Effect of Distributor Shape on Gas-Solid Bubbling Fluidized Bed Reactor

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***Abstract*— This work presents a computational study of flow behaviour in a bubbling fluidized bed polymerization reactor. The model is developed by using the commercial CFD code Fluent 6.3. The model is based on an Eulerian description of the gas and the particle phase. The effect of distributor shape on fluidization is analyzed using the in-built Gidaspow and Syamlal-O'brien Drag Model. The computational results are validated against classical calculated data. Both these drag model parameters are found to affect the bubble behavior and hydrodynamics of different shaped distributors. The dimension of the lab-scale fluidized bed is 0.33 × 0.33 × 0.9 m. The simulations are performed with spherical particles with mean particle size of 0.001m and density 910 kg/m^3 while that of gas is 21.56 kg/m^3. The superficial gas velocity is 0.3 m/s.**

***Keywords*- Multiphase flow; Fluidization; Computation; Modeling; CFD; Two-dimensional**

NOMENCLATURE

C_D	[-]	Friction coefficient
d_s	[m]	Particle diameter
e	[-]	Coefficient of restitution
g_i	[m/s^2]	Acceleration due to gravity
g_0		Radial distribution function
K_{qm}	[kg/m^3·s]	Coefficient for the interface force between the fluid phase and the solid phase
p	[Pa]	Fluid pressure
p_s	[Pa]	Solid phase pressure
Re_s	[-]	Particle Reynolds number
U_{qi}	[m/s]	Velocity vector for phase q
v_r	[m/s]	Terminal velocity

Special characters

α_q	[-]	Volume fraction of phase q
δ_{ij}	[-]	Kroenecker delta
ρ_q	[kg/m^3]	Density of phase q
$\bar{\tau}_{ij}$	[kg/m·s^2]	Stress tensor
μ	[kg/m·s]	Viscosity
ξ_s	[kg/m·s]	Bulk viscosity
Θ_s	[m^2/s^2]	Granular temperature

Subscripts

i, j, k	i, j and k directions
g	Gas phase
s	Solid phase

I. INTRODUCTION

Fluidised Bed Reactors are found in many operations in the chemical, petroleum, pharmaceutical, agricultural, biochemical, food, electronic, and power-generation industries. In a fluidized bed gas is passing upwards through a bed of particles supported on a distributor. Fluidized beds are applied in industry due to their large contact area between phases, which enhances chemical reactions, heat transfer and mass transfer. The efficiency of fluidized beds is highly dependent of flow behaviour and knowledge about flow behaviour is essentially for scaling, design and optimization; however, the precise analysis of the flow field has not been achieved yet because of these complex phenomena between the gas and particles. In order to improve the performance of these processes of actual systems, detailed knowledge about the distribution of the different solid species throughout the bed in different operations is required. The mechanism for bubble formation, bubble growth and particle segregations are one of the most difficult research subjects in the gas bubbling fluidized beds Computational fluid dynamics, CFD is an emerging technique for predicting the flow behavior of these systems, as it is necessary for scale-up, design, or optimization [1].

In the last decade considerable progress has been made in the area of hydrodynamic modelling of the fluidized beds. Work is required to make CFD suitable for fluidized bed reactor modeling and scale-up. Broadly speaking, two different categories of models can be distinguished: Eulerian models and Lagrangian models. Lagrangian models solve the Newtonian equations of motion for each individual particle, taking into account the effect of particle collisions and forces acting on the particle by gas. Particle collisions are described by collision laws, that account of energy dissipation due to non-ideal particle interactions by means of the empirical coefficient of restitution and friction (hard sphere approach) or an empirical spring stiffness and a friction coefficient (soft sphere approach).The distinct element method, DEM, is one of the trajectory models, which can calculate the particle velocity and the corresponding particle trajectory to examine interactions, such as those due to multibody collisions. Though models based on a DEM allow the effects of various particle properties on the motion of fluid to be studied, it is computationally intensive. Due to computational limitations, the Eulerian–Lagrangian model is normally limited to a relatively small numbers of particles. The number of particles that these models can handle (typically less than 10^6) is orders of magnitude lower than that encountered in most (industrial) fluidised beds. Therefore, continuum models constitute a more natural choice for hydrodynamic modelling of engineering scale systems, whereas discrete particle models can be applied as a valuable research tool to verify and further develop closure laws for these continuum models. Therefore, the multi-fluid model is the preferred choice for simulating macroscopic hydrodynamics.

Eulerian models considered all phases to be continuous and fully interpenetrating. The equations employed are a generalization of the Navier Stokes equations. In this scheme, collections of particles are modeled using continuous medium mechanics. The solid particles are generally considered to be identical having a representative diameter and density. The general idea in formulating the multi-fluid model is to treat each phase as an interpenetrating continuum, and therefore to construct integral balances of continuity, momentum and energy for both phases, with appropriate boundary conditions and jump conditions for phase interfaces. Since the resultant continuum approximation for the solid phase has no equation of state and lacks variables such as viscosity and normal stress, certain averaging techniques and assumptions are required to obtain a momentum balance for the solids phase.

Averaging theorems are applied to construct a continuum for each phase in order for the Eulerian description of single-phase flows to be extended to the multiphase flow. Although the transport coefficients of the gas phase may be reasonably represented by those for a single-phase flow with certain modifications, the transport coefficients of the solid phases must account for gas–particle interactions and particle–particle collisions.

The interphase momentum transfer between gas and solid phases is one of the dominant forces in the gas- and solid phase momentum balances. This momentum exchange is represented by a drag force. The application of different drag models significantly impacted the flow of the solid phase by influencing the predicted bed expansion and the solid concentration in the dense phase regions of the bed. Researchers have shown that their models are sensitive to drag coefficient. In general, the performance of most current models depends on the accuracy of the drag formulation. A number of different drag models have been proposed in modelling of fluidized beds. Ergun developed a drag model that was derived empirically for Newtonian flow through packed beds in a narrow band of porosities around 0.4. In an active fluidized bed the void fraction can vary over the whole range from zero to unity and the models used in numerical simulations should be equally versatile. Gidaspow combined the Ergun equation with the equations of Rowe and Wen and Yu to get a drag model that can cover the whole range of void fractions. Syamlal and O Brian have also developed an empirical drag model that that can cover the whole range of void fractions.

The application of kinetic theory to model the motion of a dense collection of nearly elastic spherical particles is based on an analogy to the kinetic theory of dense gases. A granular temperature is defined to represent the specific kinetic energy of the velocity fluctuations or the translational fluctuation energy resulting from the particle velocity fluctuations. In granular flow, particle velocity fluctuations about the mean are assumed to result in collisions between particles being swept along together by the mean flow. The granular particle temperature equation can be expressed in terms of production of fluctuations by shear, dissipation by kinetic and collisional heat flow, dissipation due to inelastic collisions, production due to fluid turbulence or due to collisions with molecules, and dissipation due to interaction with the fluid(Gidaspow, 1994). Numerous studies have shown the capability of the kinetic theory approach for modeling bubbling fluidized beds

The success of numerical computation of bubbling fluidized beds critically depends upon the ability to handle dense packing of solids. At high solid volume fraction, sustained contacts between particles occur and the resulting frictional stresses might be accounted for in the description of the solid phase stress. Granular flows can be classified into two flow regimes, a viscous regime and a plastic regime. In a viscous or rapidly shearing regime, the stresses arise because of collisional or translational transfer of momentum, whereas in a plastic or slowly shearing regime, the stresses arise because of Coulomb friction between grains in enduring contact [11].

In the present study the Eulerian approach is used to investigate gas-solid flow in a two dimensional fluidized bed. The in-built drag model of Syamlal & O Brien in Fluent 6.3, and the simulations in the present study are based on these two drag models. The frictional stresses are not included in the simulations.

II. PHYSICAL DESCRIPTION OF BED DYNAMICS

Computational studies have been performed on a 2-dimensional fluidized bed. Spherical particles with a mean diameter of 0.001m and a density of 910 kg/m^3 are used. The behaviour of particles in fluidized beds depends on a combination of their mean particle size and density. Geldart fluidization diagram [12], shown in Figure 1, is used to identify characteristics associated with fluidization of powders. The current particles are classified as Geldart B particles, but are very close to Geldart A particles. The fluidization properties for these two groups of particles differ significantly from each other.

Particles characterized in group A are easily fluidized and the bed expands considerably before bubbles appear. This is due to inter-particle forces that are present in group A powders [13]. Inter-particle forces are due to particle wetness, electrostatic charges and van der Waals forces. Bubble formation will occur when the gas velocity exceeds the minimum bubble velocity and the bubbles rise faster than the gas percolating through the emulsion. For group B particles the inter-particle forces are negligible and bubbles are formed as the gas velocity reaches the minimum fluidization velocity. The bubble size increases with distance above the gas distributor and increases also with increasing excess gas. The bed expansion is small compared to group A particles.

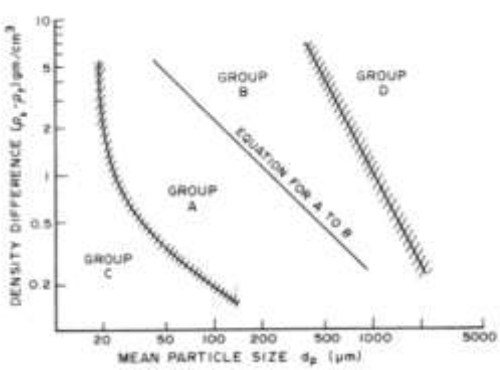


Figure 1 Geldart classification of particles according to their fluidization behaviour [12]

III. NUMERICAL METHOD

The computational work is performed by using the commercial CFD code Fluent 6.3. The model is based on an Eulerian description of the gas and the particle phase. The default settings in Fluent 6.3 are used to describe the granular phase [14]. The energy equation is not solved, and it is assumed that there is no mass transfer between the phases.

The continuity equation for phase q can then be expressed as:

$$\frac{\partial}{\partial t}(\alpha_{ij}\,\rho_q) + \frac{\partial}{\partial x_i}(\alpha_q \rho_q U_{i,q}) = 0 \qquad (1)$$

The momentum equation in the j direction for phase q is:

$$\frac{\partial}{\partial t}(\alpha_q \rho_q U_{j,q}) + \frac{\partial}{\partial x_i}(\alpha_q \rho_q U_{i,q}\, U_{j,q})$$

$$= \alpha_q \frac{\partial P}{\partial x_j} + \frac{\partial \tau_{ij,q}}{\partial x_j} + \alpha_q \rho_q g_j + \sum_{m=1,\, m\neq q} K_{qm}\,(U_{j,m} - U_{j,q}) \qquad (2)$$

where the terms on the lower line represent the pressure forces, viscous forces, mass forces and drag forces respectively. The gas phase stress tensor is expressed by:

$$\tau_{ij,g} = \mu_g \left[\left(\frac{\partial U_{gj}}{\partial x_i} + \frac{\partial U_{gi}}{\partial x_j} \right) - \frac{2}{3}\delta_{ij} \left(\frac{\partial U_{gk}}{\partial x_k} \right) \right] \qquad (3)$$

and the solid phase stress tensor is:

$$\tau_{y,s} = -\,p_s \delta_{ij} + \mu_s \left(\frac{\partial U_j}{\partial x_j} + \frac{\partial U_{si}}{\partial x_i} \right)_s + \left(\xi_s - \frac{2}{3}\mu_s \right) \delta_{ij} \left(\frac{\partial U_k}{\partial x_k} \right) \qquad (4)$$

In the simulations the bulk viscosity is set to zero, and the solid viscosity is set constant. The solid phase pressure is modelled based on the kinetic theory of granular flow and is expressed by the following equation [14]:

$$p_s = \alpha_s\, \rho_s\, \Theta_s\, [\,1 + 2(1+e)\, g_0\, \alpha_s\,] \qquad (5)$$

where the terms on the right hand side represent the kinetic and the collisional contribution to the solid pressure respectively.

The radial distribution function expresses the probability of collisions between the particles. The function will approach unity for dilute regions and infinity in the dense regions of the bed. The radial distribution function is given by [15]:

$$g_0 = \left[1 - \left(\frac{\alpha_s}{\alpha_{s,\max}} \right)^{1/3} \right]^{-1} \qquad (6)$$

In a bubbling fluidized bed the concentration of particles varies from very low to very high. In dilute regions, the kinetic of the particles will dominate the solids viscosity, and the solid pressure will be close to zero. In regions with higher concentration of particles, the collisions between particles will dominate the solids viscosity, and the solid pressure will increase. At very high concentration of particles, the frictional stresses dominate the solid viscosity. In this study the frictional stresses are not accounted for.

A. *Drag models*

The drag describes the momentum exchange between phases and is expressed by the drag coefficient K_{qm} in the momentum equation. In this work two different drag models are used, The Gidaspow drag model and the Syamlal & O Brien drag model. The Gidaspow drag model is a combination of the Ergun equation and the drag model of Wen and Yu. The Ergun equation is developed for packed beds and is only valid at high particle concentrations. To get a model that covers the whole range of particle concentrations, the Wen and Yu equation is used for the lower concentrations. The Gidaspow model for gas particle drag is:

$$K_{sg} = 150\,\frac{\alpha_s\,(1-\alpha_g)\,\mu_g}{\alpha_g\,d_s^2} + 1.75\,\frac{\rho_g\alpha_g\,|U_g - U_s|}{d_s} \qquad (7)$$

This is the Ergun equation and is valid for $\alpha_g \le 0.8$. The Wen and Yu equation is valid for $\alpha_g > 0.8$, and is expressed by:

$$K_{sg} = C_D\,\frac{3\alpha_s\alpha_g\rho_g\,|U_g - U_s|}{4d_s}\,\alpha_g^{-2.65} \qquad (8)$$

The friction coefficient is developed by Rowe, and is related to the Reynolds number:

$$C_D = \frac{24}{\mathrm{Re}_s}\,(1+0.15\,\mathrm{Re}^{0.687}),\ \mathrm{Re} \le 1000 \qquad (9)$$

$$C_D = 0.44, \qquad \mathrm{Re} > 1000$$

The Syamlal & O Brien drag model is:

$$K_{sg} = C_D\,\frac{3\alpha_s\alpha_g\rho_g\,|U_g - U_s|}{4v_r^2\,d_s} \qquad (10)$$

The formula for the terminal velocity is developed by Garside and Al Dibuouni [14] and is an analytical formula:

$$v_r = 0.5\left(A - 0.06\,\mathrm{Re}_s + \sqrt{(0.06\,\mathrm{Re}_s)^2 + 0.12\,\mathrm{Re}_s(2B - A) + A^2}\right) \qquad (11)$$

The constants *A* and *B* are:

$$A = \alpha_g^{4.14} \qquad (12)$$

$$B = 0.8\,\alpha_g^{1.28}, \qquad \alpha_g \le 0.85$$

$$B = \alpha_g^{2.65}, \qquad \alpha_g > 0.85$$

The drag factor is proposed by Dalla Valle [14] and is expressed by:

$$C_D = \left(0.63 + \frac{4.8}{\sqrt{\mathrm{Re}_s/v_r}}\right)^2 \qquad (13)$$

The granular temperature is a measurement for the random movement of the particles and influences on the solid pressure. In Fluent 6.3 there are two options for calculation of the granular temperature. The granular temperature can be described with a separate conservation equation or with an algebraic expression [14]. The algebraic expression is used in this work.

The governing equations are solved by a finite volume method, where the calculation domain is divided into a finite number of non-overlapping control volumes. The simulations are performed using three-dimensional Cartesian co-ordinates. The conservation equations are integrated in space and time. This integration is performed using first order upwind differencing in space and is fully implicit in time.

IV. Computational Set-Up

A computational study of bubble behaviour in a 2-D fluidized bed is performed. The cross section area of the bed is 0.33 m × 0.33 m and the height is 0.9 m. The initial bed height is 0.75 m, and the initial void fraction in the packed bed is 0.63. A two dimensional Cartesian co-ordinate system is used to describe the fluidized bed. The grid resolution is 5 mm in horizontal and vertical direction and the total number of control volumes is 11330. Spherical particles with a diameter of 0.001 m and density 910 kg/m^3 are used. The coefficient of restitution is set to 0.9. The boundary conditions are given as velocity inlet and pressure outlet. The inlet superficial gas velocity is set to 0.3 m/s and the outlet pressure is 1.4 atm. The simulations are run for about 3 s real time, and the computational results are compared to experimental data obtained on a corresponding fluidized bed with the same set-up and flow conditions. The calculated minimum fluidization velocity for particles with diameter of 0.001 µm and density 910 kg/m^3 is 0.12 m/s [6], and according to Geldart fluidization diagram the particles are characterized as D particles.

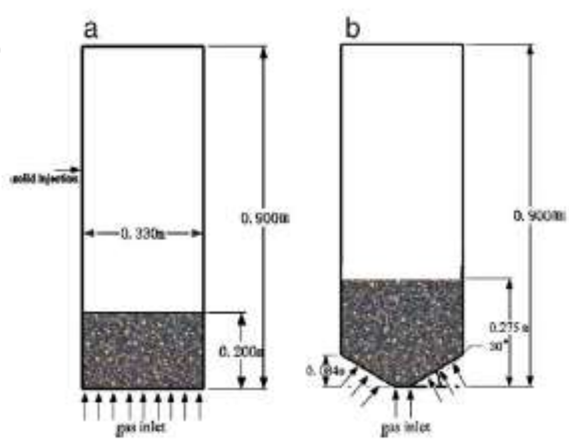
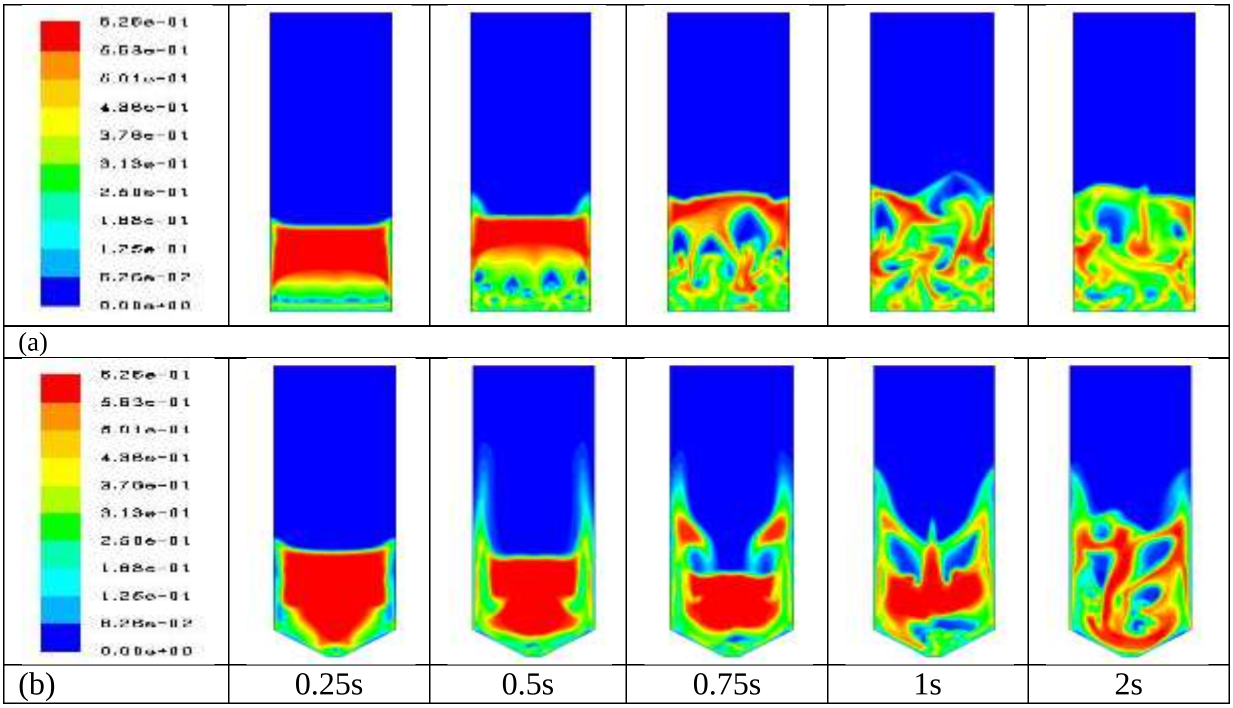


Figure 2: Geometry Configuration: (a) Plane Distributor (b) Triangular Distributor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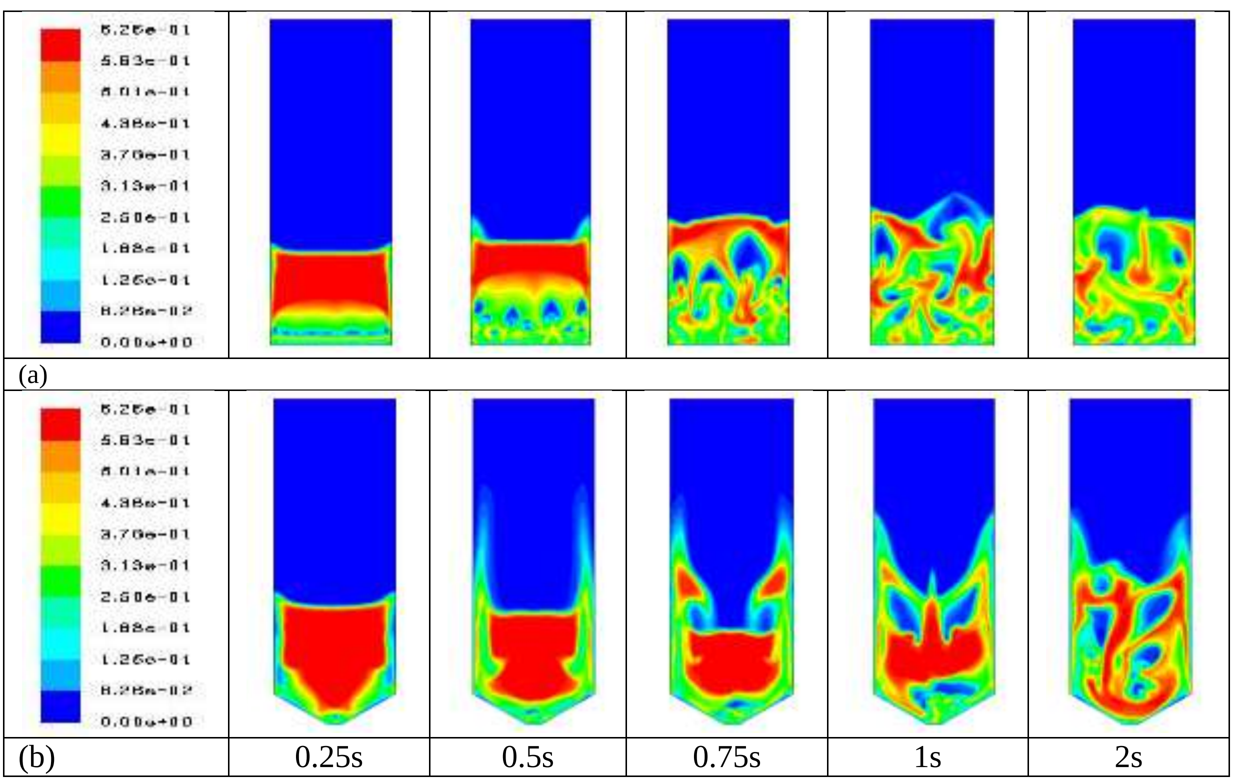
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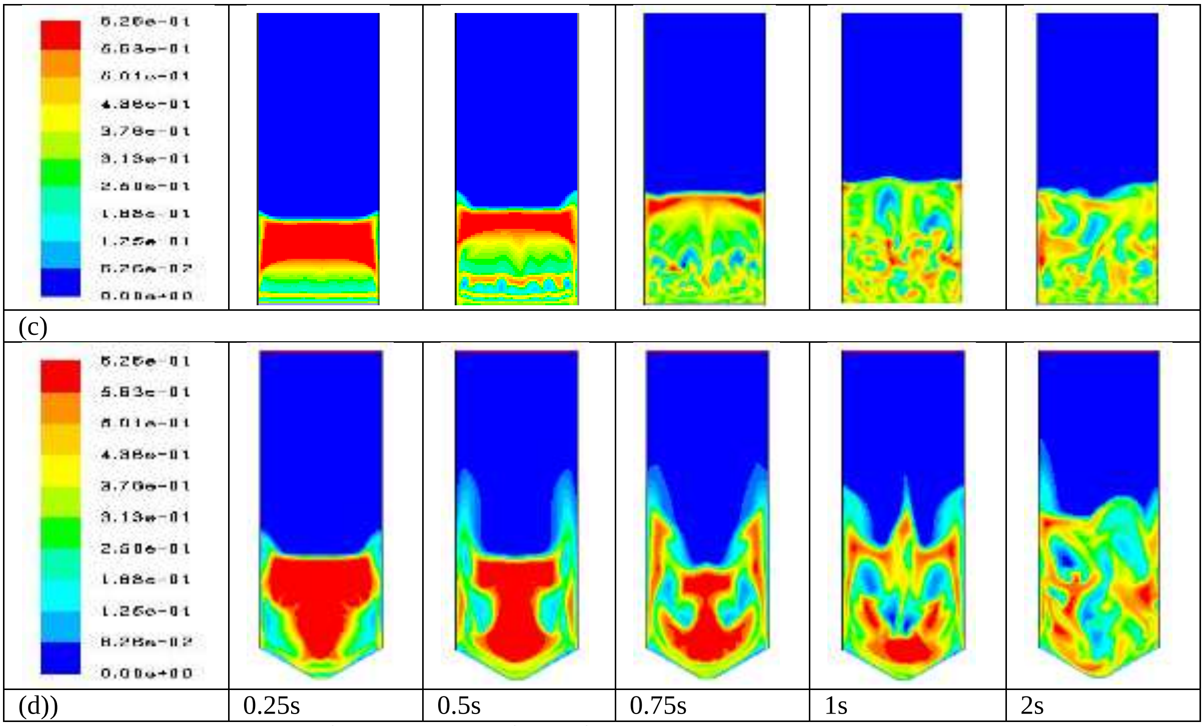

Figure3: Visual Representation of solid phase volume fraction for Gidaspow and Syamlal Drag Model in plane and triangular distributor (a) Plane Distributor for Gidaspow Drag Model (b)Triangular Distributor for Gidaspow Drag Model(c) Plane Distributor for Syamlal Drag Model (d)Triangular Distributor for Syamlal Drag Model

V. RESULTS

The quality of bubbling fluidization is strongly influenced by the type of gas distributor used. As one of the key components of the gas–solid FBRs, the distributor makes gas-distribution uniform and leads to an excellent quality in the FBRs. Furthermore, selecting the proper distributor shape is crucial for designing the distributor. At present, many open papers reported the plane distributor .In the industry, however, the triangle distributor is employed widely. Here, the bubble behaviors at the two type distributor shapes are obtained as shown in Figure.3..Figure 3 shows the contour plots of solid phase volume fraction in plane and triangular distributor for Gidapsow and Syamlal Drag Model.

For the two distributor shapes, the initial solid packing height is adjusted to make sure that they have the same amount of solid at the beginning. As shown in Fig.3, a traditional plane shape distributor and the triangle shape distributor are compared in this section. Fig.3 illustrates the difference between the two shapes, and the final bubble behaviors described via particle fraction are almost the same as shown in Fig.3. Compared to the plane shaped distributor, the action areas of particles increases due to the flexural direction of gas flow vector at the triangle shape. Accordingly, the solid particles are more easily to be fluidized due to the quicker change of bubbles toward the uniform particle distribution in the FBR with the triangle shape distributor than that with the plane distributor. Indeed, from Fig.3 one can find that the fluidized time to the uniform particle distribution at the triangle shape is less than that at the plane shape. Therefore, in practice, to apply the triangle shape distributor is in order to change the gas flow vector (gas flow rate and direction), which makes the particles fluidized more easily.

Also one may observe from the figure that incase of plane distributor, the formation of bubbles is more prominent in Gidaspow Drag Model rather than that in Syamlal Drag Model. The bubbles are easily penetrated through as they move along into the bed for Gidaspow Drag Model. For triangular shaped distributor similar arguments holds true. However in both the cases, the top of the bed is very unstable for Gidaspow Drag Model and shows chaotic bubble formation.

VI. CONCLUSION

The CFD code Fluent 6.3 is used to study flow behaviour in a 2-dimensional bubbling fluidized bed with different shaped distributor. The Eulerian approach is used to describe the gas and the solid phase. The simulations are performed with Gidaspow drag model and the drag model developed by Syamlal & O Brian. The results from the simulations with these two drag models differ significantly from each other. Both the models give high bed expansion, and shows chaotic bubble formation. However the bubble formation is more predominant in Gidaspow drag model rather than Syamlal Drag model. Another important result to note is that in order to simplify the geometry, plane distributors are employed in the computational domain of fluidized bed reactors. However in general, the fluidized bed reactors are triangular in shape especially those employed in industries. The triangular shaped distributors show the ease of fluidization as compared to plane distributors and thus needs to be considered during CFD simulation of FBR's.

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