Powder Technology 253 (2014) 657–676

Contents lists available at ScienceDirect

Powder Technology

journal homepage: www.elsevier.com/locate/powtec

Hydrodynamic comparison of spherical and cylindrical particles in a gas–liquid–solid fluidized bed at elevated pressure and high gas holdup conditions

Dominic Pjontek, Arturo Macchi *

Chemical and Biological Engineering Department, University of Ottawa, 161 Louis Pasteur, Ottawa, Ontario K1N 6N5, Canada

ARTICLE INFO

Article history:
Received 25 October 2013
Accepted 14 December 2013
Available online 21 December 2013

Keywords:
Fluidization
Hydrodynamics
Particle shape
Gas–liquid–solid
High pressure
Surfactant

ABSTRACT

Experiments were carried out to validate the use of spheres in lieu of cylinders when investigating the global hydrodynamic features of a co-current gas–liquid–solid fluidized bed. Two sizes of glass spheres with diameters of 4 and 1.5 mm were compared to aluminum cylinders with equivalent volume/surface area ratios (i.e., matching Sauter mean diameters). Lengths/diameters of the larger and smaller cylinders were 7.5/3.2 mm and 3.1/1.2 mm, respectively, which resulted in equal particle sphericity of 0.8 for both sizes. The particle properties of the larger particles led to the inertial settling flow regime ($Re_{LT\infty} > 500$) in water while the smaller particles were in the intermediate regime ($0.2 < Re_{LT\infty} < 500$). High gas holdup conditions were obtained by increasing the system pressure to 6.5 MPa and/or adding a surfactant. Atmospheric conditions were also studied for comparison. Experiments were conducted in a 101.6 mm diameter column with tap water or a 0.5 wt.% aqueous ethanol solution as the liquid phase while the gas phase was a combination of air and nitrogen. Global phase holdups measured from the dynamic pressure profiles characterized the hydrodynamic behavior of the fluidized bed and studied the impact of particle shape. Standard deviations of the mean holdups aided the comparison and also examined the fluctuations of the bed interface. Liquid–solid fluidized bed experiments demonstrated that equivalent Sauter mean diameters resulted in comparable bed porosities. Gas–liquid–solid fluidized bed dynamics of equivalent size spherical and cylindrical particles were similar in the dispersed bubble flow regime whereas differences were observed in the presence of larger coalescing bubbles.

1. Introduction

Many industrial applications of gas–liquid–solid fluidized beds, e.g. the LC-Finer^{SM} hydroprocessor used for resid upgrading [1], employ extruded cylindrical catalysts. Most gas–liquid–solid fluidized bed experimental studies currently available in the open literature use spherical glass beads due to their ease of use, cost, and availability. Although some studies have used cylindrical extrudates [2–4], the validity of simulating cylindrical particles with spheres in a gas–liquid–solid ebullated bed needs to be investigated.

Flow through a fixed bed of particles can provide a starting point in the literature when accounting for particle shape in a fluidized bed. The Ergun equation [5] is one of the most widely used correlations to determine the pressure drop of a fixed bed.

$$\Delta P/\Delta L = 150\frac{\mu_F U_F}{{d_{SV}}^2}\frac{(1-\varepsilon)^2}{\varepsilon^3} + 1.75\frac{\rho_F {U_F}^2}{d_{SV}}\frac{(1-\varepsilon)}{\varepsilon^3}. \quad (1)$$

Eq. (1) accounts for the shape of non-spherical particles by using the diameter of a sphere with an equal surface area to volume ratio, generally referred as the Sauter mean diameter (d_{SV}). Previous experiments have used the Ergun equation to measure the sphericity (ϕ) of irregular particles in a fixed bed at very low flow rates where viscous forces dominate [6].

Drag on particles must also be considered where a particle's terminal settling velocity, when the force balance is equal to zero, is a key parameter for fluidized beds. The gravitational, buoyant and drag forces acting on a particle at its terminal velocity in a liquid are related as follows:

$$\frac{\pi}{6}(\rho_S - \rho_L) g {d_V}^3 = \frac{1}{2}\rho_L {U_{LT\infty}}^2 C_D A_P \quad (2)$$

where the left hand side is the net gravitational force and the right hand side is the drag force. Examining the previous equation, the drag coefficient (C_D) and projected area (A_P) of the settling particle are required to determine the terminal velocity. Drag coefficients for spherical particles can be estimated via available correlations in the literature [7–10] and the projected area of a sphere can be calculated. These parameters are not as easily determined for cylinders as the projected area and drag coefficient of a cylindrical particle depends upon its orientation. Lau et al.

* Corresponding author. Tel.: +1 613 562 5800x6939; fax: +1 613 562 5172.
E-mail address: arturo.macchi@uottawa.ca (A. Macchi).

0032-5910/$ – see front matter
http://dx.doi.org/10.1016/j.powtec.2013.12.030

[11] observed that the settling of a cylinder in the inertial regime ($Re_{LT\infty} > 500$) resulted in both horizontal and inclined orientations due to wall effects. Some drag coefficient correlations developed for cylinders estimated the projected area based on the diameter of an equal volume sphere while experimentally measuring the terminal velocities [8,12,13]. Although the estimated projected areas may not be accurate, the product of the interrelated drag coefficient and projected area is the parameter required to estimate the terminal velocity. Nonetheless, the orientation of a single cylinder falling in a tube differs from the orientations of many particles in a fluidized bed. The previous correlations used the particle sphericity to account for shape effects. The terminal velocity of cylindrical particles has thus been related using the volume equivalent diameter and particle sphericity.

In liquid–solid fluidized beds, the bed porosity (ε) of spherical particles can be estimated using the Richardson and Zaki [14] empirical correlation.

$$\frac{U_L}{U_{LT\infty}} = k\varepsilon^n. \tag{3}$$

The terminal free settling velocity of the particles ($U_{LT\infty}$), the wall effect factor (k) and the n index can be estimated for spheres using available correlations [10,15]. Gabitto and Tsouris [13] experimentally demonstrated that the Haider and Levenspiel [8] terminal settling velocity predictions for cylinders are relatively accurate for isometric particles with $\phi \geq 0.7$. Wall effects for cylindrical particles have been estimated by Chhabra [16], where non-spherical particles usually experience smaller wall effects compared to spheres, with the exception of cylinders with a length over diameter ratio greater than 10. Unfortunately, no reliable set of correlations have been developed yet to estimate the n index for non-spherical particles [17].

Another method to predict the bed porosity of non-spherical particles assumes that the liquid immobilizes around the surface irregularities, where the particles then behave as smooth spheres [18,19]. This leads to an effective particle volumetric concentration ($K\varepsilon_S$) based on a hydrodynamic volume factor K, defined as the liquid envelope and solid volume divided by the solid volume. Eq. (3) is modified as follows.

$$\frac{U_L}{U_{LT\infty}} = k(1-K\varepsilon_S)^n. \tag{4}$$

The effective volumetric concentration can be estimated by assuming that the settled bed porosity is equivalent to the bed porosity at minimum fluidization [20], which is related to the particle sphericity. The definition of the hydrodynamic volume factor results in effective particle diameters and densities to then estimate the bed porosities using correlations for spheres. The particle properties used to quantify shape and size when estimating bed porosities are again the volume equivalent diameter and sphericity.

The fluid dynamic characteristics of cylindrical particles in gas–liquid–solid fluidized beds have been experimentally studied by some authors. Soung [4] studied the bed expansion of commercial cobalt–molybdenum cylindrical catalysts with n-heptane and nitrogen as the liquid and gas phases, respectively. A correlation was developed that accounted for particle shape via the product of sphericity (ϕ) and the diameter of a sphere with equivalent volume (d_V). Song et al. [3] investigated the hydrodynamic characteristics of seven hydrotreating catalysts consisting of cobalt and molybdenum oxide on extruded porous alumina supports in water and a 0.5 wt.% aqueous t-pentanol solution. The Sauter mean diameter of the particles ranged from 1.51 to 1.90 mm. The authors discussed that particle shape effects were dependent on the bubble/particle size ratio. Bed void fractions for the water fluidized bed were compared to the Begovich–Watson [21] correlation, which underestimated the experimental data. The fit was improved by adding particle sphericity, although its exponent prevents the direct use of the Sauter mean diameter. A separate bed porosity correlation was developed by Song et al. [3] for the surfactant system using d_{SV} to account for particle size and shape. Minimum liquid fluidization velocities (U_{lmf}) and bed porosities of fresh and equilibrium hydrocracking catalysts were studied by Ruiz et al. [2] in water, diesel or jet fuels as the liquid phase and air or nitrogen as the gas phase. Experimental U_{lmf} values were compared to many correlations and the sphericity was successfully incorporated to improve the fit of the two correlations with the best initial predictions (Begovich–Watson [21] and Ermakova et al. [22]). Particle sphericity was again added to the Begovich–Watson [21] correlation for bed porosity to improve the fit for the studied particles.

In summary, the previous gas–liquid–solid studies compared their experimental data obtained using non-spherical particles to correlations developed for spheres. Lack of fit was then corrected by adding the particle sphericity to the existing correlations and fitting the exponent using experimental data. The previous studies however did not directly compare spheres and cylinders in a single gas–liquid–solid fluidized bed to determine a methodology to account for particle shape. As some of the modified correlations did not directly substitute the Sauter mean diameter, it is difficult to conclude whether this parameter effectively accounts for particle shape when comparing the global fluid dynamic behavior of spheres and cylinders. In addition, the gas holdups, an important parameter for ebullated beds, were only measured by Song et al. [3].

Sinha et al. [23] compared the gas–liquid–solid bed porosities of cylindrical and spherical particles using kerosene and heptane as the liquid phases and nitrogen as the gas phase. Although the authors concluded that the spheres and cylinders were equivalent, some experimental observations reveal that the effect of particle shape may not have been fully isolated in the study. The spheres and cylinders used in the study had an apparent size distribution, where the solid phase ordered itself axially based on size when operated as a liquid–solid fluidized bed. The author also mentioned that the pressure profiles along the length of the column were curved, implying that the bed densities were not constant. The previous observations and the exclusion of gas holdup measurements render it difficult to fully compare the fluidized bed behavior of the studied spheres and cylinders.

The objective of this study is thus to experimentally investigate whether the Sauter mean diameter can be used to account for particle shape effects on the global hydrodynamics in a gas–liquid–solid fluidized bed. A comparison of two sets of spheres and cylinders with equivalent Sauter mean diameters was completed in the same experimental system. Particles were selected to minimize particle size and density distribution effects, hence focusing on shape effects. Global gas, liquid and solid holdups in the bed and freeboard regions and fluidization characteristics are compared and discussed over relevant ranges of gas and liquid superficial velocities. Interactions between bubble characteristics and particle shape are studied by increasing the system pressure and/or adding a surfactant. The previous operating conditions also led to high gas holdup conditions which are relevant when studying the fluid dynamics of industrial gas–liquid–solid ebullated beds.

2. Experimental setup

Experiments were carried out in a gas–liquid–solid fluidization system (Fig. 1), purchased from Zeton Inc. (Burlington, Ontario), which is capable of reaching pressures up to 10 MPa. The fluidization column is made of stainless steel with an inner diameter of 101.6 mm and a maximum expanded bed height of 1.8 m. Glass viewing windows with dimensions of 118.75 mm × 15.63 mm are located at heights of 244 mm, 603 mm, and 956 mm above the top of the distributor plate. At the top of the column, an expanded overflow section was designed as the primary gas–liquid separation stage. The liquid is conveyed into a partitioned liquid storage tank for further degassing and then recycled to the column. The system was pressurized using industrial grade nitrogen cylinders. National Instruments hardware and software are used for data acquisition.