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Bed height and material density effects on fluidized bed hydrodynamics

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ABSTRACT

Characterizing the hydrodynamics of a fluidized bed is of vital importance to understanding the behavior of this multiphase flow system. Minimum fluidization velocity and gas holdup are two of these key characteristics. Experimental studies addressing the effects of bed height and material density on the minimum fluidization velocity and gas holdup were carried out in this study using a 10.2 cm diameter cylindrical fluidized bed. Three different Geldart type-B particles were tested: glass beads, ground walnut shell, and ground corncob, with material densities of 2600, 1300, and 1000 kg/m^3, respectively. The particle size range was selected to be the same for all three materials and corresponded to 500–600 μm. In this study, five different bed height-to-diameter ratios were investigated: $H/D=0.5$, 1, 1.5, 2, and 3. Minimum fluidization velocity was determined for each H/D ratio using pressure drop measurements. Local time-average gas holdup was determined using non-invasive X-ray computed tomography imaging. Results show that minimum fluidization velocity is not affected by the change in bed height. However, as the material density increased, the minimum fluidization velocity increased. Finally, local time-average gas holdup values revealed that bed hydrodynamics were similar for all bed heights, but differed when the material density was changed.

1. Introduction

Fluidized bed hydrodynamic behavior is very complex and must be understood to improve fluidized bed operations. One of the most important parameters to characterizing fluidized bed conditions is the minimum fluidization velocity, U_{mf} (Ramos Caicedo et al., 2002), which is proportional to the drag force needed to attain solid suspension in the gas phase. The minimum fluidization velocity also constitutes a reference for evaluating fluidization intensity when the bed is operated at higher gas velocities (Zhong et al., 2008). In general, U_{mf} is a function of particle properties/geometry, fluid properties, and bed geometry.

Sau et al. (2007) determined the minimum fluidization velocity for a gas–solid system in a tapered fluidized bed and studied the effects of bed geometry, specifically the tapered angle, on the minimum fluidization velocity. Results showed that as the tapered angle increased, U_{mf} increased, which implied a dependence of the minimum fluidization velocity to the geometry of the fluidized bed. Moreover, Hilal et al. (2001) concluded that both the bed diameter and the type and geometry of the gas distributor affected U_{mf}.

The influence of bed height on minimum fluidization velocity has been studied using different types of fluidized beds. Zhong et al. (2006) concluded that the static bed height for a 2D spouted bed influenced the minimum spouting fluidization velocity; increasing the bed height increased the spouting velocity. Sau et al. (2007) used a conical tapered fluidized bed to find the minimum fluidization velocity and the pressure drop across the bed. They concluded that U_{mf} was independent of bed height for their conical tapered fluidized bed.

Ramos Caicedo et al. (2002) studied the minimum fluidization velocity for gas–solid 2D fluidized beds operated over a range of conditions. They concluded that as the static bed height increased, U_{mf} increased. Moreover, Sánchez-Delgado et al. (2011) also studied the effects of bed height on the minimum fluidization velocity for a 2D fluidized bed, and concluded that there were negligible differences in U_{mf} as bed height changed, which did not corroborate the results obtained by Ramos Caicedo et al. (2002). Gunn and Hilal (1997) studied cylindrical fluidized beds and concluded that for all the material and experimental conditions used in their study, there was no significant change in the minimum fluidization velocity when the bed height was increased. Cranfield and Geldart (1974) studied the fluidization characteristics of large particles and showed that for 3D beds, the minimum fluidization velocity remained constant no matter the bed height.

Gas holdup (ε_g) is another important parameter that characterizes fluidization quality, homogenous mixing, and process efficiency in a fluidization system, and is defined as the volumetric fraction of gas

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present within the bed. Using an optical probe, Zhu et al. (2008) determined the solid volumetric fraction $(1-\varepsilon_g)$ in a gas–solid system for bubbling and turbulent fluidization regimes. The turbulent regime showed that solid concentrations were not uniform in the axial or radial direction. In the bubbling regime, the nonuniformity increased as the superficial gas velocity increased. Zhu et al. (2008) also studied the effects of static bed height on the solid concentration. Results showed that increasing the static bed height produced an increase in the solid concentrations mainly in the bed central region, while the wall region had no significant changes.

Du et al. (2003) measured the solid concentration for a turbulent fluidized bed. Results showed that at high gas velocities, especially in the turbulent regime, the cross-sectional solids holdup exhibited a radially symmetric distribution, while this was not the case for the bubbling regime. At low gas velocities in the bubbling regime, dispersed bubbles produced a lower solids concentration in the center of the bed. The asymmetric distribution of solids concentration was attributed to the spiral motion of bubbles in the bed.

X-ray computed tomography (CT) imaging was used by Grassler and Wirth (2000) to determine the solids concentration in a 0.19 m diameter circulating fluidized bed with 50–70 μm glass beads as the bed material. For an up-flow system, they showed that the radial solid concentration exhibited a parabolic shape with a maximum concentration close to the wall of the reactor and a minimum concentration in the center of the bed. In a down-flow system, the solid concentration distribution was much more complex and depended upon the gas–solid distributor operating conditions. Franka and Heindel (2009) also used X-ray CT imaging to determine local time-average gas holdup in a 10.2 cm fluidized bed. Using different materials (glass beads, ground corncob, and ground walnut shell), superficial gas velocities (U_g), and side air injection flow rates (Q_{side}), they identified the local variations in the bed hydrodynamics. They found that with side air injection, the side air flow rose near the wall but then expanded into the bed as height and Q_{side} increased. As U_g increased, the effects caused by the side air injection were less pronounced, and the overall gas holdup in the system increased. Fluidization among the different materials had similar behaviors with some notable differences. Side air injection was less influential in the less dense material and gas holdup was the lowest for the denser material. Finally, they demonstrated the usefulness of X-ray computed tomography in visualizing the internal features of fluidized beds.

In the current study, two of the particle types are natural (biomass) components. Fluidizing biomass particles is challenging due to their irregular size, shape, and density. Understanding the influence of these particular characteristics on the fluidized bed hydrodynamics is also important. Zhong et al. (2008) studied the effects of particle size, density, and shape on the minimum fluidization velocity using wood chips, mung beans, millet, corn stalks, and cotton stalks. In their study, they used a rectangular fluidized bed with a cross section of 0.4 m × 0.4 m and air as the fluidizing gas. They determined that for long, thin biomass types, the minimum fluidization velocity increased with increasing length-to-diameter (L/d_{pt}) ratio. Their experiments showed that biomass did not fluidize when $L/d_{pt} > 20$, indicating that the biomass size and shape affect fluidization.

In this paper, the effects of bed height and material density on the minimum fluidization velocity and local time-average gas holdup in a 3D cylindrical fluidized bed are investigated.

2. Experimental setup

The reactor used in these experiments is a 3D cold flow fluidized bed. The cylindrical fluidized bed was fabricated with 10.2 cm internal diameter (ID) acrylic tubes with a 0.64 cm wall thickness. As shown in Fig. 1, the reactor consists of three main chambers: the top chamber or freeboard region, the bed chamber, and the plenum. Fluidization occurs in the bed chamber, which is 30.5 cm tall and 10.2 cm ID. Square flanges (16.5 cm × 16.5 cm) connect each section. An aeration plate is located immediately below the bed chamber; it is fabricated from a 1.27 cm thick acrylic plate with 62, 1 mm diameter holes spaced approximately 1.27 cm apart in a circular grid for a total open area of 0.62%. A 45 mesh screen with 0.04 cm openings is attached to the plate using silicone adhesive to avoid material blocking the aeration holes.

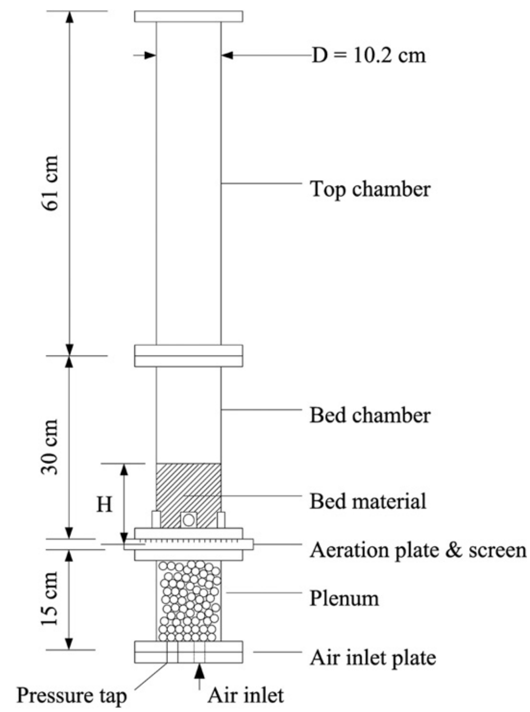

Fig. 1. Fluidized bed reactor (not to scale). The static bed height is identified by H.

Compressed air from the laboratory's building air supply is used as the fluidizing gas for this research. The fluidized bed air flow is regulated with a manual stainless steel pressure regulator and attached filter. The regulated air flows through two different mass flow meters: a 0–1000 Lpm stainless steel Aalborg GFM771 flow meter, which is used for high gas flow applications, and a 0–200 Lpm Aalborg GFM571 flow meter, used in lower gas flow applications. This allows for better measurement resolution. The flow through the respective mass flow meter is directed by ball valves. The mass flow meters for this study have a maximum error of 2% of full scale.

Pressure is measured with a Dwyer 0–34.5 kPa (0–5 psig), 4–20 mA output pressure transducer located in the bottom of the plenum. The signals obtained from the pressure transducer and mass flow meters are connected to a computer controlled data acquisition system. Pressure readings have an estimated error of 1–4%, with the largest error corresponding to the smallest pressure readings.

Computer controlled data acquisition software is used to record real-time pressure and flow rate measurements over a user-specified period, and then the average pressure and flow rate are calculated and recorded. Average measurements are necessary due to the highly variable pressure signal caused by the bubbling bed. In this study, data collection occurs at a rate of 1000 Hz for a time interval of 5 s; average pressure, and average flow rate are subsequently written to a data file.