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Hydrodynamics of Bubbling Fluidized Bed by Computational Fluid Dynamics

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Abstract

Gas–solid fluidized bed reactors have many industrial applications and have been studied by many researchers. In this study, the 2D hydrodynamics of fluidized bed was investigated by using CFD analysis. To perform the simulation of fluidized bed, the two-fluid model (TFM) by the kinetic theory of granular flow (KTGF) was used to describe the solid particles as continuum phase. Also for this system (dense gas-solid fluidized bed), an algebraic granular energy-balance equation is proposed to determine the granular temperature instead of solving the full granular energy balance equation. This simplification does not lead to different results, but significantly reduces the computational effort of the simulation. The computational fluid dynamics simulation results were compared to bubbling fluidized bed containing the average spherical glass beads of 275 micron in diameter. The results of this simulation for local voidage and local solid velocity are in a good agreement with the experimental data and show a suitable trend in comparison with the theoretical findings. Whilst the superficial gas velocities were 0.1, 0.38 and 0.48 m/s.

Key words: multiphase, fluidized bed, hydrodynamics, CFD, KTGF,

Introduction

Fluidization is an effective method for mixing the solid particles in gas or liquid phase for many industrial applications such as drying, freezing and chemical reaction processes. Gas-solid fluidized bed reactors have many applications in chemical, oil, medical, agricultural, biochemical, electronic and power industries. These reactors have been widely used because of their suitable mixing characteristics and high surface contact between phases [1]. For the purpose of increasing efficiency, modeling and hydrodynamic behavior of these systems are essential. Understanding of fluidized bed hydrodynamics is essential for selecting suitable operation parameters in fluidization regime. CFD addition of construct cost and saving the time, have several advantages in simulation of fluid bed such as optimum design and scale up of these systems. Today CFD has shown that it can be powerful instrument for modeling of multiphase flows [2]. Some of the difficult and challenge related to CFD validation of fluidized bed were reviewed by Grace and Taghipour [3]. For obtaining basic sight of suspension dynamic of fluidized bed behavior, elementary education of material properties and physical properties such as drag forces, friction forces, energy dissipations and solid fluctuations are needed. With this sight many researchers can drive several models that able to good prediction of gas – solid flow in fluidized bed.

Basic approaches to model gas–solid flows are Eulerain- Eulerian and Eulerian- Lagrangian. For gas-solid modeling, usually, Eulerian-Lagrangian and Eulerain-Eulerian models are called discrete particle and granular flow models, respectively. Both approaches were compared in

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literatures [4, 5, 6]. The two phases have interpenetrating continuum behavior in the first approach. Eulerian-Lagrangian models are limited to number of particles that Kuipers et al. studied in this problem [7].
Eulerian - Eulerian is the most commonly used approach for fluidized bed simulations. Generally multi fluid model is a suitable select for macroscopic hydrodynamics simulations [8]. CFD modeling based on Eulerian approach is a feasible approach for performing parameter investigations and scale-up and design studies [9].
There are two types of governing equations for multi phase flow. They are Anderson and Jackson (1967) and Ishii (1975) equations [10, 11]. These equations were derived from first principles, but the inherent assumption in these two type of governing equations constrain the types of multi phase flows to which they can be applied. Ishii's equations are appropriate for a dispersed phase containing of fluid droplet and Anderson and Jackson's equations are appropriate for a dispersed phase containing of solid particles [12].
Generally to describe solid shear stress, solid pressure and solid viscosity for identifying solid phase as a fluid (TFM) in gas-solid dense flow, the kinetic theory of granular flow (KTGF), is the best method. KTGF has been developed based on particle velocity fluctuations [13, 14].

$$q = \frac{1}{3} < n_S'^2 > \quad (1)$$

Numerous studies have shown the capability of the kinetic theory approach for modeling the bubbling fluidized beds [8, 9, 15, 16, and 17].
The KTGF has been modified to be applicable to fine and semi-cohesive particles such as FCC powders, from Geldart's group A and C. The modification accounted for the cluster formation and/or agglomeration of the particles [18].
Recognition of major forces in fluidized bed, play a key rule in successful simulation of fluidized bed. In the gas-solid fluidized bed, gravity and drag are the governing forces in the flow while in the dense flow, frictional pressure play an important role [12]. Drag force is used for describing the effect of interactions between gas and particle phases. Numerous correlations for calculating the momentum exchange coefficient of gas-solid systems have been reported in the literature, for example Syamlal and O'Brien (1989), Gidaspow (1994), and Wen and Yu (1966) drag models are amongst the famous models that are used in simulation of fluidized bed [19, 20, 21].
In this research, two-fluid model was used for saving calculation time and performing the suitable simulation of the bubbling fluidized bed. Algebraic granular energy-balance equation was solved instead of full granular energy balance equation for determination of the granular temperature this will reduce the computational effort without loosing the accuracy. The model applies the best constitutive equations.

CFD Model

The simulation of fluidized bed was performed using Eulerian-Granular model of CFD code, containing the set of momentum and continuity equations for each phase, which are linked through pressure and interphase exchange coefficients. The solid properties are obtained by using the kinetic theory of granular flow (KTGF). The governing equations of the system contain the conservation of momentum, mass and energy. The continuity equation for *q*th phase without any mass transfer between the phases is written as:

$$\frac{\partial}{\partial t}(\alpha_q \rho_q) + \nabla \cdot (\alpha_q \rho_q \vec{v}_q) = 0 \quad (2)$$

Where ρ_q and $\vec{v}_q$ are the density and velocity of *q*th phase, respectively. The conservation of momentum for gas phase (g) is written as:

$$\frac{\partial}{\partial t}(\alpha_g \rho_g \vec{v}_g) + \nabla \cdot (\alpha_g \rho_g \vec{v}_g \cdot \vec{v}_g) = -\alpha_g \nabla P + \nabla \overline{\overline{\tau}}_g + \alpha_g \rho_g g - K_{gs}(\vec{v}_g - \vec{v}_s) \quad (3)$$

The conservation of momentum for solid phase (s) also is written as:

$$\frac{\partial}{\partial t}(\alpha_s \rho_s \vec{v}_s) + \nabla \cdot (\alpha_s \rho_s \vec{v}_s \cdot \vec{v}_s) = -\alpha_s \nabla P - \nabla P_s + \nabla \overline{\overline{\tau}}_s + \alpha_s \rho_s g - K_{gs}(\vec{v}_g - \vec{v}_s) \quad (4)$$

Where $a_s = 1 - a_g$.

The balance of granular energy $(\frac{3}{2}\Theta)$ related to velocity particle fluctuations for completing the momentum and continuity equations in two phases is needed. Therefore the conservation of the kinetic energy of the moving particles is described as following granular temperature Θ_s that is derived from the KTGF:

$$\frac{3}{2}\left[\frac{\partial}{\partial t}(\alpha_s \rho_s \Theta_s) + \nabla \cdot (\alpha_s \rho_s \vec{v}_s \Theta_s)\right] = \left(-P_s \overline{\overline{I}} + \overline{\overline{\tau}}_s\right) : \nabla \vec{v}_s + \nabla \cdot (k_{\Theta s} \nabla \Theta_s) - \gamma_{\Theta s} + f_{gs} \quad (5)$$

Description of the relations

Structural correlations for solid pressure and viscosity as well as drag coefficient and stress tensor are important in momentum conservation equations. One of the best methods for identifying these parameters in Eulerian-Eulerian approach is kinetic theory of granular flow that was used in this work. In Eq. 5 the first term on the right-hand side represents the production of fluctuating energy due to shear in the particle phase, the second term represents the diffusion of fluctuating energy along gradient in Θ_s , $\gamma_{\theta s}$ represent the collisional dissipation of energy while the last term (f_{gs}) is the net rate of transfer of the fluctuation energy between the phases.

Some researchers have assumed that the granular energy is in steady state and this parameter is dissipated locally thus the convection and diffusive terms are negligible [22, 23, 9]. With this assumption Eq. 5 will be expressed as:

$$0 = \left(-\nabla P_s \overline{\overline{I}} + \overline{\overline{\tau}}_s\right) : \nabla v_s - \gamma_s \quad (6)$$

This simplification is only valid under the particular assumptions of high volume fraction and low velocity of solid phase. As regards to bed conditions, granular temperature can be calculated by solving equations (5) or (6). To obtain granular temperature for the dense gas-solid fluidized bed, equation (6) was used instead of solving full granular temperature transport equation.

Except the drag model another constitutive relations were needed to close any governing relations. Equation (7) represents the constitutive relations that were used in the current model.

The Syamlal-O'Brien drag model [19] is calculated as:

$$K_{gs} = \frac{3}{4} \frac{\alpha_s \alpha_g \rho_g}{{v_{r,s}}^2 d_s} C_D \left(\frac{Re_s}{v_{r,s}}\right) \left|\vec{v}_s - \vec{v}_g\right| \quad (7)$$

Where the drag coefficient is written as:

$$C_D = \left(0.63 + \frac{4.8}{\sqrt{Re_s / v_{r,s}}}\right)^2 \quad (8)$$

The terminal velocity also is written as:

$$v_{r,s} = 0.5[A - 0.06 Re_s + \sqrt{(0.06 Re_s)^2 + 0.12 Re_s (2B - A) + A^2}] \quad (9)$$

Where

$$A = \alpha_g^{4.14}, B = 0.8\alpha_g^{1.28} \text{ for } \alpha_g \le 0.85$$
$$A = \alpha_g^{4.14}, B = \alpha_g^{2.65} \text{ for } \alpha_g > 0.85 \quad (10)$$

It has been reported that behavior of fluidization in bubbling fluidized bed contain Geldart B, is less sensitive to the chosen drag models [24,25].
Solid phase stress tensor is expressed as:

$$\overset{\approx}{\tau}_s = \alpha_s \mu_s \left(\nabla \vec{v}_s + \nabla \vec{v}_s^{\,T}\right) + \alpha_s \left(\lambda_s - \frac{2}{3}\mu_s\right)\nabla.\vec{v}_s \overset{=}{I} \quad (11)$$

The Radial distribution function which represents the probability of particle collisions when the granular phase becomes dense is calculated as:

$$g_{0,ss} = [1 - (\frac{\alpha_s}{\alpha_{s,max}})^{1/3}]^{-1} \quad (12)$$

The other required relations are:
Collision dissipation energy

$$\gamma_{\Theta m} = \frac{12(1 - e_{ss}^2) g_{0,ss}}{d_s \sqrt{\pi}} \rho_s \alpha_s^2 \Theta_s^{3/2} \quad (13)$$

Transfer of kinetic energy

$$\varphi_{gs} = -3 K_{gs} \Theta_s \quad (14)$$

Solid pressure

$$p_s = \alpha_s \cdot \rho_s \cdot \Theta_s + 2 \cdot \rho_s \cdot (1 + e_{ss}) \cdot \alpha_s^2 \cdot g_{0,ss} \cdot \Theta_s \quad (15)$$

Solid shear viscosity

$$\mu_s = \mu_{s,col} + \mu_{s,kin} + \mu_{s,fr} \quad (16)$$

Solid collision viscosity

$$\mu_{s,col} = \frac{4}{5} \alpha_s \rho_s d_s g_{0,ss} (1 + e_{ss}) \left(\frac{\Theta_s}{\pi}\right)^{1/2} \quad (17)$$

Kinetic viscosity (Syamlal–O'Brien) [19]

$$\mu_{s,kin} = \frac{\alpha_s d_s \rho_s \sqrt{\Theta_s \pi}}{6(3 - e_{ss})} \left[1 + \frac{2}{5}(1 + e_{ss})(3e_{ss} - 1)\alpha_s g_{0,ss}\right] \quad (18)$$

Frictional viscosity, in dense flow at low shear, where the volume fraction for a solid phase is closes to the packing limit, will be used. In this study the generation of stress is mainly due to friction between particles, thus the frictional viscosity is calculated as [26]:

$$\mu_{s,fr} = \frac{p_s \cdot \sin\varphi}{2.\sqrt{I_{2D}}} \quad (19)$$

The solid bulk viscosity accounts for the resistance of the solid phase for compression and expansion of the bed is expressed as [26]:

$$\lambda_s = \frac{4}{3}\alpha_s \rho_s d_s g_{0,ss}(1+e_{ss})\left(\frac{\Theta_s}{\pi}\right)^2 \tag{20}$$

In this study the value of restitution coefficient, 0.9, was accepted [24, 25].

Geometry and operating condition of the bed

In this work the experimental data that obtained by Taghipour et al[24] was used. The height of bed is 1 m, the width of bed is 0.28 m, and the bed thickness is negligible compared to other dimensions of the bed. The average diameter of solid particle (Geldart B) is 275 micron with density of 2500 kg/m^3. The density of the air input was used is 1.225 kg/m^3, Also the range of superficial gas velocity that were used are 0.1 to 0.46 m/s. The 2D simulation domain is discretizing by 11200 rectangular cells and the governing equations in this system were solved by finite volume approach. The time step 0.001with 20 iteration per each time step was chosen. The phase-coupled SIMPLE (PC-SIMPLE) algorithm was used for the pressure–velocity coupling.

Initial and Boundary conditions

All velocities for the two phases and the granular temperature in the bed were set to zero m/s and 0.0001 m/s, respectively. The upper section of the simulated geometry, or freeboard, was considered to be occupied by gas, only. The initial static height of solid in the bed is 0.4 m and the solid volume fraction at this height is 0.6.

The condition of no slip of the two phases for the inside wall of the bed was assumed. Dirichlet boundary conditions were employed at the bottom of the bed to specify a uniform gas inlet velocity. Out put of the bed was defined as outflow, in this region all flow quantities were given zero normal gradient, except pressure gradient.

Results and discussions

The unsteady state CFD simulations using Eulerian-Granular approach performed with the gas superficial velocities of 0.1, 0.38 and 0.46 m/s .

Fig. 1 shows the contour plot of solid volume fraction. Initially, the height of solid in the bed increased until it leveled off at the steady state height of the bed, although the apparent fluctuation is due to bubble growth and coalescence. As the fluidization proceeds the bubbles have split and coalesced continuously resembling boiling of water.

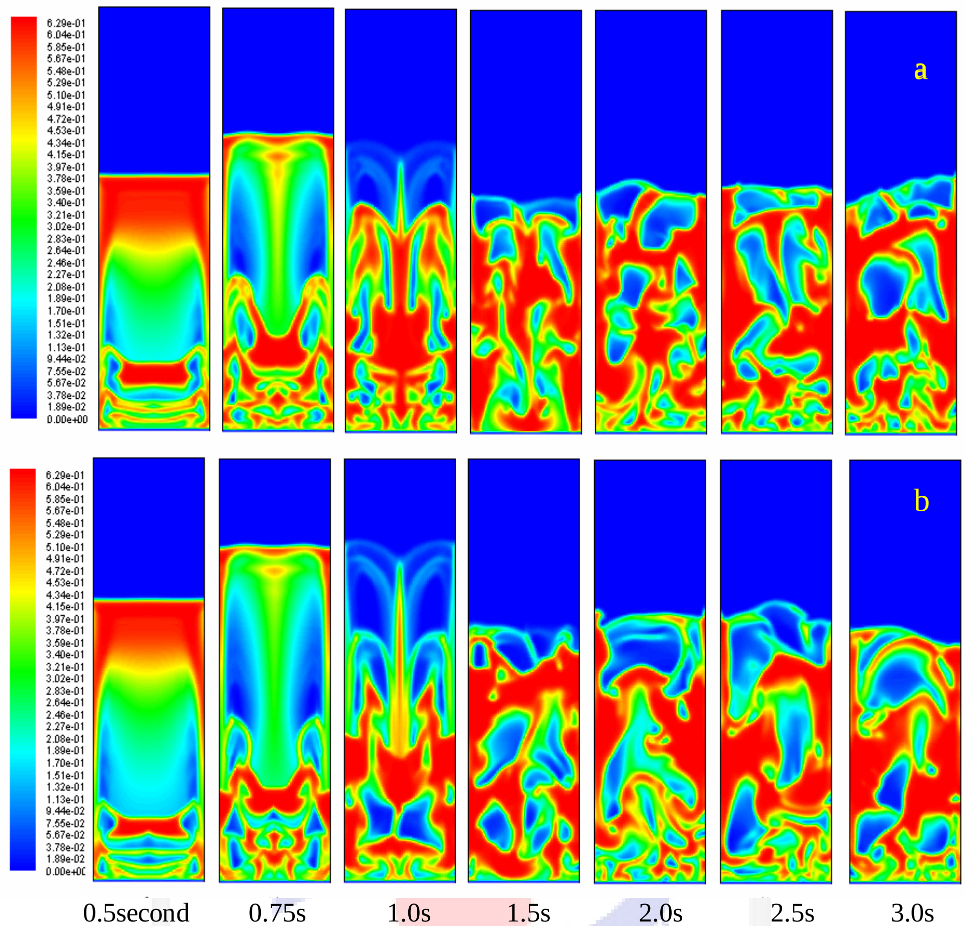


Fig. 1. Contour plot of solid volume fraction at different time (a) U=0.38 m/s, (b) U=0.46 m/s

The bed expansion ratio and contour of bubble distribution against gas velocity are shown by Figs. 2 and 3, respectively. In this analysis the time-average bed expansion ratio for the purpose of validating simulation and experimental results were taken after about 3seconds of simulation. The model does not predict fluidization of the bed at U = 0.1 m/s as shown in Fig. 2. This is due to the adapted constant value of drag coefficient that should be used in the simulation model on basis of the minimum fluidization velocity. However, under estimation of the model predictions and the experimental data is about 10%.

It has been reported that over-prediction of the bed expansion would have been occurred for Geldart group A by using original form of Syamlal-O'Brien drag model, but good agreement with experimental results by using modified Syamlal-O'Brien drag model. A great challenge in CFD modeling of gas-solid two-phase flows is to obtain realistic predictions of the fluidization behavior of small particles such as Geldart A particles [27].

The experimental observations indicated small bubbles near the distributor and large bubbles at the top of the bed; the bubbles grow as they rise to the top of the bed with coalesces. The reason of this phenomenon is the wall effects and interaction between bubbles. This issue is illustrated in Fig. 3.

Fig. 4 shows the pressure drop versus time at two gas inlet velocities. The total pressure drop has an oscillation in bubbling fluidized bed, because of bubble slip and coalescence. This issue is represented in Fig. 4. The higher gas velocity causes the higher bubble size. This increases bubble splitting and coalescence. Therefore a fluctuation in pressure drop will be seen.

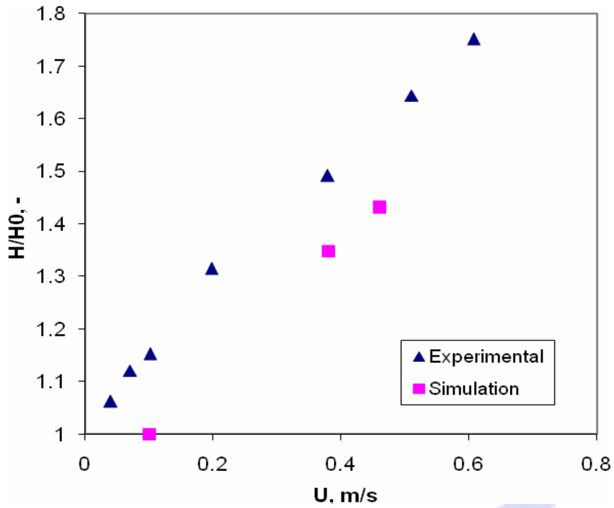
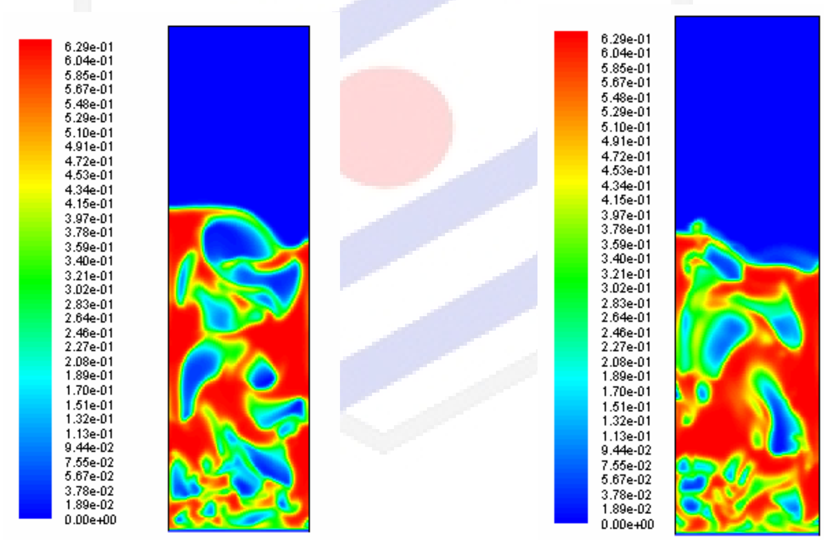


Fig. 2. Bed expantion ratio at different velocities



(b) U = 0.46 m/s
(a) U = 0.38 m/s

Fig. 3. Contour plot of solid volume fraction at different velocities (before 20 second)

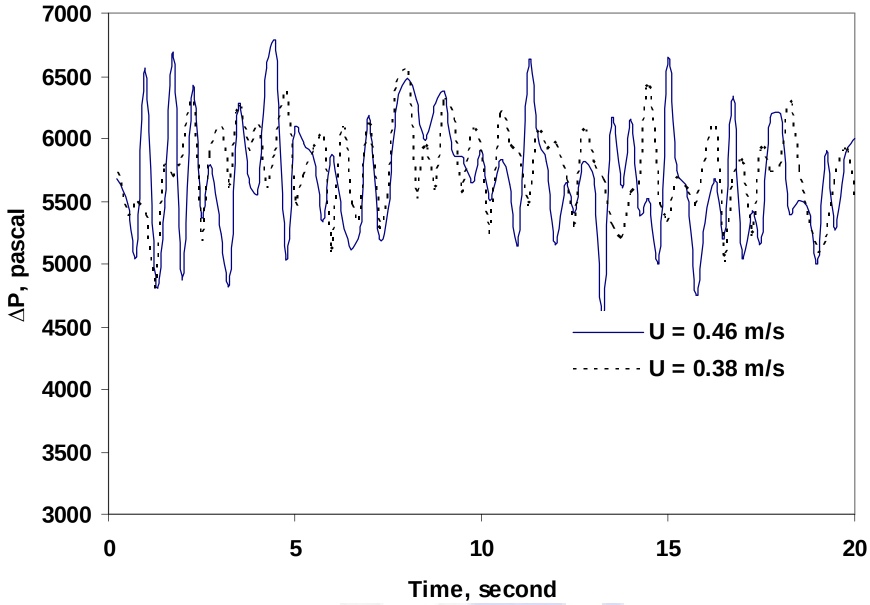


Fig. 4. Pressure drop in bed versus time, at two different velocities

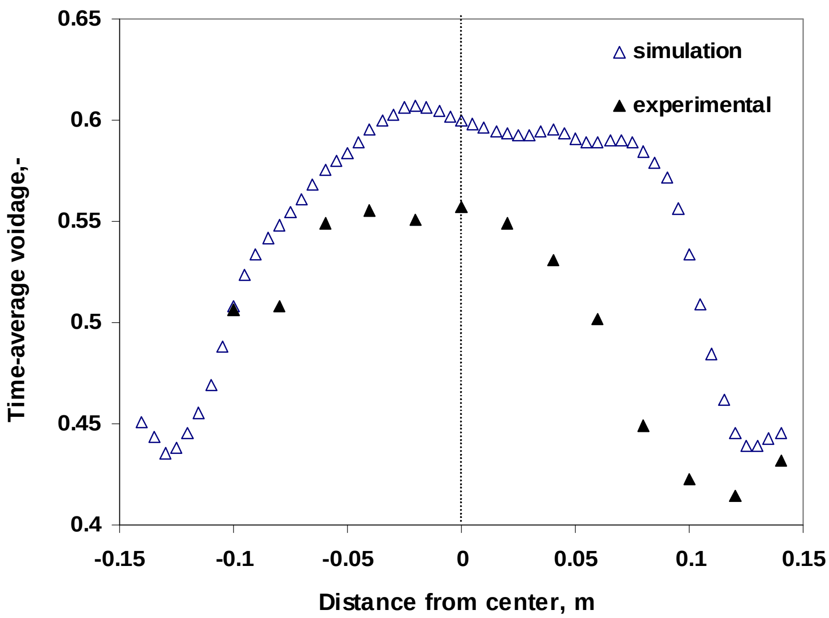


Fig. 5. Experimental and simulated time-average local voidage at z = 0.2m above the distributor

Time- averaged voidage in the bed for real time simulation of 5 to 24 seconds is compared with experimental data at superficial gas velocity of 0.46 cm/s in Fig. 5. For both experimental and simulation results, voidage is increased symmetrically and voidage profiles are flatter near the center of the bed,. Further from distributor, the more developed flow causes a flatter voidage and velocity profile. Such behavior confirms this fact that velocity of both phases increase in central region of the bed in comparison with the velocity close to the wall region. As an important result any discrepancy could be due to the effect of the gas distributor, which was not considered in the

CFD modeling of fluid bed. At higher simulation time the fluctuation of simulation results diminishes.

Fig. 6 shows the results of CFD simulation for time-average local voidage at different velocities. It shows that the time-average local voidage has increased with the increase of superficial gas velocity; as Zimmermann and Taghipour [27] results for Geldart A. Fig. 6 clears that at higher velocity (0.46 m/s) the time-average voidage behavior is flatter and symmetric than at lower velocity.

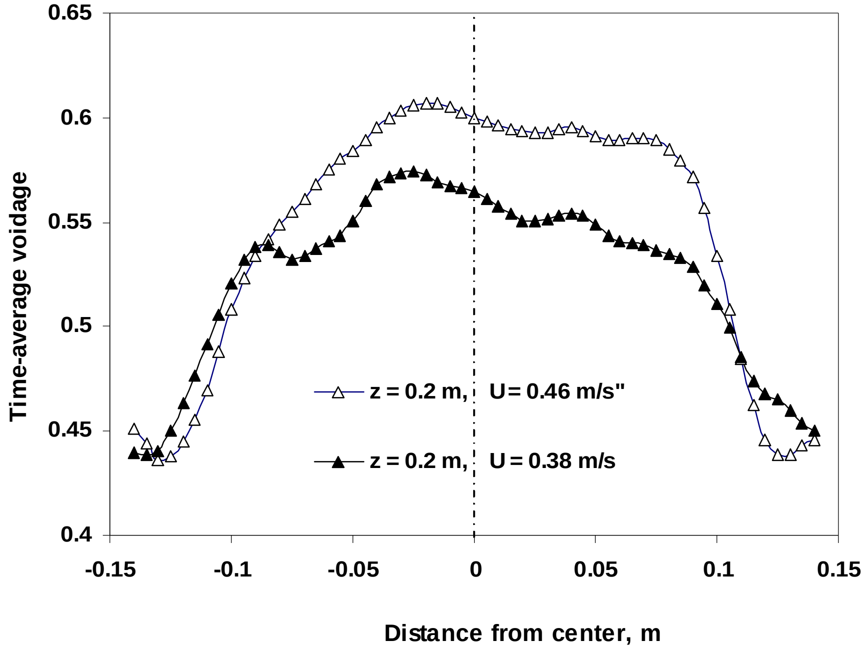


Fig. 6. Time-average voidage simulated result at different velocities

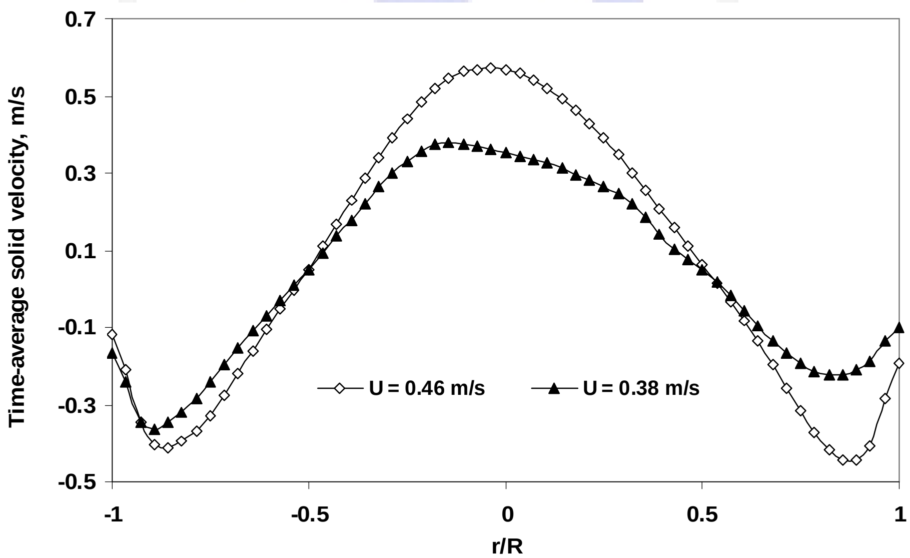


Fig. 7. Time-average voidage simulated result at different velocities

To study local solid velocity profile in the bed, we encountered to limitation of experimental data. As the available measured data was not local and not measured by LDA or PIV method. However

the trend of simulated result was investigated.
Fig. 7 is shown the time-average local solid velocities at 0.2 meter above the distributor for different superficial gas velocities. It. shows that the center of the bed has a maximum time-average velocity because of the existence of large bubble and lack of wall effects. It is visible that increase of superficial gas velocity cause an increase in the solid velocity at the center of the bed and inverse behavior of particles near the wall, i.e. the particles that have high upward velocity in the center will have a high downfall velocity near the wall. This reverse flow causes accumulation of solid particle by the wall. This is the main cause of erosion in many operative beds. The center of the bed is the dilute region of the particles.

Conclusion

The Two-Fluid Model, using the kinetic theory of granular flow for description of solid phase rheology, can predict the gas-solid fluidized bed behavior suitably. The result of CFD simulation, considering the pattern flow (a bubbling regime), are in a good agreement with the experimental results such as: bubble formation, distribution of the bubble through the bed, expansion of the bed and pressure drop. Comparison between time-average local voidage profile obtained from the result of simulation and experiment studies represents a similar trend. The time-average of the local solid velocity profile from simulation results represents an acceptable trend as well. In this study, the best constitutive equations were used to obtain the acceptable results. Also to reduce the computational effort, algebraic granular energy-balance equation was used instead of full energy-balance equation, and this simplification does not lead to different results. As a consequence, the simulation of the dense gas-solid fluidized bed by the Geldart B, which uses the kinetic theory of granular flow, the algebraic granular equation can be solved instead of the full granular energy-balance equation.

Notation

C_D drag coefficient, dimensionless
d_i diameter, m
e_{ss} restitution coefficient, dimensionless
$g_{0,ss}$ radial distribution coefficient, dimensionless
g acceleration due to gravity, m/s2
H expanded bed height, m
H_0 static bed height, m
$\bar{\bar{I}}$ stress tensor, dimensionless
I_{2D} second invariant of the deviatoric stress tensor, dimensionless
$k_{\Theta s}$ diffusion coefficient for granular energy, kg/sm
K_{gs} gas/solid momentum exchange coefficient, dimensionless
P pressure, Pa
r radial coordinate, m
Re Reynolds number, dimensionless
t time, s
U superficial gas velocity, m/s
v_i velocity, m/s
z height coordinate measured from distributor, m
v_s' fluctuating particle velocity of the particulate phase
$< v_s' >$ ensemble averaging of fluctuating particle velocity

Subscripts

g gas
s solids
i general index
t terminal (e.g. v_t is the terminal velocity)
mf minimum fluidization
T stress tensor

References

[1] D. Kunii, O. Levenspiel, Fluidization Engineering, second edition, 1991.
[2] Versteeg and Malalsekera, "An Introduction to Computational Fluid Dynamics: the finite volume method", 1960.
[3] Grace, J.R., Taghipour, F., 2004. Verification and validation of CFD models and dynamic similarity for fluidized beds. Powder Technology 139, 99–110.
[4] Gera, D., Gautam, M., Tsuji, Y., Kawaguchi, T., Tanaka, T., 1998.Computer simulation of bubbles in large-particle fluidized beds. Powder Technology 98, 38–47.
[5] Matteo Chiesa, V. Mathiesen, J. A. Melheim, B. Halvorsen," Numerical simulation of particulate flow by the Eulerian–Lagrangian and the Eulerian–Eulerian approach with application to a fluidized bed", Computers and Chemical Engineering., 29, 291–304(2005).
[6] M.A. van der Hoef, M. van Sint Annaland, J.A.M. Kuipers," Computational fluid dynamics for dense gas–solid fluidized beds: a multi-scale modeling strategy", Chemical Engineering Science 59 (2004) 5157 – 5165.
[7] Kuipers, J. A. M., Hoomans, B. P. B., and van Swaaij, W. P. M., "Hydrodynamic Models of Gas-Fluidized Beds and thek Role for Design and Operation of Fluidized Bed Chemical Reactors", in "Fluidization IXT, Knowlton, T. M. and Fan, L.-S., Eds., Engineering Foundation, NY, pp. lS-30,1998.
[8] Pain, C.C., Mansoorzadeh, S., de Oliveira, C.R.E., 2001. A study of bubbling and slugging fluidized beds using the two-fluid granular temperature model. International Journal of Multiphase Flow 27,527–551.
[9] van Wachem, B.G.M., Schouten, J.C., Krishna, R., van den Bleek, C.M., "Eulerian simulations of bubbling behavior in gas–solid fluidized beds". Computers & Chemical Engineering 22, S299–S306,1998.
[10] Anderson, T. B. and R. Jackson, "A fluid mechanical description of fluidized beds", Ind. Eng. Chem. Fundam., vol. 6, p.p. 527, 1967
[11] Ishii, M., Thermo-Fluid Dynamic Theory of Two-Phase Flow, direction des Etudes et Recherches d'Electricite´ de France, Eyrolles, Paris, 1975

[12] van Wachem, B. G. M.; Schouten, J. C.; Krishna, R.; van den Bleek, C. M.; Sinclair, J. L. Comparative analysis of CFD models of dense gas-solid systems. AIChE J. 2001, 47, 1035- 1051.

[13] Lun, C.K.K., Savage, S.B., Jeffrey, D.J., Chepurniy, N., 1984. Kinetic theories for granular flow: inelastic particles in Couette flow and slightly inelastic particles in a general flow field. Journal of Fluid Mechanics 140, 223–256.
[14] Jenkins, J. T.; Savage, S. B. A theory for the rapid flow of identical, smooth, nearly elastic, spherical particles. J. Fluid Mech. 1983, 30, 187-202.
[15] Ding, J. and Gidaspow, D., "A Bubbling Fluidization Mode1 Using Kinetic Theory of Granuiar Flow". AIChE Journal, Vol. 36,4, pp. 523-538, 1990.
[16] Ding, J.M., Lyczkowski, R.W., "Three-dimensional kinetic-theory modeling of hydrodynamics and erosion in fluidized-beds". Powder Technology 73, 127–138, 1992.
[17] Enwald, H., Almstedt, A.E., "Fluid dynamics of a pressurized fluidized bed: comparison between numerical solutions from two-fluid models and experimental results". Chemical Engineering Science 54, 329–342, 1999.

[18] Seu-Kim, H. and Arastoopour, H., "Simulation of FCC Particles Flow Behavior in a CFB Using Modified Kinetic Theory", Can. J. Chem. Eng., Vol. 73, 5, pp.603-6 1 1, 1995.

[19] Syamlal , M., O'Brien, T.J., 1989. Computer simulation of bubbles in a fluidized bed. A.I.Ch.E. Symposium Series 85, 22–31.

[20] Gidaspow, D., 1994. Multiphase Flow and Fluidization: Continuum and Kinetic Theory Descriptions. Academic Press, San Diego.

[21] Wen, C.-Y., Yu, Y.H., 1966. Mechanics of fluidization. Chemical Engineering Progress Symposium Series 62, 100–111.

[22] Syamlal, M.; Rogers, W.; O'Brien T. J. MFIX Documentation: Theory Guide; National Technical Information Service Springfield, VA, 1993; Vol. 1.

[23] Boemer, A., H. Qi, U. Renz, S. Vasquez, and Boysan, "Eulerian computation of fluidized bed hydrodynamics-a comparison of physical models", Proc. Of the Int. Conf. on FBC, Orlando, FL, p. 775, 1995

[24] Taghipour, F.; Ellis, N.; Wong, C. Experimental and Computational Study of Gas-Solid Fluidized Bed Hydrodynamics. Chem. Eng. Sci. 2005, 60, 6857-6867.

[25] Nasr Esfahani, M., Ramimi, R., Hosseini, S. H., 2006. Hydrodynamics of fluidized bed by computational fluid dynamic (CFD). 11 th Iranian public congress, Tarbiat Modares university.

[26] FLUENT Inc. FLUENT 6.3 User's Guide; FLUENT Inc.

[27] S. Zimmermann , F. Taghipour. CFD Modeling of the Hydrodynamics and Reaction Kinetics of FCC Fluidized-Bed Reactors. Ind. Eng. Chem. Res. 2005, 44, 9818-9827