally, the 15.2 cm ID reactor only has 2 bosses located on the bed chamber walls but has an additional two bosses on the plenum. A pressure tap is located in one of these plenum bosses instead of in the air inlet plate. Pressure in both reactors is measured with a Dwyer 0-34.5 kPa (5 psig), 24 vdc pressure transducer connected to a pressure tap in the plenums.
The pressure transducer has a maximum error of ±0.25% of the full scale (± 86 Pa).

Compressed air from the laboratory supply serves to fluidize the various beds. This air is also injected through the side port. Air flow to the plenum is controlled through a series of ball valves, a pressure regulator, and one of two flow meters; a 0-500 Lpm stainless steel Aalborg GFM671S flow meter is used in high gas velocity applications, and a 0-200 Lpm Aalborg GFM571 flow meter is used in lower gas velocity applications to improve resolution. Similarly, air flow through the injector port is controlled through a series of ball valves, a pressure regulator, a solenoid control valve, and one of two flow meters; a 0-100 Lpm Aalborg GFM471 flow meter is used for high side air flow conditions, and a stainless steel 0-30 Lpm Aalborg GFM371S flow meter is used for low side air flow applications to improve resolution. Error on the flow meters is ±2% of the full scale.

The flow meters and pressure transducers are interfaced with a data acquisition (DAQ) system connected to a computer. The DAQ system, described by Jones [22], is composed of a National Instruments 6030 E series multifunction data acquisition card, a National Instruments SCB-68 shielded connector block, a 24 vdc Cole-Parmer PS2-24-15-007 instrument power supply, and National Instruments LabView 8.0 acquisition software. A LabView VI records the average pressure and flow measurements at a sample rate of 1000 Hz per sensor. Data are then output to Excel files for manipulation.

Material Selection, Minimum Fluidization Velocity, and Flow Conditions

The fluidization hydrodynamics of three bed materials are compared in this study: glass beads, ground walnut shell, and ground corncob. These Geldart type-B materials are chosen according to previous work [12]. The static bed height is set to one column diameter (H = D) for all tests and the weight of each bed is measured. From the bed volume and weight, the bulk densities of each material are calculated. Table 1 provides a summary of the bed characteristics in this study.

<table>
<thead>
<tr>
<th>TABLE 1: BED MATERIAL PROPERTIES.</th>
<th>Glass Beads</th>
<th>Walnut Shell</th>
<th>Corncob</th>
</tr>
</thead>
<tbody>
<tr>
<td>Reactor ID (cm)</td>
<td>5.2</td>
<td>5.2</td>
<td>5.2</td>
</tr>
<tr>
<td>Particle density (kg/m³)</td>
<td>2600</td>
<td>2600</td>
<td>1180</td>
</tr>
<tr>
<td>Particle diameter (μm)</td>
<td>500-600</td>
<td>500-600</td>
<td>660</td>
</tr>
<tr>
<td>Bed height (cm)</td>
<td>10.2</td>
<td>10.2</td>
<td>10.2</td>
</tr>
<tr>
<td>Bed weight (g)</td>
<td>1180</td>
<td>4158</td>
<td>1496</td>
</tr>
<tr>
<td>Bulk density (kg/m³)</td>
<td>1481</td>
<td>196</td>
<td>579</td>
</tr>
<tr>
<td>Bulk void fraction (ε)</td>
<td>0.43</td>
<td>0.42</td>
<td>0.58</td>
</tr>
</tbody>
</table>

The minimum fluidization, Umf, is one of the most important fundamental parameters in fluidization hydrodynamics and is used to normalize flow conditions between the reactors. In addition, one of the main purposes of this study is to examine how minimum fluidization is affected by bed geometry, material, and side air injection. Umf is experimentally determined for each bed material, reactor, and side air flow condition. Initially, beds are fluidized with air at 200 Lpm in the 10.2 cm column, and 300 Lpm in the 15.2 cm column. The pressure drop across the plenum, distributor plate, and bed is subsequently measured. Gas flow is then decreased in increments of 5 Lpm and pressure is again measured. By measuring pressure as the gas flow rate decreased, initial packing effects of the bed material are removed. This procedure is repeated for a reactor with no bed material in order to determine the pressure drop across the plenum and distributor plate. By interpolating and subtracting the empty reactor pressures from the fluidized bed pressures at each gas flow rate (or superficial gas velocity), the pressure drop across the bed is calculated. The minimum fluidization velocity is defined as the point where the pressure drop across the bed stops increasing linearly with gas velocity, but remains constant. To illustrate, Figure 2 shows an example of an experimental fluidization curve for glass beads in the 10.2 cm reactor, with no side air injection. The plot clearly shows a linear increase in pressure with increasing Ug until reaching Umf, at which point the pressure drop is constant with increasing Ug. For this example, Umf = 21.3 cm/s.

To determine the effects of superficial gas velocity and side air injection on fluidization hydrodynamics, a variety of flow conditions are examined in this research. For this study, five side air flow rates, Qs, are investigated for each bed; Qs = 0, 0.05, 0.10, 0.15, and 0.20Qmfs. These values are based off the minimum fluidization air flow rates with no side air injection, Qmfs. In CT imaging, only side air flow conditions Qs = 0, 0.05, and 0.10Qmfs are evaluated. Additionally, for CT imaging two superficial gas velocities, Ug, are studied which are referenced to the minimum fluidization velocity without side air injection for each bed; Ug = 1.5Umfs, and Ug = 2Umfs.

XFloViz Facility

The Iowa State University XFloViz facility is used to image the fluidized beds and has been described in detail elsewhere [11]. Consequently, only a brief summary is presented here. Two LORAD LPX200 portable X-ray tubes provide 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 combinations of 1 mm thick copper and aluminum filters, depending on the object of interest. The XFloViz imaging...
room, which is where X-ray imaging is performed, is shown in figure 3.

![Figure 3: XFLOVIZ Imaging Room](image)

Located opposite each X-ray source is a corresponding X-ray detector which consists of a detector and CCD camera pair. Three detectors are available in XFLOViz facility, and are interchangeable to allow for varying visualization techniques. The first two detectors consist of identical 40.6 cm diameter Precise Optics PS164X image intensifier screens with a 35.0 mm output image diameters coupled to DVC-1412 monochrome digital CCD cameras. These detectors are primarily used for radiographic imaging, due to their relatively high temporal resolutions ranging from 10 to 60 frames per second (fps), depending on binning options. Despite the usefulness of radiographic data in understanding fluidized bed hydrodynamics, the image intensifier detectors are not utilized in this study. Franka et al. [12] describes radiographic imaging of fluidized beds in the XFLOViz facility.

The third detector in the XFLOViz facility is primarily used for CT imaging because of its high spatial resolution. This detector is composed of a square 44×44 cm cesium-iodide scintillator screen which transforms radiation into visible light. A 50 mm Nikon lens with variable exposure rates captures images on the scintillator which are subsequently digitized by an Apogee Alta U9 CCD system. This system has 3072×2048 active pixels and is thermoelectrically cooled to allow long exposure times with low noise. Adjustable binning sizes combine pixels into “effective pixel” clusters which can be used to decrease file size and acquisition time. In this study, an exposure time of 1 second with 4×4 binning is used to minimize acquisition time while maintaining X-ray signal strength. Hence, the resulting spatial resolution in the fluidized bed is ~0.6 mm.

Both X-ray detectors and sources are located on a 1.0 m ID rotation ring that can rotate 360° around the object of interest. This allows various imaging orientations without affecting the object being imaged. Data from the CCD cameras are acquired with software developed by Iowa State University’s Center for Nondestructive Evaluation (CNDE) and stored on a personal computer. The software allows for control of both detector/camera pairs, and provides rotation ring motion control.

In this study, two calibrations are applied to the CT data to remove artifacts induced by the imaging hardware [11]. A normalization algorithm accounts for non-uniform pixel response to incident X-ray energy. The normalization calibration utilizes a linear interpolation method to scale pixel response during acquisition. The second calibration is only applied to glass bead fluidized beds, and accounts for beam hardening. This artifact is a result of the preferential attenuation of polychromatic X-rays in high density material, and causes reconstructed objects to appear less dense in the center than in the edges. To correct for beam hardening, an “effective attenuation” calibration is applied to raw CT files before reconstruction; this requires knowledge of X-ray attenuation as a function of material thickness. A radiograph of a glass calibration wedge is used to calculate the required calibration parameters. Once calibrations have been applied, volumetric reconstruction of CT data is performed through a filtered back projection algorithm using 14 nodes of CNDE’s 64-node LINUX cluster. This provides 3-D time-averaged fluidization data.

### Gas Holdup and CT Images

In order to quantify CT data, the time-averaged local gas holdup is calculated for the various fluidization test conditions. Gas holdup (void fraction or volumetric gas fraction) is a ratio describing the amount of void space in the bulk material. It is particularly useful in understanding heat and mass transfer phenomena. For each material and reactor combination, two U conditions are examined, \( U_g = 1.5U_{mf,0} \) and \( 2U_{mf,0} \), as well as three \( U_m \) conditions, \( Q_m = 0Q_{mf,0}, 0.05Q_{mf,0}, \) and \( 0.1Q_{mf,0} \).

Since CT intensity is proportional to X-ray attenuation, the local time-averaged gas holdup (\( \varepsilon_g \)) is determined from local CT intensities for the fluidization test conditions (CT). Also, local CT intensities for the bulk material (CT_b) and for the reactor without material (CT_g) are necessary for calibrations. Consequently, for each fluidization test condition, CTS of the fluidizing bed, the bulk material, and the empty reactor are necessary. These CTS must all have identical X-ray power settings and CT acquisition parameters. The following equation is then applied to each 3-D pixel (voxel) using the three CT files and the resulting local gas holdup voxels are recompiled into a 3-D image.

\[
\varepsilon_g = \frac{CT - CT_b + (CT_g - CT)(\varepsilon_{g,b})}{CT - CT_b}
\]

where the void fraction of bulk material is shown in Table 1 and is defined as:

\[
\varepsilon_{g,b} = 1 - \frac{\rho_b}{\rho_p}
\]

Bulk density (\( \rho_b \)) is determined experimentally, while particle density (\( \rho_p \)) was provided by the material manufacturer. In addition to finding gas holdup, the local calculations reduce some artifacts created by the bed geometry. Noise is reduced with a smoothing method that averages neighboring voxels.

Three-dimensional images are viewed using internally developed software, which allows viewing of the 3-D images at any location within the imaging volume. Additionally, the program captures “slices” of the 3-D image to produce 2-D

![Image](image)
Multiple tests ensure repeatability and improve the accuracy of the data. The software also applies a colorizing routine to the slice images to enhance certain features. The colorizing scales can be independently adjusted for each material to improve image contrast. In the resulting images, low gas holdup corresponds to more bed material, while high gas holdup corresponds to high air flow.

**FIGURE 4: CT IMAGING PLANES.**

**RESULTS AND DISCUSSION**

**Minimum Fluidization Velocity**

The minimum fluidization velocity without side air injection \( (U_{mf,0}) \) is experimentally determined according to the procedure outlined previously. The minimum fluidization air flow rate without side air injection \( (Q_{mf,0}) \) is then calculated from \( U_{mf,0} \). Multiple tests ensure repeatability and improve the accuracy of the data. Table 2 shows \( U_{mf,0} \) and \( Q_{mf,0} \) for each bed in both reactors with the \( \pm \) band representing one standard deviation from 8-10 replicates. Note that \( U_{mf} \) corresponds to minimum fluidization velocity while \( U_{mf,0} \) corresponds to the minimum fluidization velocity without side air injection.

**TABLE 2: \( U_{mf,0} \) AND \( Q_{mf,0} \) FOR ALL MATERIALS IN BOTH REACTORS.**

<table>
<thead>
<tr>
<th>Material</th>
<th>( U_{mf,0} ) (cm/s)</th>
<th>( Q_{mf,0} ) (L/min)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Glass beads, 10.2 cm FB</td>
<td>21.7 ± 0.5</td>
<td>105.0</td>
</tr>
<tr>
<td>Walnut shell, 10.2 cm FB</td>
<td>18.4 ± 0.6</td>
<td>89.0</td>
</tr>
<tr>
<td>Corncob, 10.2 cm FB</td>
<td>17.1 ± 0.3</td>
<td>83.0</td>
</tr>
<tr>
<td>Glass beads, 15.2 cm FB</td>
<td>20.2 ± 0.4</td>
<td>220.0</td>
</tr>
<tr>
<td>Walnut shell, 15.2 cm FB</td>
<td>18.1 ± 0.2</td>
<td>197.0</td>
</tr>
<tr>
<td>Corncob, 15.2 cm FB</td>
<td>16.4 ± 0.3</td>
<td>179.0</td>
</tr>
</tbody>
</table>

From the table it can be seen that for both beds, \( U_{mf,0} \) is largest for glass beads, followed by walnut shell, and lastly by corncob. This trend may be partly due to density since increased density appears to increase \( U_{mf,0} \); however, \( U_{mf,0} \) relies on many factors such as bulk density and particle sphericity, so the effects of density alone cannot be determined. It is also observed that \( U_{mf,0} \) decreases (slightly) with increasing bed diameter, which has also been observed by others [10, 16] and this is independent of bed material.

The \( U_{mf,0} \) tests without side air injection also show that the high density materials exhibit a larger bed pressure drop than the low density materials. Also, higher pressure drops are recorded in the 15.2 cm beds than in the 10.2 cm beds. Increasing the material density or volume increases the bed weight. In order for fluidization to occur, the force of the fluidizing air must overcome the weight of the bed and, consequently, larger pressure drops are required to fluidize high density materials or large bed volumes. To illustrate, Figure 5 shows the pressure drop between fluidization curves for glass beads, walnut shell, and corncob without side air injection in both reactors.

**FIGURE 5: EXPERIMENTAL FLUIDIZATION CURVES FOR ALL MATERIALS IN BOTH REACTORS WITH \( Q_g = 0 \).**

Since testing is performed with side air injection, it is also desired to see how side air injection affected the minimum fluidization velocity. For each material and reactor combination, air is injected at the desired flow rates and \( U_{mf} \) is determined. Each condition is tested multiple times to improve accuracy and repeatability. The fluidization curves with side air injection are more difficult to analyze than the curves generated without side air; as \( Q_g \) increases, the slope of the curves becomes more gradual, and thus, \( U_{mf} \) becomes increasingly difficult to determine. In addition, with large values of \( Q_g \), there appears to be a non-linear region introduced in the fluidization curves below \( U_{mf} \) and this is observed in both beds for all three materials. It is hypothesized that with side air injection, fluidized beds undergo two fluidization points. Because side air decreases the particle friction in the bed, the bed undergoes partial fluidization with \( U_p \) well below \( U_{mf} \). As \( U_p \) increases past the partial fluidization point, the pressure-velocity slope becomes more gradual until reaching a constant pressure, which is the fully fluidized \( U_{mf} \). For the purposes of this study, this fully fluidized superficial gas velocity is recorded as \( U_{mf} \).

Figure 6 identifies the partial fluidization phenomena on the experimental fluidization curves for glass beads, walnut shell, and corncob in both reactors with \( Q_g = 0.20Q_{mf,0} \). Note that the pressure scale in Figure 6 has been normalized to the maximum pressure in each set of data.
reported and arranged in the following figures. Each figure shows a collection of gas holdup images in a given material with a specific $U_g$ and three $Q_s$ conditions. Both reactors are shown in the figures; the three image groups on the left correspond to the 10.2 cm reactor while the three image groups on the right correspond to the 15.2 cm reactor. The 15.2 cm reactor images are also scaled to match the 10.2 cm reactor images for easier comparison between geometries. The image edges represent the walls of the fluidized beds; the bottom of each y-slice represents the top of the distributor plate and the edges of the y- and z- slices represent the bed chamber walls. Z-slices are captured at the top of the injection ports in both reactors, 3.2 cm above the distributor plates, and at 9.0 cm in the 10.2 cm reactor and 13.4 cm in the 15.2 cm reactor, corresponding to H/D = 0.88. Horizontal dashed lines on the y-slices show the locations of the z-slices, and the “grayed” region on the bottom-right of the y-slices show the location of the injection port. At the top of each collection of slices, is the color scale corresponding to gas holdup. The gas holdup color scales are identical for all slices in a material but are different between materials. These are modified individually here for each material in order to improve color resolution. As a result, the bed materials cannot be directly compared by color; however, the actual gas holdup values are unaffected.

Figure 8 shows the local time-averaged gas holdup for glass beads with $U_g = 1.5U_{mf,0}$ and highlights the effects of increasing side air injection on fluidization behavior. With no side air injection, y- and z-slices in both reactors show that the beds generally fluidize uniformly; however, it can be noticed that drag from the injection port causes slight asymmetry in the bed hydrodynamics, even without side air injection. In the y-slice images for $Q_s = 0Q_{mf,0}$, a region of relatively high gas holdup is observed directly above the injection port. Additionally, the z-slices at 3.2 cm show gas distribution from the distributor plate primarily occurs near the wall in both reactors.

Bed hydrodynamics are influenced by side air injection. With side air injection, a path of high gas holdup extends from

**Effects of Side Air Injection**

For each 3-D gas holdup data set, y- and z-slices are reported and arranged in the following figures. Each figure

Average $U_{mf}$ for each side air injection condition is plotted in Figure 7. Comparing fluidization between the two reactors, it is evident that side air injection affects the two beds similarly. While there are differences in $U_{mf}$ between the materials for a given $Q_s$, the data do not appear to be significantly different between the two reactors. It is also observed that for beds of walnut shell and corncob, as $Q_s$ increases, $U_{mf}$ decreases. With increasing $Q_s$, more air is introduced into the system, reducing the friction between the bed particles and decreasing $U_{mf}$. Also, the corncob bed shows the largest drop in $U_{mf}$ under the five side air conditions. This may be partly due to the low density of corncob. It is likely that $Q_s$ has a greater impact on reducing the pressure required to fluidize the bed than for glass beads. This could also explain why $U_{mf}$ for beds of glass beads did not significantly vary with $Q_s$; a significant pressure is required to fluidize the heavier glass bed, regardless of $Q_s$. Error in $U_{mf}$ calculations is approximately ±1 cm/s.

Gas Holdup

For each 3-D gas holdup data set, y- and z-slices are reported and arranged in the following figures. Each figure

![Figure 7: $U_{mf}$ with varying $Q_s$ for all materials in both reactors.](image)

![Figure 6: Experimental fluidization curves for all materials in both reactors with $Q_s = 0.20Q_{mf,0}$.](image)

\[ U_g (\text{cm/s}) \]

\[ 0.00 \ 0.05 \ 0.10 \ 0.15 \ 0.20 \ 0.25 \]

\[ U_{mf}(\text{cm/s}) \]

\[ 5.0 \ 10.0 \ 15.0 \ 20.0 \ 25.0 \]

Side air ratio ($Q_s/Q_{mf,0}$)

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the injector to the bed surface. This air path gradually expands into the bed as it rises up the bed, indicating that horizontal dispersion increases with axial height. This effect is captured in the z-slices; at z = 3.2 cm, a small region of high gas holdup occurs at the injector while at z = 9.0 cm, the high gas holdup region expands over much of the bed. As Q increases, the injector air path extends further into the bed due to a higher side air velocity. Side air injection also produces a non-uniform average bed height above the injector. The highest average bed height occurs near the wall, directly above the injector. As Q increases, the bed height above the injector increases while the surrounding bed remains approximately constant. While the y-slice images show non-symmetrical bed fluidization with side air injection, the z-slices show that fluidization remains somewhat symmetric about the x-axis.

Another structure shown in the \( U_g = 2U_{mf,0} \) y-slices is a region of lower gas holdup on the edge of the column across from the injector and slightly below the bed surface. This region can also be seen in the 9.0 cm z-slices and shows the low gas holdup spans a section of the wall. This feature can be attributed to bubble coalescence; bubbles tend to coalesce in the center of the column after reaching a certain height and thus, air flow along the column wall is reduced. The flow structures caused by the fluidizing air seem to be unaffected by side air injection, except by the injector port where the injection air flow path can be seen.

In addition to the flow structures caused by the fluidizing air, \( U_g \) also significantly affects the side air injection flow path. By increasing \( U_g \), side air penetrates further into the bed. In the \( U_g = 1.5U_{mf,0} \) images, side air follows the reactor chamber wall vertically and does not appear to penetrate the bed except near the top, while in the \( U_g = 2U_{mf,0} \) y-slices, side air expands further into the bed. The increase in side air penetration occurs because of an increase in the local gas holdup throughout the bed. With a high gas holdup, less force is required for the side air to penetrate the bed material. Increasing \( U_g \) also changes the boundary profiles of the side air flow paths. For \( U_g = 1.5U_{mf,0} \), the side air paths are very distinguishable for each \( Q \). Increasing \( U_g \) to \( 2U_{mf,0} \) shows much less distinguishable side air flow paths. This also shows that by increasing \( U_g \), fluidization becomes more uniform, and the effects of side air injection are greatly reduced. The reduced effect of \( Q \) occurs because the percentage of the side air flow relative to the total air flow through the entire system (fluidizing air and side air) is greatly reduced as \( U_g \) increases.

Superficial gas velocity is also shown to affect the region directly above the distributor plate. The y-slices show regions of high gas holdup extending from the distributor plate into the bed, which indicate the air jets from the distributor plate orifices. These jets are caused by a high gas velocity passing through each orifice hole in the distributor plate. Bed material between these orifices has slightly lower gas holdup than the surroundings showing that fluidization is not uniform at the surface of the distributor plate. As with the side air injection, by increasing \( U_g \), the penetration of the air from the distributor plate increases.

The gas holdup images also show that increasing \( U_g \) also increases the average bed height. Additionally, increasing \( U_g \) causes the top of the bed to become less distinguishable. The y-slice images for \( U_g = 1.5U_{mf,0} \) clearly show the upper bed boundaries whereas the average bed height is much less distinguishable for \( U_g = 2U_{mf,0} \). This indicates that for higher \( U_g \), larger bubbles are breaking the surface of the fluidized bed. Also, the y-slices for \( U_g = 1.5U_{mf,0} \) show that the top of the beds are approximately level with the exception of the region above the injector port. Increasing \( U_g \) to \( 2U_{mf,0} \) shows areas of lower gas holdup located near the reactor walls. This phenomenon is again caused by bubbles breaking in the center of the bed surface; as entrained bed particles are ejected by the bubbles, they fall towards the reactor wall and reenter the bed.

**Effects of Bed Materials**

Figures 10-13 show the effects of bed material on fluidization hydrodynamics and gas holdup. It should be noted...
that the color mapping scales are not consistent between the materials to allow for better color resolution, so the analysis presented here is not based on the image absolute colors. The images show that $Q_g$ and $U_8$ have similar effects on fluidization; however a few differences exist between the materials. It is evident that as the density of the material decreases, the gas holdup increases. The glass bead beds (Figures 8 and 9) have a lower gas holdup than both walnut shell and corncob beds while corncob has a higher gas holdup than walnut shell beds. This effect is also related to the bulk gas holdup for each material: $e_{gb}$ of glass is lower than walnut shell which is in turn lower than corncob (see Table 1).

The images also show the effects of material on noise in the calculations. Since the attenuation of glass beads is much higher than air, the signals obtained from glass bead CTs are much stronger and have higher intensity resolution compared to the surrounding air. In walnut shell and corncob beds, the attenuation difference between the bed material and the surrounding air is much less than with glass. This reduces the intensity resolution of the CT data. In addition, glass beads are extremely homogeneous and the bulk glass CT data needed in gas holdup calculations (Eq. (1)) has high uniformity. As a result, very little noise is introduced in the calculations by the bulk glass CT. In contrast, the bulk CT files for walnut shell and corncob are much more heterogeneous and, consequently, noise is introduced into the gas holdup calculations. For example, all y-slice images of corncob in the 10.2 cm bed show small circular regions of low gas holdup located along the left side of the image above the horizontal dashed line. These features can be attributed to noise and should be disregarded when making comparisons.
In general, the fluidization behavior between all three materials is similar; however side air appears to affect the beds differently. In glass beads, the side air path is very clearly defined for all Qs conditions. For walnut shell beds, the side air flow path is distinguishable, but not as well defined as for glass beads. In the corncob beds, the side air flow path is even less distinct. This illustrates another effect of the material. With side air injection, the corncob fluidizes most uniformly followed by walnut shell and lastly by glass beads. It is evident that side air affects the lower density corncob and walnut shells less than the higher density glass beads.

**Effects of Reactor Geometry**

By comparing the gas holdup images for the 10.2 cm reactor to the 15.2 cm reactor images, it is evident that fluidization hydrodynamics appear to be very similar, regardless of reactor geometry, for the two bed diameters studied here. When scaled by the column diameter, the profiles of the injector air flow appear to be consistent as Qs changes. Additionally, the bed heights are similar under similar flow conditions. Internal flow features mentioned previously are also shown in both reactors, although these differ slightly due to noise and randomness in the experimental data. While local differences may be observed between reactors, the general trends are consistent between the reactors and no significant fluidization differences are noticed.

**CONCLUSIONS**

Minimum fluidization velocity was analyzed for three materials in two different reactors. Without side air injection, glass bead beds were found to have the highest \( U_{mf,0} \) followed by walnut shell and corncob. Side air injection was found to decrease \( U_{mf} \) in walnut shell and corncob beds, likely due to the reduction in pressure necessary to fluidize the bed. In glass bead beds, side air did not significantly affect \( U_{mf} \). Side air injection was also found to influence the pressure-gas velocity fluidization curves for all materials. At high Qs, a partial fluidization location was observed at \( U_g \) under \( U_{mf} \). Differences in reactor geometry were not found to significantly affect \( U_{mf} \).

X-ray computed tomography was also applied to fluidized beds in order to determine the effects of side air injection, superficial gas velocity, bed material, and reactor geometry on fluidization behavior. From the CT data, local time-averaged gas holdup was calculated for the fluidized beds. It was found that gas fluidization was approximately uniform without side air injection. With side air injection, the side air flow was found to preferentially rise near the side of the bed but gradually expanded into the bed as height increased, and this expansion increased with increasing Qs. By increasing \( U_g \), the beds were found to fluidize more uniformly and the effects of side air injection were less pronounced. In addition, increases in \( U_g \) increased the bed expansion and the overall gas holdup in the system. \( U_g \) was also found to affect the internal flow structure; at \( U_g = 2U_{mf,0} \), air flow was concentrated around the bottom edges of the bed leaving a region of low gas holdup in the bed center, however, the air flow gradually migrated toward the top center of the bed. Penetration of the air from the distributor plate also increased with higher \( U_g \). Various materials were found to have similar fluidization behaviors with a few notable differences. Fluidization with side air injection was most uniform in corncob beds, followed by walnut shell and glass bead beds, indicating Qs had the least effect on the lower density materials. Also, glass bead beds showed the lowest gas holdup followed by walnut shell and corncob beds. More noise was shown in the corncob and walnut beds than in glass bead beds due to a decrease in X-ray signal strength in the lower density materials and the heterogeneous nature of bulk corncob and walnut shell. Comparisons between reactor geometries showed very few differences between the 10.2 cm and the 15.2 cm reactors.

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