Analysis of bed voidage characteristic of a gas-liquid-solid fluidized bed by CFD simulation and experiment

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Abstract

The bed expansion profile of a co-current three-phase fluidized bed have been studied using liquid as the continuous phase and gas as the discontinuous phase for different particle size and bed mass. Air, water and glass beads are used as the gas, liquid and solid phases respectively. The experiments were carried out in a 0.1m ID, 1.88m height vertical Plexiglas column. A two dimensional transient model has been developed to simulate the bed voidage behaviour of the mentioned gas–liquid–solid three-phase fluidized bed using the computational fluid dynamics (CFD) method. An Eulerian Granular Multiphase model has been used in the present study and simulations are carried out using the commercial CFD package Fluent 6.2.16. The bed voidage is found to increase with the liquid velocity significantly, but the increase in bed voidage with gas velocity is meagre. The bed voidage is found to decrease with particle size and static bed height. The CFD simulation predictions are compared with the experimental data. The simulation results are in good agreement with the experimental values. Finally empirical correlations have been developed to predict bed voidage behaviour.

Keywords: three-phase fluidization, cfd, bed voidage, multiphase flow.

Introduction

Gas-liquid-solid fluidization also known as three-phase fluidization is a subject of fundamental research since the last four 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. 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 [2]. The successful design and operation of a gas-liquid-solid fluidized bed system depends on the ability to accurately predict the fundamental hydrodynamic characteristics [3-8]. Knowledge of bed expansion (alternatively the bed voidage) is essential for sizing of the system and operation of gas-liquid-solid fluidized beds. 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 [4-10].

The design of three-phase fluidized beds is usually done using correlations available for hydrodynamic characteristics [4,11]. Tarmy and Coulaloglu [12] showed that there was no three-phase hydrodynamic model in the literature and that there was a need for such a model. The complex hydrodynamics of these systems are not well understood due to complicated phenomena. Computational fluid dynamics (CFD) has been promoted as a useful tool for the understanding multiphase reactors for precise design and scale up [13] and has emerged as a new paradigm for modeling multiphase flow and fluidization. The report on the computational models for the hydrodynamic characteristics of three-phase (gas-liquid-solid) fluidized bed is very limited. Hardly there is any literature which is focused on the effect of various variables on the bed voidage (bed expansion) behaviour.

Aim

The prime objective of this work is to investigate numerically the bed voidage behaviour of a three-phase gas-liquid-solid fluidized bed using the above mentioned Eulerian model and then validate the same with the experimental results. The bed voidage directly depends on the bed expansion, in this article the terminology bed expansion has been used very often.

Computational model and experiments

In the present work, an Eulerian granular multiphase model is adopted where gas, liquid and solid phases are all treated as continua interpenetrating and interacting with each other everywhere in the computational domain. The pressure field is assumed to be shared by all the three phases, in proportion to their volume fraction. The motion of each phase is governed by respective mass and momentum conservation equations.

Continuity equation: $\frac{\partial}{\partial t}(\varepsilon_k \rho_k) + \nabla(\varepsilon_k \rho_k \boldsymbol{u}_k) = 0$ (1)

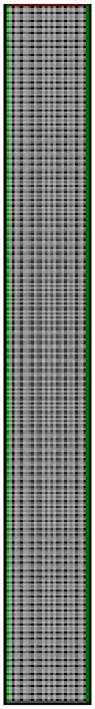
where ρ_k is the density, ε_k is the volume fraction and $\boldsymbol{u}_k$ is the velocity of phase $k = L, g, s$. The volume fraction of the three phases satisfies; $\varepsilon_L + \varepsilon_g + \varepsilon_s = 1$ (2)

Momentum equations: For liquid and gas phases by Eq. (3) and for solid phase by Eq. (4).

$$\frac{\partial}{\partial t}(\rho_k \varepsilon_k u_k) + \nabla.(\rho_k \varepsilon_k u_k u_k) = -\varepsilon_k \nabla p + \nabla.\tau_k + \rho_k \varepsilon_k g + F_{i,k} \quad (3)$$

$$\frac{\partial}{\partial t}(\rho_s \varepsilon_s \boldsymbol{u}_s) + \nabla.(\rho_s \varepsilon_s \boldsymbol{u}_s \boldsymbol{u}_s) = -\varepsilon_s \nabla p - \nabla p_s + \nabla.\boldsymbol{\tau}_s + \rho_s \varepsilon_s g + \boldsymbol{F}_{i,s} \quad (4)$$

where P is the pressure shared by all phases. The second term on the R.H.S of solid phase momentum Eq. (4) is the term that accounts for additional solid pressure due to solid collisions. The terms $\boldsymbol{F}_{i,L}$, $\boldsymbol{F}_{i,g}$, and $\boldsymbol{F}_{i,s}$ of the above momentum equations represent the inter-phase momentum exchange. The terms $\boldsymbol{\tau}_L$, $\boldsymbol{\tau}_g$ and $\boldsymbol{\tau}_s$ are the stress-strain tensors. The inter-phase drag force between the liquid and the solid phases is obtained by Gidaspow drag model. For gas-liquid interaction Schiller and Neumann drag model has been used. The solid phase pressure gradient results from normal stresses resulting from particle–particle interactions accounted based on the kinetic theory of granular flow (KTGF). To describe the effects of turbulent fluctuations of velocity and scalar quantities the simplest but complete: two-equation standard k-ε model has been used [14]. The model equations are solved using Fluent 6.2.16. The fluidized bed simulated is described in Table 1. Fig. 1 depicts the typical numerical mesh used for this simulation. The governing equations are discretized using element based finite volume method and for spatial discretization of the governing equations. For time discretization of the governing equations, a second order backward Euler scheme is used. Two dimensional computational geometry of the fluidization column have been generated by using top-down technique (Face primitive - Rectangle) by using commercial software GAMBIT 2.2.30. After geometry creation, a uniform mesh has been generated with map structured Quadrilateral elements containing height to width ratio of 1. Totally 7520 cells with size of each cell 0.005m x 0.005m have been used for computation.



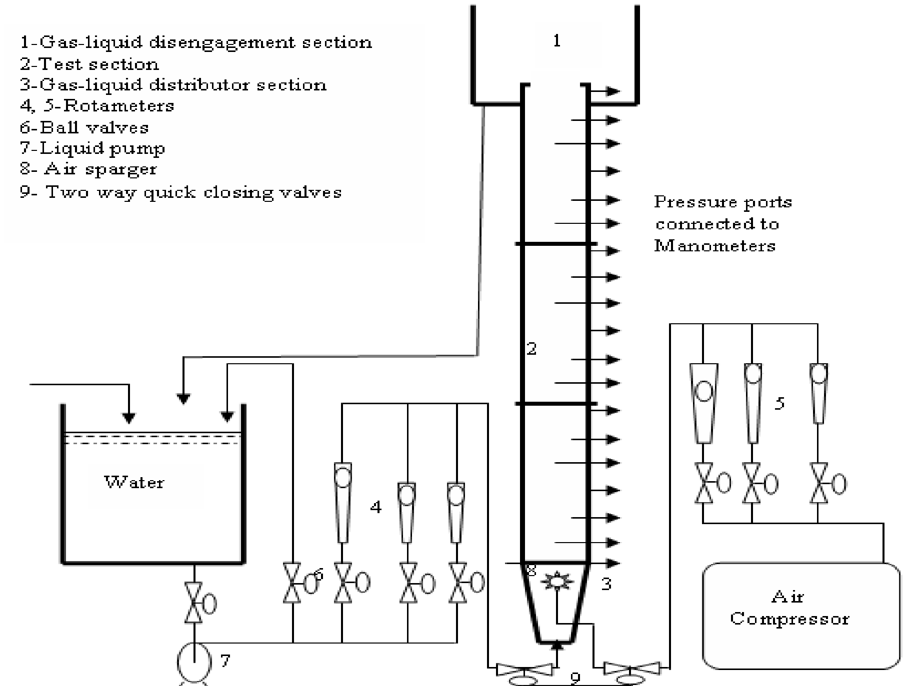


Fig. 1. 2D mesh **Fig. 2. Schematic representation of the experimental three-phase fluidized bed.**

Inlet boundary condition is a uniform liquid and gas velocity at the inlet, and outlet boundary condition is the pressure boundary condition, which is set as1.013×10^5 Pa. Wall boundary conditions are no-slip boundary conditions for the liquid phase and free slip boundary conditions for the solid phase and the gas phase. The volume fraction of the gas at the inlet and in the free board region is based on the inventory. Table 1 shows the boundary and initial conditions. The Phase Coupled SIMPLE method [15] has been chosen for pressure–velocity coupling. The second-order upwind scheme has been used for discretization of momentum, turbulence kinetic energy and turbulence dissipation rate and the first-order upwind scheme has been used for discretization of volume-fraction equations. The time step size of 0.001s has been used. The convergence criteria for all the numerical simulations are based on monitoring the mass flow residual and the value of $1.0e^{-04}$ was set as converged value. The following under relaxation factors have been used for different flow quantities: pressure = 0.3, density = 1, body forces = 1, momentum = 0.2, volume fraction = 0.5, granular temperature = 0.2, turbulent kinetic energy = 0.8, turbulent dissipation rate = 0.8 and turbulent viscosity =1. The simulations have been carried out till the system reached the quasi-steady state i.e., the averaged flow variables are time independent. Fig. 3 shows the variation in the bed profile with time. Simulations continued for 60 s and the averages over the last 20 s were used in the analysis. Once the fully developed quasi-steady state is reached, the averaged quantities in terms of time, axial and radial direction have been calculated.

A schematic representation of the experimental setup is shown in Fig. 2. 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 part where fluidization takes place. 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 gas-liquid disengagement section at the top of the column is a cylindrical section of larger diameter which allows gas to escape and liquid to be circulated. The detailed design of the experimental setup descried elsewhere [5-9]. The experimental conditions are reported in Table.1.

Table 1: Description of system used in simulation

Diameter of column:	0.1 m	Liquid phase (water), 30^0C	
Height of column:	1.88 m	Viscosity, Pas:	7.98×10^{-4}
Solid phase (glass beads):		Density, Kg/m^3:	995.7
Particle size, mm:	2.18	Gas phase (air), 30^0C	
Particle density, Kg/m^3:	2470	Viscosity, Pas:	1.794×10^{-5}
Initial static bed height, m:	0.171, 0.213	Density, Kg/m^3:	1.166
Bed inventory, kg:	1.965, 2.450	Superficial liquid velocity:	0.004246 to 0.1746 m/s
Static bed voidage:	0.41	Superficial gas velocity:	0 to 0.1019 m/s

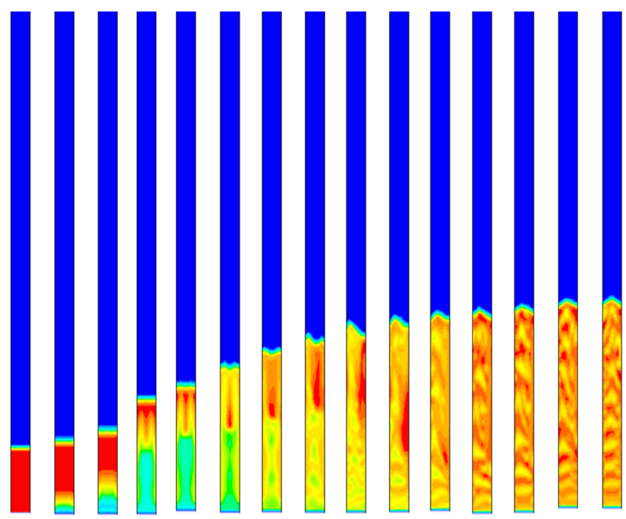


Fig. 3. Contours of volume fraction of 2.18 mm glass beads at water velocity of 0.12 m/s and air velocity of 0.0125 m/s for initial bed height 0.213 m [at time 0,0.5, 1, 2, 2.5, 4, 6, 8, 10, 12, 14, 16, 18, 20, 30 seconds from left to right].

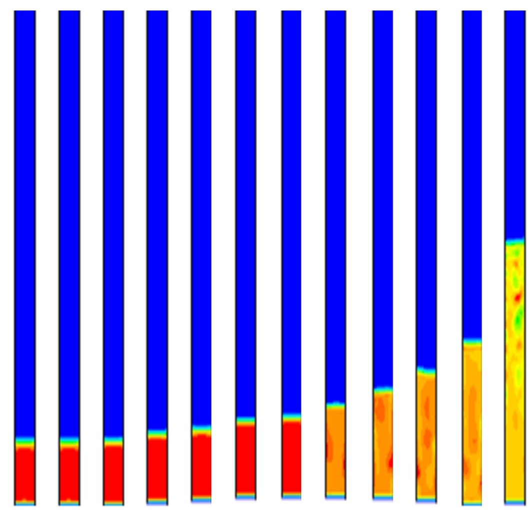


Fig. 4. Contour plot of solid volume fraction at liquid velocities: 0.035, 0.04, 0.045, 0.05, 0.055, 0.06,0.071, 0.082, 0.093, 0.104, 0.11 and 0.12 m/s for 2.18 mm glass beads at air velocity of 0.0125 m/s for initial bed height 0.213 m.

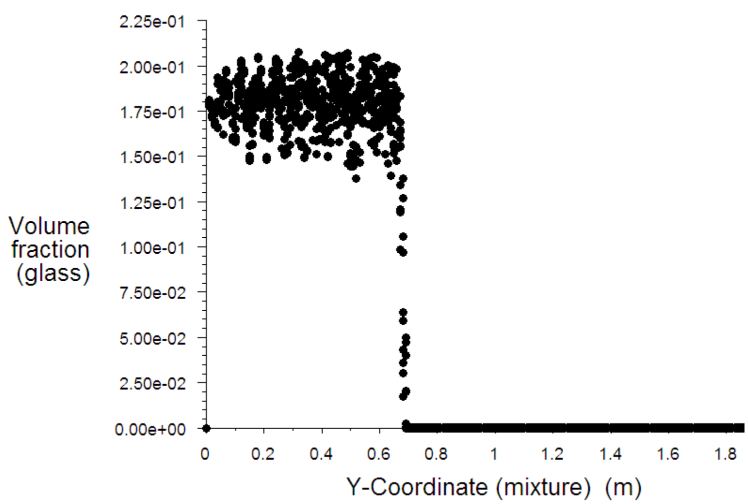


Fig. 5. XY Plot of solid volume fraction.

Results

The bed expansion dynamics has been represented in the form of contours and XY plots. Figs. 3 and 4 represent the contours of volume fractions of solid. The contours for glass beads illustrates that bed is in fluidized condition. The bed height has been determined from the XY plot of the solid volume fraction w.r.t. the axial distance from the base of the column (in 2D mesh it is noted as y-coordinate) [Fig. 5]. The point where the solid fraction sharply decreases to zero value can be taken as the height of the bed. Fig. 6 shows the plot of expanded bed height vs. liquid velocity obtained at different values of inlet gas velocity. Fig. 7 shows a comparison of the simulated and the experimental values of expanded bed height. A very good agreement is seen between the values for the gas velocity of 0.05 m/s, while for gas velocity of 0.10 m/s, the values have been found to deviate.

Conclusions and Discussion

In gas-liquid-solid system with increase in liquid velocity at a constant gas velocity, the expanded bed height increases and the voidage of the bed also increases. It can be seen from the contours of solid volume fraction (as shown in Fig. 4) that there is steady increase in bed height with liquid velocity above the minimum fluidization condition.

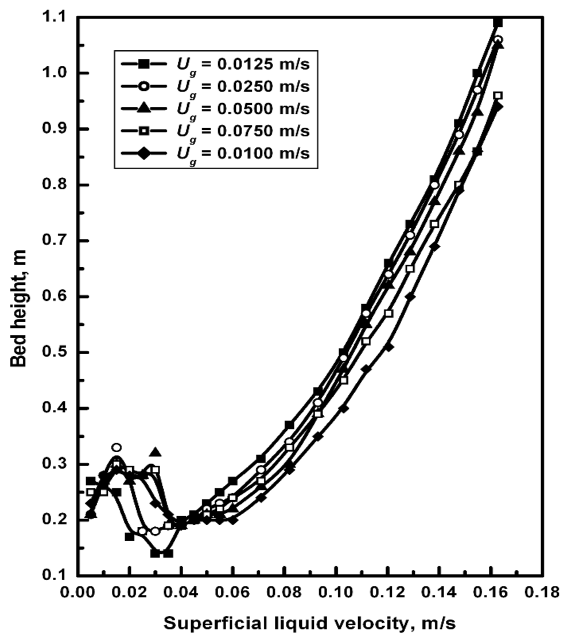


Fig. 6. CFD simulation result of bed expansion behaviour of 2.18 mm glass beads at H_s=0.171 m.

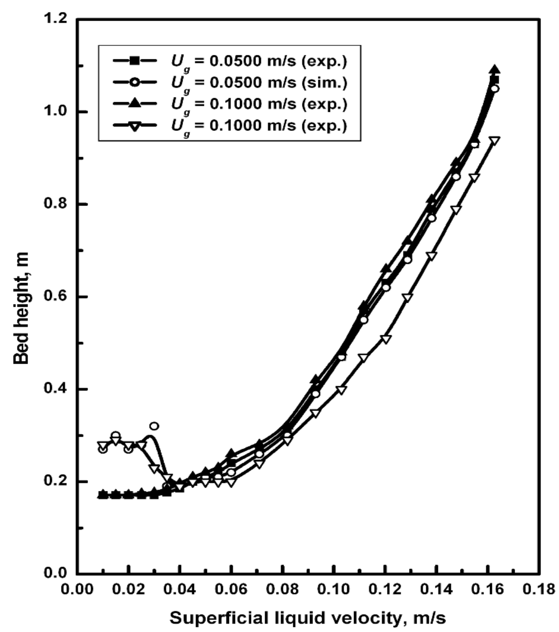


Fig. 7. Comparison of bed height obtained from experiment and CFD simulation.

At this higher gas velocity the simulated bed height has been found to be less than the experimental one. In experiment, a slight increase in bed height has been observed with the increase in the gas velocity. The computational model used is based on the prescription by some investigators; they have found a decrease in bed height with the increase in gas velocity. The bed contraction truly occurs for particles of sizes close to 1 mm or less than that. Above all in the bed expansion regime (condition above minimum fluidization), the expanded bed height value from experiment and from simulation agrees within 10 %. In Figs. 6 and 7 at low liquid velocity, the bed height has been found to be higher, then it decreases and further increases with increase in liquid velocity. This indicates the presence of an agitation in the bed by the larger size gas bubbles where few particles are lifted giving the pseudo feeling of fluidization. Larger size bubbles appear at low liquid velocities and as liquid velocity increases the bubble size decreases.

In experiment the agitated bed height has been neglected while measuring the bed height. CFD simulation result shows the agitated bed. In CFD result the agitated bed vanished near the minimum fluidization state. The good agreement between the values obtained from CFD simulation and experimental ones for the range of the present operating variables justify that the Eulerian-Eulerian multi-phase granular flow approach is capable to predict the overall performance of gas–liquid–solid fluidized bed.

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