Proceedings of the National Seminar on Recent Advances in Chemical Engineering Operation and Process in Chemical and Allied Industries, Dept. of Chem. Engg., Institute of Technology, Guru Ghasidas University, Bilaspur (C.G.), Feb. 2008

Bed Expansion Behaviour of Cylindrical Particles in a Three-phase Fluidized Bed

H. M. Jena*, G. K. Roy and K. C. Biswal
Department of Chemical Engineering, National Institute of Technology (NIT), Rourkela, Orissa, Pin - 769008, India
hara.jena@gmail.com

Abstract

The bed expansion profile of a co-current three-phase fluidized bed with an antenna type air sparger have been studied using liquid as the continuous phase and gas as the discontinuous phase. Air, water and ceramic raschig rings of equivalent diameter 0.006864 m are used as the gas, liquid and solid phases respectively. The experiments have been carried out in a 0.1 m internal diameter, 1.88 m height vertical Perspex column. The expanded bed height in the fluidized bed regime has been measured visually. Correlations for bed voidage have been developed. The developed correlation for bed voidage can be used for the sizing of such systems. The bed voidage is a strong function of both the liquid velocity and gas velocity. The experimental values have been compared with those predicted by the correlations and have been found to agree well.

Key words: Gas-liquid-solid fluidization; Bed voidage; Hollow cylindrical particles

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 cocurrent 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 [2].

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 [3, 4]. Knowledge of bed expansion (alternatively the bed voidage) is essential for the successful design (sizing of the system) and operation of gas-liquid-solid fluidized beds. For such systems the bed expansion is represented as the bed expansion ratio, which is the ratio of the expanded bed height to the initial static bed height of the solid phase [5, 6].

In case of environmental applications of liquid-solid fluidized bed as anaerobic bioreactor and gas-liquid-solid fluidized bed as aerobic bioreactor for waste water 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 micro-organisms 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 [7].

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 – 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.

In the present study experiments have been conducted to examine the bed expansion behaviour 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 bed expansion behaviour in low-moderate Reynolds number range. Correlation bed voidage have been developed from experimental data by dimensional analysis approach and compared with the correlation of Song et al. [8] as they have used cylindrical particle as the solid phase.

Experimental Set-up and Techniques

A schematic representation of the experimental setup is shown in Fig. 1. The experimental fluidized bed consists of three sections, v.i.z., 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 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 4.5^0 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.

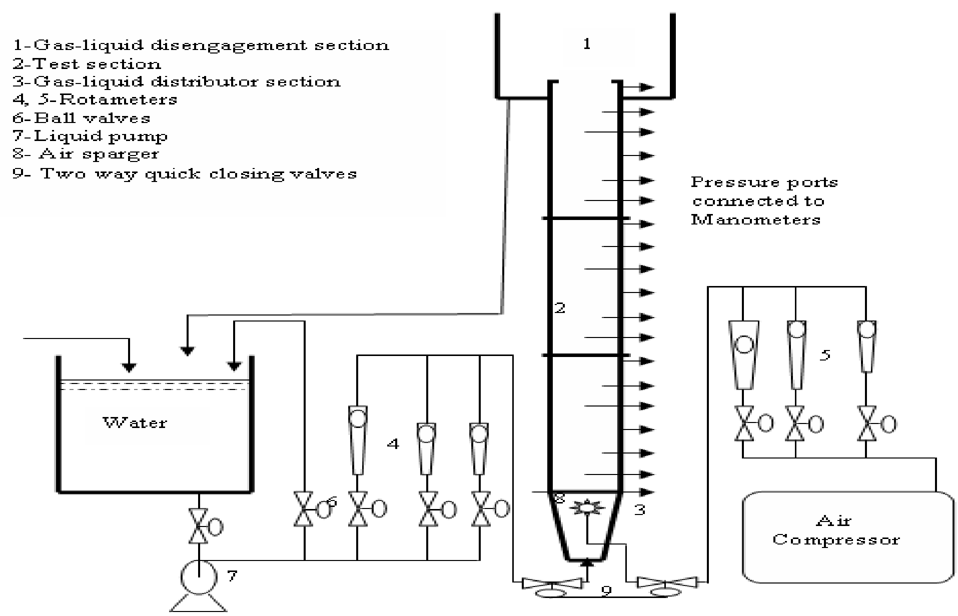

Fig. 1. Schematic representation of the three-phase fluidized bed.

Table 1: Scope of the experiment.

A. Properties of bed materials				
Materials	**L = OD, m**	**ID, m**	**Spherical volume-equivalent diameter, m**	**ρ_p (kg/m^3)**

Ceramic raschig ring	0.0066	0.0033	0.006864	1670
Initial static bed height (m)	0.154	0.214	0.264	0.314
B. Properties of fluidizing medium			**ρ (kg/m^3)**	**μ (Pa.s)**
Air at 25^{0}C			1.187	0.0000181
Water at 25^{0}C			997.15	0.000891
C. Properties of manometric fluid			**ρ (kg/m^3)**	**μ (Pa.s)**
Mercury			13,574	0.001526
Carbon tetra-chloride (CCl_4)			1,600	0.000942

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 was fed into the column and adjusted for a specified initial static bed height. Water was pumped to the fluidizer at a desired flow rate using water rotameter. Then air was injected into the column through the air sparger at a desired flow rate using air rotameter. Three calibrated rotameters with different ranges each for water as well as air have been used to accurately record the flow rates. Approximately five minutes was allowed to make sure that the steady state was reached. Then the readings of the expanded heights of the bed were noted.

Results and discussion

Experiments were conducted with the gas and liquid flow rates which varied from 0 - 0.127389 m/sec and from 0 – 0.14862 m/sec respectively. To ensure steady state in operation at least five minutes were allowed after which the readings for bed expansion were noted down. The experimental results have been presented graphically in this section. Empirical equations have been developed.

The expanded bed height was measured by visual observation and the data were compared with the pressure drop profile along the length of the column measured by a number of manometers. Fairly good agreement was observed between two measurements. The bed expansion study carried out by varying liquid velocity (at a constant gas velocity) has been presented in terms of bed expansion ratio in Fig. 2. It is seen from the figure that the bed expansion ratio increases with increase of both the liquid and the gas velocities.

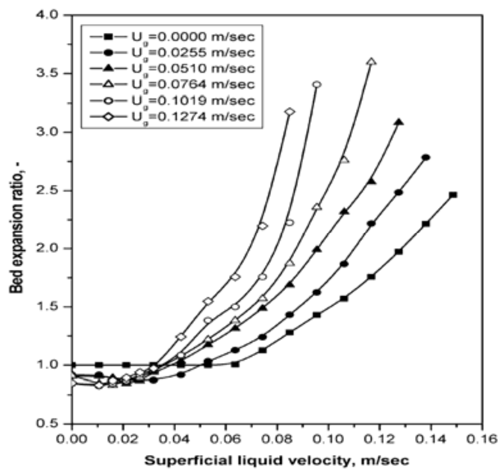

Fig. 2. Variation of bed expansion ratio with liquid velocity for different gas velocities at $H_s = 0.214$ m.

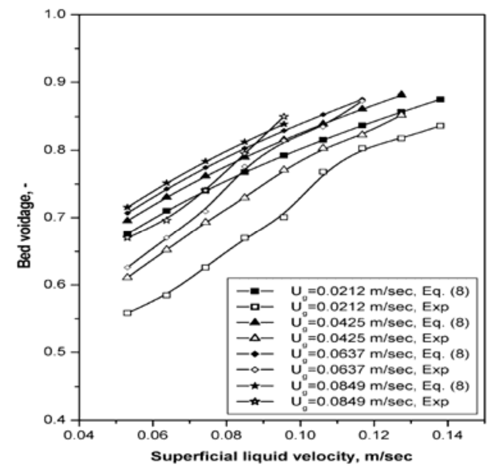

Fig. 3. Variation of bed voidage with liquid velocity for different gas velocities at $H_s = 0.214$ m.

The bed voidage or bed porosity is defined as the fraction of the bed volume occupied by both liquid and gas phases and as such directly proportional to the expanded bed height. As in the present study hollow cylindrical particles have been used as the solid phase, the bed expansion simply does not relate to the bed voidage unless the hollow volume is taken into account. The bed voidage has been calculated by considering the hollow volume and has been represented graphically in Fig. 3 for the conditions above the minimum fluidization. Song et al. [8] have rewritten the original Begovich and Watson [3] correlation by introducing shape factor term for calculation of bed voidage of cylindrical particles in three-phase fluidized bed with air and water as the gas and the liquid phases respectively. The correlation of song et al. [8] is given as under:

$$\varepsilon = 3.93\phi_s^{-0.424} V_l^{0.271} V_g^{0.0410} \mu_l^{0.0550} d_e^{-0.268} D_c^{-0.0330} (\rho_s - \rho_l)^{-0.316} \qquad (1)$$

The bed voidage for the three-phase fluidized bed calculated from the above equation has been presented graphically in Fig, 3. Bed voidage both experimental and calculated from Eq. (1) have been plotted against superficial liquid velocity for different constant gas velocities. It is clear from this that the bed voidage increases with increase of both the liquid velocity and the gas velocity. The bed voidage is a strong function of liquid velocity, but is a weak function of gas velocity. For almost all cases the experimental bed voidage is found to be less than that is predicted from Eq. (1). There is higher deviation between the values at lower liquid velocities but close agreement is seen for higher liquid velocities. This may be due to the fact that the correlation developed by song et al. [8] is from their experimental bed voidage obtained at higher gas and liquid velocities than the present study. Fig. 5 shows the variation of bed voidage with the ratio of superficial liquid to gas velocity. It is clear from the plot that for higher gas velocities (liquid to gas velocity ratio <1.5), there exists very close agreement between the experimental bed voidage and that calculated from Eq. (1).

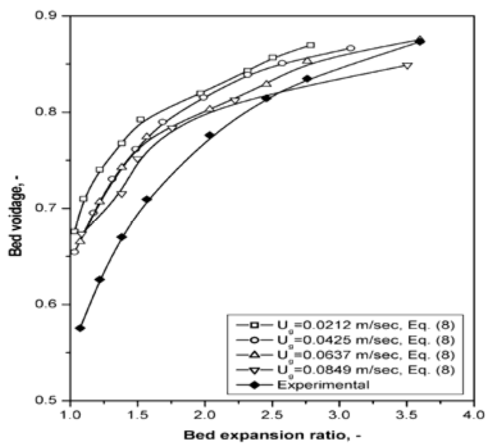

Fig. 4. Variation of bed voidage with bed expansion ratio for different gas velocities at $H_s = 0.214$ m.

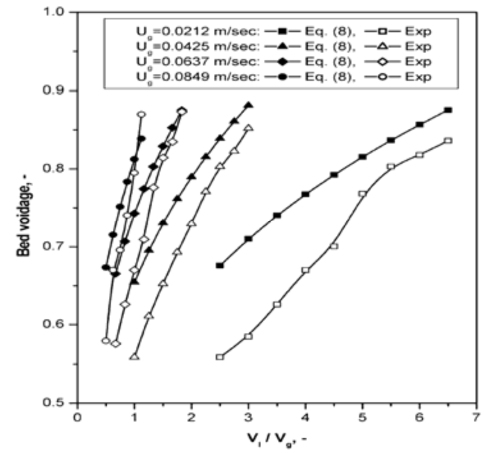

Fig. 5. Variation of bed voidage with ratio of liquid to gas velocity for different gas velocities at $H_s = 0.214$ m.

A simple correlation has been developed from the experimental data of bed voidage above the minimum fluidization condition upto the expanded bed height of about 3.5 times that of the initial static bed height. In this experiment the particle size, sphericity and density of the solid, viscosity and density of liquid, and column diameter are constant. Thus the bed expansion or bed voidage determined here is a simple function of gas and liquid velocities and initial static bed height. Fig. 6 is the plot of the variation of bed expansion ratio with superficial liquid velocity at a constant gas velocity for different initial static bed heights. It is clear from the plot that the bed voidage is not a function of the initial static bed height as the bed expansion ratio is more or less same for all the cases. Thus the bed voidage for this case is a function of gas and liquid velocities only and can be written as:

$$\varepsilon = k_1 V_l^m V_g^n \qquad (2)$$

The results have been fitted to power-law equation for the range of 0.0425 m/sec $\leq V_l \leq$ 0.1380 m/sec and 0.0212 m/sec $\leq V_g \leq$ 0.1062 m/sec, this leads to:

$$\varepsilon = 3.29 V_l^{0.4220} V_g^{0.1469} \qquad (3)$$

(with a standard deviation of 0.02483, mean deviation of 0.01910 and a correlation factor of 0.9701)

The bed voidage values calculated from Eqs. (1) and (3) are compared with the experimental ones in Fig. 7. Fairly good agreement is seen between the experimental values and with those calculated by both the equations for air-water system. More than 65% of the data of Song et al. [8] and all values from present correlation are within 10%. Where as all the values from Eq. (1) is within 20% but with almost positive deviation from experimental ones.

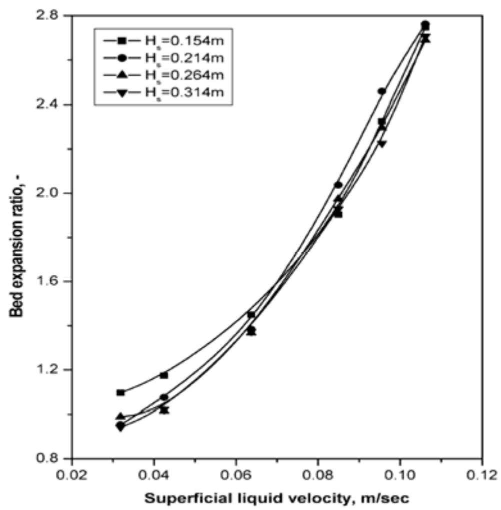

Fig. 6. Variation of bed expansion ratio with liquid velocity for different initial static bed heights at gas velocity of 0.0637 m/sec.

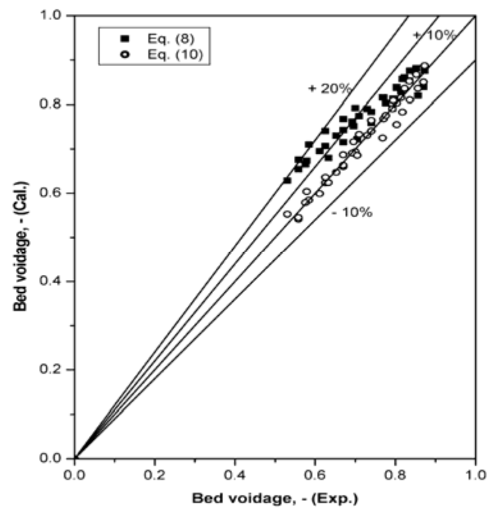

Fig. 7. Comparison of calculated values of bed expansion ratio from Eqs. (1) and (3) with the experimental values.

Conclusions

The new type of hollow cylindrical particles has been used in this study, which can find better applications in process engineering where high surface-volume ratio is required. Especially such particles of smaller size and little higher density than of water can be used as the solid matrix for immobilization cells in fluidized bioreactor for wastewater treatment.

The bed voidage has been found to increase with both gas and liquid velocities in the fluidization regime. The bed voidage at minimum fluidization found has been to be 0.54586 for liquid-solid fluidization, which suddenly decreases with the introduction of the gas. Later with increase in gas velocity the bed voidage increases, but for all cases the bed voidage at minimum fluidization in three-phase has been found to less than the liquid-solid bed.

Correlation for bed voidage developed in the present study is found to be quite significant from the correlation coefficient of 0.9701 and standard deviation of 0.0248 mean deviation of 0.0190 respectively. The developed correlations for bed voidage can be used for the design of similar systems with respect to the sizing of the unit and its accurate operation.

Nomenclature

d_e	equivalent diameter of particle (m)
H	total height of test section (m)
H_e	height of expanded bed (m)
H_s	initial static bed height (m)
R	bed expansion ratio (H_e/H_s)
V_l	liquid velocity (m/sec)
V_g	gas velocity (m/sec)

Greek symbols

ε bed voidage

ρ_g, ρ_l, ρ_s gas, liquid and particle density (kg/m^3)

μ_l liquid viscosity (Pa.s)

Subscripts

l liquid phase

g gas phase

s solid phase

[1] K. Muroyama, L.S. Fan, Fundamentals of gas-liquid-solid fluidization, AIChE. J. 31 (1985) 1-34.

[2] N. Epstein, Three-phase fluidization: Some knowledge gaps, Can. J. Chem. Eng. 59 (1981) 649-757.

[3] J.M. Begovich, J.S. Watson, Hydrodynamic characteristics of three-phase fluidized beds, in: Fluidization, J.F. Davison, D.L. Keairns (Eds.), Cambridge University Press, Cambridge, 1978, pp.190-195.

[4] T.J. Lin, C.H. Tzu, Effects of macroscopic hydrodynamics on heat transfer in a three-phase fluidized bed, Catalysis Today 79–80 (2003) 159–167.

[5] W. Y. Soung, Bed expansion in three-phase fluidization, Ind. Eng. Chem. Process Des. Dev. 17 (1) (1978) 33-36.

[6] H. Yu, B.E. Rittman, Predicting bed expansion and phase hold-up for three-phase fluidized bed reactors with and without biofilm, Water Res. 31 (1997) 2604-2616.

[7] A. Delebarre, J. M. Morales, L. Ramos, Influence of bed mass on its fluidization characteristics, Chem. Eng. J. 98 (2004) 81-88.

[8] G. H. Song, F. Bavarian, L. S. Fan, Hydrodynamics of three-phase fluidized bed containing cylindrical hydrotreating catalysts, Can. J. Chem. Eng. 67 (1989) 265-275.