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Hydrodynamics of a gas–liquid–solid fluidized bed with hollow cylindrical particles

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ABSTRACT

The hydrodynamic characteristics, viz. the pressure drop, bed expansion and phase holdup profile of a co-current three-phase fluidized bed with an antenna-type air sparger have been determined. Correlations for minimum liquid fluidization velocity, bed voidage and gas holdup have been developed. The bed voidage is found to increase with the increase of both liquid and gas velocities. The gas holdup increases with gas Froude number, but decreases with liquid Reynolds number. The gas holdup is a strong function of the Froude number. The experimental values have been found to agree with the correlations.

1. Introduction

Gas–liquid–solid fluidization also known as three-phase fluidization is a subject of fundamental research since the last three decades due to its industrial importance. Three-phase fluidized beds have been applied successfully to many industrial processes such as in the H-oil process for hydrogenation and hydro-desulfurization of residual oil, the H-coal process for coal liquefaction, Fischer–Tropsch process, and the bio-oxidation process for wastewater treatment. Three-phase fluidized beds are also often used in physical operations [1]. The co-current gas–liquid flow in the three-phase fluidized bed with liquid continuous and gas in dispersed state is quite significant compared to other types [1,2]. The co-current gas–liquid–solid fluidization is defined as an operation in which a bed of solid particles is suspended in upward flowing gas and/or liquid media due to the net gravitational force (i.e. gravitational force – buoyancy force) on the particles. Such an operation generates considerable intimate contact among the gas, liquid and solid particles in the system and provides substantial advantages for applications in physical, chemical or biochemical processing involving gas, liquid and solid phases [3].

The successful design and operation of a gas–liquid–solid fluidized bed system depends on the ability to accurately predict the fundamental characteristics of the system, viz. the hydrodynamics, the mixing of individual phases, and the heat and mass transfer characteristics [4,5]. Knowledge of minimum liquid fluidization velocity is essential for the successful operation of gas–liquid–solid fluidized beds. For such systems the minimum liquid fluidization velocity is the superficial liquid velocity at which the bed becomes fluidized for a given gas superficial velocity [6]. The minimum liquid flow rate required to achieve fluidization is determined by a plot of the bed pressure drop against the superficial liquid velocity at a constant gas flow rate and the corresponding liquid velocity is taken as the minimum liquid fluidization velocity [4]. Visual observation determines the minimum liquid fluidization velocity as either the velocity at which the bed first begins to expand or as the velocity at which any particle within the bed continuously shifts position with neighboring particles [7].

For chemical processes, where mass transfer is the rate-limiting step, it is important to be able to estimate the gas holdup since this relates directly to the rate of mass transfer [8–10]. The following equations have typically been used to determine the volume fraction (holdup) of each phase in a three-phase fluidized bed:

$$\varepsilon_g + \varepsilon_l + \varepsilon_s = 1 \quad (1)$$

$$\frac{\Delta P}{\Delta H} = g(\rho_g \varepsilon_g + \rho_l \varepsilon_l + \rho_s \varepsilon_s) \quad (2)$$

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$$\varepsilon_s = \frac{M_s}{\rho_s A_c H_e} \quad (3)$$

where the bed height in Eqs. (2) and (3) is obtained either visually or from the measured pressure drop gradient [1,11]. A more direct method of measuring gas holdup is to simply isolate a representative portion of the test section by simultaneously shutting two quick closing valves and measuring the fraction of the isolated volume occupied by the gas [2]. Other most promising methods of measuring the local gas holdup are electroresistivity and electro conductivity methods, γ-ray transmission measurements and radioactive tracer techniques [2–4,12,13].

In case of environmental applications of liquid–solid fluidized bed as anaerobic bioreactor and gas–liquid–solid fluidized bed as aerobic bioreactor for wastewater treatment, a higher bed inventory is preferred as it increases the global depollution efficiency of the operation. Biological treatment is a slow process and needs long residence time for the waste water in the bed. Thus in this particular application, the design and operation of the fluidized bed should ensure a good quality of fluidization as well as sufficiently long residence time for the liquid (i.e. low liquid velocity). Sufficient contact time between microorganisms and the pollutants is achieved in the system and maximum solid surface is available to the liquid. This emphasizes the study of the effect of bed mass on liquid–solid and gas–liquid–solid fluidization characteristics [14].

In the above cited literature the solid phase used is spherical particles: like glass beads, steel balls, plastic beads and other spherical catalyst particles; cylindrical particles: like aluminum cylinders and PVC cylinders; other cylindrical catalyst particles and irregular particles like: sand, irregular gravel, quartz particles, etc. having sphericity ranging from 0.7 to 1.0 approximately. Three-phase fluidized beds have been applied successfully in the bio-oxidation process for wastewater treatment in which various low-to-moderate density solid particles of different shape and size are used as cell support. In such reactors high surface area of the particle is desirable, which can be used as solid support for microorganisms, thus resulting in higher mass transfer rate. This can be achieved by the use of hollow cylindrical particles as, these possess very high surface to volume ratio, i.e. of low sphericity.

The present study has been conducted to examine the hydrodynamic behaviour, viz. the pressure drop, minimum liquid fluidization velocity, bed expansion and phase holdup of a co-current gas–liquid–solid three-phase fluidized bed with a modified air sparger using liquid as the continuous phase and gas as the discontinuous phase. Ceramic raschig rings having sphericity of 0.58 have been used as the solid phase as it is of moderate density and high surface to volume ratio due to its hollow cylindrical structure. These have been done in order to develop a good understanding of the gas holdup in low-moderate Reynolds number range. Correlations for minimum fluidization velocity, bed voidage and gas holdup have been developed from experimental data by dimensional analysis approach and compared with the correlation of Song et al. [15] and Safoniuk et al. [8] as they have used cylindrical particle as the solid phase.

2. Experimental setup and techniques

A schematic representation of the experimental setup is shown in Fig. 1. The experimental fluidized bed consists of three sections, viz. the test section, the gas–liquid distributor section, and the gas–liquid disengagement section. The test section is the main component of the fluidizer where fluidization takes place. It is a vertical cylindrical Plexiglas column of 0.1 m internal diameter and 1.88 m long. Any entrained particles are retained on the 16-mesh screen attached to the top of the column. The gas–liquid distributor

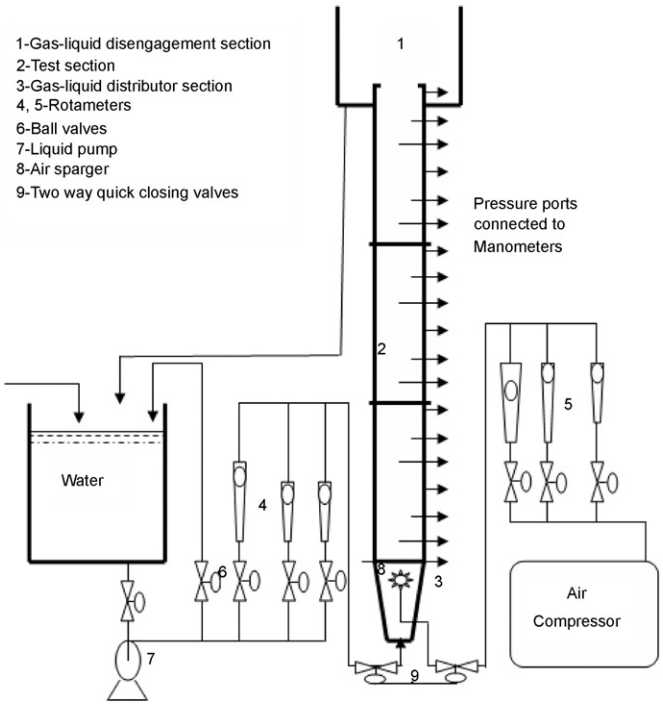

Fig. 1. Schematic representation of the three-phase fluidized bed.

is located at the bottom of the test section and is designed in such a manner that uniformly distributed liquid and gas mixture enters the test section. The distributor section made of Perspex is fructo-conical of 0.31 m in height, and has a divergence angle of 4.5° with one end of 0.0508 m in internal diameter and the other of 0.1 m in internal diameter. The liquid inlet of 0.0254 m in internal diameter is located centrally at the lower cross-sectional end. The higher cross-section end is fitted to the test section, with a perforated plate made of G.I. sheet of 0.001 m thick, 0.12 m diameter having open area equal to 20% of the column area with a 16-mesh (BSS) stainless steel screen in between. This has been done with a view to have less pressure drop at the distributor plate and a uniform flow of the fluids into the test section. There is an antenna-type air sparger of 0.09 m diameter just below the distributor plate containing 50 number of 0.001 m holes, for generating uniform bubbles to flow throughout the cross-section of the column. In this section the gas and liquid streams are merged and passed through the perforated grid. The mixing section and the grid ensured that the gas and liquid are well mixed and evenly distributed into the bed. The gas–liquid disengagement section at the top of the column is a cylindrical section of 0.026 m internal diameter and 0.034 m height, assembled to the test section with 0.08 m of the test section inside it, which allows gas to escape and liquid to be circulated through the outlet of 0.0254 m internal diameter at the bottom of this section.

For pressure drop measurement in the bed, the pressure ports have been fitted to the manometers filled with carbon tetrachloride. Pressure ports are available at seven different levels of equal spacing including one at bottom and one at the top of the test section. This has been done to measure the pressure drops at a particular section at three different radial positions, viz. at the wall, at the center of the column and at one-fourth of the diameter of the column from the wall. With this arrangement, the wall effect, distribution of particle concentration and the gas holdup can be studied clearly.

The three-phase solid, liquid and gas are ceramic raschig rings, tap water and the oil free compressed air, respectively. The scope of the experiment is presented in Table 1. The air–water flow was co-current and upwards. Accurately weighed amount of material