Refereed Proceedings

The 12th International Conference on Fluidization - New Horizons in Fluidization Engineering

Engineering Conferences International *Year* 2007

A CFD Study into the Influence of the Particle Particle Drag Force on the Dynamics of Binary Gas Solid Fluidized Beds

Olumuyiwa Owoyemi* Paola Lettieri†

*University College London
†University College London, p.lettieri@ucl.ac.uk

CFD STUDY INTO THE INFLUENCE OF PARTICLE PARTICLE DRAG FORCE ON THE DYNAMICS OF BINARY GAS SOLID FLUIDIZED BEDS

Olumuyiwa Owoyemi, Paola Lettieri
Department of Chemical Engineering, University College London, Torrington Place, London, WC1E 7JE, UK.

ABSTRACT

This paper presents a preliminary CFD study into the effect of the particle-particle interphase momentum transfer term on the mixing and bubble dynamics of a binary gas solid fluidized consisting of particles which only differ in size. A new fluid dynamic model, implemented within a commercial CFD code, CFX4.4, is used to model the bi- nary mixture. The governing fluid dynamic relationships for the solid phases and the fluid phase are based on “the elastic force” concept from the Particle Bed Model (1, 2). The solids pressure for each of the particulate phases is not taken into consideration in this model, however solid phase compaction for the each of the particulate phases is controlled via a numerical scheme supporting experimental validation of the computational results is also presented herein. The computational strategy employed in this work involved the use of two case studies, where one case study was carried out without the implementation of the particle-particle drag force. Results from the CFD simulations in agreement with the experimental results, initially showed an increase in bubble diameter at increasing bed height however the trend discontinued higher up in the bed, with the simulation in which particle-particle drag force was neglected giving the the poorest agreement.

INTRODUCTION

Many researchers to enable the correct theoretical prediction of various macroscopic phenomena encountered in gas-fluidized beds have successfully carried out the mono component computer modelling of dense gas-solid fluidized systems. CFD simulations have been carried out, by researchers, covering the whole range of Geldart classified powders with great success (3, 4, 5) with some authors even successfully validating their work with actual experimental results (6). The wide variety of case studies available in literature today is a testament to the applicability of eulerian-eulerian approach in tackling complex gas solid interaction phenomena such as that present in a fluidized bed. However a monosize system of particles seldom occurs in large-scale industrial fluidized beds. Industrially operated gas-solid fluidized beds typically consist of particles, which have a wide size distribution as well as different densities. The phenomenon of mixing and segregation pervades in this non-ideal system, which has led to a less than favourable development in the computational modelling of these systems. The continuum modelling of binary mixtures, within the sphere of Eulerian-Eulerian continuum modelling is typically carried out using two approaches. The first approach is characterized by the use of separate momentum equations to define each particle specie, this approach has been employed by Gidaspow et al. (7), Cooper and Coronella (8) and Bell (9) whilst the second approach makes use of the averaged mixture properties for the formulation of a mixture momentum equation coupled with the use of averaged constitutive relations has been employed by Van Wachem et al. (10) to predict the flow of a binary mixture in a fluidized bed. The use of separate momentum equations for each particulate phase requires an extra term to account for the collisions between particles that belong to different particle phases. This "extra" contribution is termed the particle-particle drag force. An investigation of the effect of the force on the dynamics of a binary mixture forms the primary aim of this paper.

CFD MATHEMATICAL MODEL

The governing fluid dynamic relationships for the fluid and solid phases utilize the concept of "particle phase elasticity" originally proposed by in the Particle Bed Model. The original model was initially described through one-dimensional equations by Gibilaro (2). In their model they introduced the particle phase elasticity force in the momentum balance equation of the particle phase in order to describe the transfer of momentum between particles. The elasticity term was expressed as the scalar product of the elastic modulus and the gradient of voidage in the vertical direction only. In the model proposed herein, the fluid particle interaction force is made up of the pressure gradient, drag force, derived from the expression of Di Felice (11), and the elastic force, which is the scalar product of the Elastic Modulus, E, and the gradient of the local voidage (12) parallel to the direction of the drag force. The solid stress tensor has been ignored in the current model and the solid packing is controlled via a numerical algorithm. The drag force between the two particulate phases is modelled in terms of the product of a momentum transfer coefficient and the relative velocities of the phases. Several investigators have put forward empirical correlations to account for this momentum transfer co-efficient (7, 13, 9). In this paper the drag law proposed by Syamlal et al. (13) has been used. The equations used in this work are summarised in Table 1.

EXPERIMENTAL

The experimental set-up used in this work, shown in Fig. 1, consists of a two-dimensional plexiglass rectangular column, 600 mm high, 350 mm wide and 10 mm thick. The distributor is a uniformly permeable sintered bronze rectangular plate with a thickness of 3:5 mm. Fluidizing gas, air, is supplied via rotameters. The air is also dehumidified and filtered to remove impurities present in the air mixture. Pressure taps are installed 100 mm apart along the height of the bed from which pressure readings are collected via an electronic manometer. Two-way valves are also installed on the rig to allow for instantaneous evacuation of air. The binary system investigated is characterized by components that differ in size and have the same density. Ballotini powder samples of sizes 200 μm and 350 μm with a density of

2500 kg/m^3 were used as the particle system in the fluidization experiments. The larger ballotini particle represents the jetsam particle whilst the flotsam particle is the smaller particle.

Table 1. Governing equations applied to binary gas-solid flow

Continuity	
Fluid Phase	$\frac{\partial \varepsilon}{\partial t} + \nabla \cdot (\varepsilon \mathrm{u}) = 0$ $\frac{\partial \phi}{\partial t} + \nabla \cdot (\varepsilon \mathrm{v}) = 0$
Momentum	
Fluid Phase	$\rho_f \varepsilon \frac{D_f \mathbf{u}}{Dt} = \varepsilon \nabla \cdot \Omega_f - \sum_{i=1}^{2} \left\{ \beta_i (\mathbf{u} - \mathbf{v}) \lVert \mathbf{u} - \mathbf{v} \rVert + \mathrm{E}_i \cdot \left(\nabla \varepsilon \bullet \frac{F_{D_i}}{\lVert F_{D_i} \rVert} \right) \right\} + \rho_f \varepsilon \mathbf{g}$
Solid Phase	$\rho_s \phi \frac{D_s \mathbf{v}}{Dt} = \phi \nabla \cdot \Omega_f + \beta (\mathbf{u} - \mathbf{v}) \lVert \mathbf{u} - \mathbf{v} \rVert + \mathrm{E} \cdot \left(\nabla \varepsilon \bullet \frac{F_D}{\lVert F_D \rVert} \right) -$ $\sum_{i=1}^{n} \zeta_i (\mathrm{v}_{i+1} - \mathrm{v}_i) \lVert \mathbf{v}_{i+1} - \mathbf{v}_i \rVert + \rho_s \phi \mathbf{g}$
Closure Relations	
$\Omega = -p + \mu \left[\nabla U + (\nabla U)^T \right]$ $\beta = \left[\left(\frac{17.3}{Re_p} \right)^{\alpha} + (0.336)^{\alpha} \right]^{\frac{1}{\alpha}} \frac{\rho_f (1-\varepsilon) \varepsilon^{-3.8}}{d_p}$ $\alpha = 2.55 - 2.1 \left[\tanh (20\varepsilon - 8)^{0.33} \right]^3$ $E = -\frac{2}{3} d_p \left[F_D (-3.8\varepsilon^{-1} + \kappa) - (1-\varepsilon)(\rho_p - \rho_f) g \right]$ $\kappa = \left\{ -\frac{1}{\alpha^2} \ln \left[\left(\frac{17.3}{\mathrm{Re}_p} \right)^{\alpha} + (0.336)^{\alpha} \right] + \frac{1}{\alpha} \left[\frac{ \left(\frac{17.3}{\mathrm{Re}_p} \right)^{\alpha} \ln \left(\frac{17.3}{\mathrm{Re}_p} \right) + 0.336^{\alpha} \ln 0.336 }{ \left(\frac{17.3}{\mathrm{Re}_p} \right)^{\alpha} + (0.336)^{\alpha} } \right] \right\} \times$ $\left\{ -6.3 \left[tanh (20\varepsilon - 8)^{0.33} \right]^2 \right\} \times \left\{ \sec h^2 (20\varepsilon - 8)^{0.33} \right\} \times \left\{ 6.6 (20\varepsilon - 8)^{-0.67} \right\}$	

The 12th International Conference on Fluidization - New Horizons in Fluidization Engineering, Art. 85 [2007]

Table 2. Particle particle drag model used in this work

Syamlal et.al. (13)
$$\zeta = \frac{3(1+e)\left(\frac{\pi}{2}+C_{fr}\frac{\pi^2}{8}\right)\phi_k\rho_k\phi_l\rho_l\left(d_k+d_l\right)^2 g_o}{2\pi\left(\rho_k d_k^3+\rho_l d_l^3\right)}\left\lvert \mathbf{v}_k-\mathbf{v}_k \right\rvert$$
Solid packing correction (14)
$$\phi_{kl} = \left[\left(\Phi_k-\Phi_l\right)+\left(1-a\right)\left(1-\Phi_k\right)\Phi_l\right]\left[\Phi_k+\left(1-\Phi_k\right)\Phi_l\right]\frac{X_k}{\Phi_k}+\Phi_l$$ $$\text{for} X_k \leqslant \frac{\Phi_k}{\Phi_k+\left(1-\Phi_k\right)\Phi_l}$$ $$\phi_{kl} = \left(1-a\right)\left[\Phi_k+\left(1-\Phi_k\right)\Phi_l\right]\left(1-X_k\right)+\Phi_k$$ $$\text{for} X_k \geqslant \frac{\Phi_k}{\Phi_k+\left(1-\Phi_k\right)\Phi_l}$$ $$a=\sqrt{\frac{d_l}{d_k}}d_k \geqslant d_l, X_k=\frac{\phi_k}{\phi_k+\phi_l}$$

The bed is initially completely segregated where the flotsam particle is first filled to height of 150 mm and the jetsam particle to a height of 300 mm, this corresponds to 0:88 kg and 0:97 kg of the flotsam and jetsam particles respectively. Table 3 shows a summary of the properties of materials used in this work. The experiments were carried out at a superficial gas velocity of 0:25 m/s, required to give a mixing index of 0.80. This fluidizing velocity was determined using the semi-empirical correlation developed by Wu and Baeyens. (16). The correlation is based on the visible bubble flow rate, U_o - U_{mf}, which is thought to be the real driving force behind mixing and segregation in gas fluidized beds. Digital video recordings of the fluid bed were made to analyse the development of bubble dynamics in the fluid bed and determine the bubble size at the fluidizing velocity. Images captured by means of a web camera at 14 frames/s, for 80 s and were then subsequently analysed using Optimas 6.0, image analysis software.

SIMULATIONS

In this work, all simulations were carried using a commercial CFD package, CFX 4.4. The governing equations described in Table 1 as well as the particle-particle drag force described in Table 2 was implemented into this code. A 2-D computational grid in which front and back wall effects are neglected was used in this work. The 2-D grid used was based on earlier work done by Lettieri et al. (5). The left and right walls of the domain were modelled using no-slip velocity boundary condition for both phases. Dirichlet boundary conditions are employed at the bottom of the bed to specify a uni- form gas inlet velocity. A pressure boundary condition is specified at the top of the bed and set to a reference value of 1:015 × 10^5 Pa. The distributor was made impenetrable for the solid phase. A second order Discretization scheme, SUPERBEE, was used for all equations to improve the computational prediction of bubble shape and behaviour The fluidization conditions used for all simulations are summarized in Table 4. Two dif- ferent simulations were carried out in which the mass fraction of both large and small particles was set to 0.5. The fluidized bed was initially filled in two layers in which the flotsam particle occupied the bottom half of the bed whilst the jetsam particle occupied the top half of the bed. The Particle-particle drag law shown in Table.2 was implemented for the first case and was omitted for the second "placebo" case. Both simulations were performed for a total of 10 secs (real time). The simulations were carried out using two Dell Xeon P4 3.2 Ghz Machines.

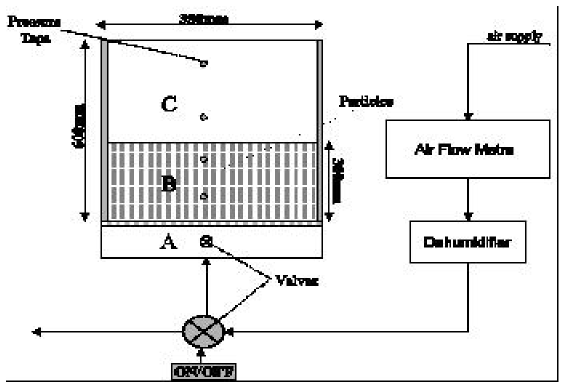

Figure 1. Experimental Apparatus (A) Windbox (B) Fluidized Bed (C) Freeboard

Table 3. Particle Physical Properties

Physical Property	Small Particle	Large Particle
Diameter	200μm	350μm
Density	2500kg/m^3	2500kg/m^3
Geldart Group	B	B

Table 4. Computational Parameters used in the CFD simulations

Description	Symbol	Value	Comments
Gas Density	ρ_g[kg/m^3]	1.2	
Gas Viscosity	μ_g[Pa s]	1.85E10^5	
Bed Height	H_{bed} [m]	0.60	
Settled Bed eight	H_s	0.30	
Grid cell size	Δx and Δy [m]	0.005	Square cells
Time step	Δt	10^{-4}	Time step
Superficial gas velocity	U_o[m/s]	0.25	
Co efficient of restitution	*e*	0.97	Syamlal et al. (13)
Co efficient of friction	C_{fr}	0.15	Syamlal et al. (13)

RESULTS AND DISCUSSION

Bed Voidage

Figures 2 shows a comparison between the experimental and simulated voidage profiles obtained using the Syamlal et al. (13) particle drag correlation. Qualitatively it is easy to observe that the both simulations capture the mixing phenomenon of the jetsam and flotsam components, shown in the experimental snapshots. It can also be observed that both simulations show bubbling phenomenon, albeit modestly, associated with the experimental bulk dense phase alongside other macroscopic phenomena like bubble coalescence and bed expansion. A distinction in terms of the effect the particle-particle drag model is immediately discernible from the computational

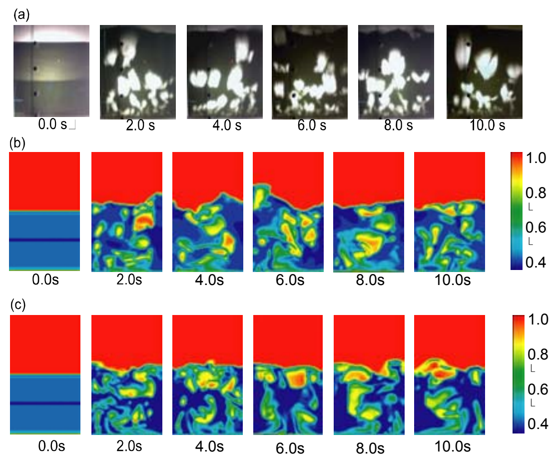

Figure 2: Snapshots showing the (a) experimental bed (b) computational bed obtained using the Particle drag expression by Syamlal et al. (13) and (c) computational bed obtained using no particle drag expression.

snapshots, it can be seen that the snapshot which has an implemented particle drag model (see Fig 2 b) displays a more vigorous bubbling bed activity, especially near the bottom of the bed, when compared with the simulation in which the particle drag model was neglected (see Fig 2 c).

Bed Height

A quantitative comparison of bulk bed properties averaged after 2s of simulation, shown in Table 5, provides an alternative way of discriminating between the simulations. The quantities have been averaged to reduce the effect of perturbations associated with the startup of the bed. One conclusion that can be drawn immediately from Table 5 is that the non-implementation of a particle drag model results in a lower prediction of averaged bed height and a lower bed voidage as consequence of the "lack of frictional hindrance", this leads to an unrealistic prediction of segregation in the bed.

Table 5. Comparison of time averaged macroscopic fluidization indicators with experimental data

Drag Model	Bed Height (m)	Bed Voidage
Experimental	0.365	0.520
Syamlal et al.(13)	0.355	0.503
No Drag implemented	0.347	0.490

Bubble Properties

The analysis of bubble diameter has been carried out by comparing simulated values with results obtained from experimental data analysis. In defining a bubble, an appropriate voidage has to be selected as the boundary between the emulsion and the gas phase. In this work, a voidage contour of 0.80 as been assumed for the simulation measurements. This subjective number is in conformity with numbers used in literature. The experimental analysis of bubble diameter was carried out using Optimas 6.0, image analysis software. The computational bubbles were obtained using the numerical algorithm recently advanced by Mazzei and Lettieri (15). Figure 4 shows a comparison between the simulations and the experimentally obtained bubble diameters. The simulations, in agreement with the experimental results, show an increase in bubble diameter at increasing bed height, initially; however the trend discontinues higher up in the bed. A possible explanation for this phenomenon might be the route of exit of the excess gas above that required for minimum fluidization. According to the two-phase theory, the quantity of gas appearing as bubbles should be equal to gas available in excess of that which is required for minimum fluidization. However several investigators have found that the above statement does not strictly hold true and indeed it is has been established that a part of the “excess gas” leaves the bed via the particulate phase. The above phenomenon could be at play in the computational simulations where a predominant excess gas flow through the particulate phase would lead to smaller computational bubble sizes.

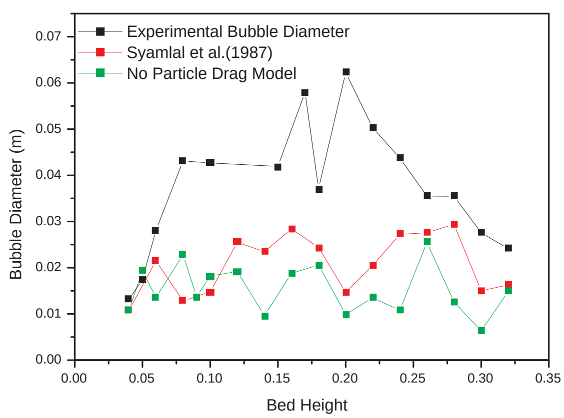

Figure 3: Comparison of experimental with the simulated bubble diameter for (a) Syamlal et al (13) particle drag model and (b) no particle drag model

CONCLUSION

This work has described the effect of the particle particle drag force on the fluid dynamics of a binary gas-solid mixture. The governing fluid dynamic relationships for the solid phases and the fluid phase were based on concepts from the Particle Bed Model (2). Results from the CFD simulations showed a match qualitatively between the experi- mental and computational snapshots, with the snapshots clearly showing the mixing phenomenon of the jetsam and flotsam components. Averaged bed height predictions were within 5% of the experimental results wherein the drag relation by Syamlal et al.(13) giving the best agreement. The simulations, in agreement with the experimental results, initially showed an increase in bubble diameter at increasing bed height how- ever the trend discontinued higher up in the bed, with the simulation in which particle drag force was neglected giving the poorest agreement.

ACKNOWLEDGMENTS

The authors wish to acknowledge financial support from the Engineering Physical Science Research Council (EPSRC). The authors wish to acknowledge financial support lso from Tioxide Europe Ltd.

REFERENCES

1. Foscolo, P.U., Gibilaro, L., (1987). Fluid Dynamic Stability of Fluidized Suspen-sions: The Particle Bed Model; Chem. Eng. Sci. 42(6), 1489-1500.
2. Gibilaro, L., (2001). Fluidization Dynamics; Butterworth Heinemann, London.
3. Boemer, A., Qi,H., Renz,U., (1997). Eulerian Simulation of bubble formation at a jet in a two-dimensional fluidized bed , Int. J Multiph. Flow, 23 (5) 927-944
4. Gelderbloom, S.J., Gidaspow D., Lyczkowski, R. W., (2003). CFD simulations of Bubbling/Collapsing Fluidized Beds for Three Geldart Groups, AIChE Journal, 49, 844-858.
5. Lettieri, P., Cammarata, L., Micale, G., Yates, J., (2003a). CFD Simulations of Gas Fluidized Beds using Alternative Eulerian-Eulerian Modelling Approaches, IJCRE, 1, A5.
6. Owoyemi, O.,Lettieri, P., Place, R., (2005). “Experimental validation of Eulerian-Eulerian simulations of rutile industrial powders” Ind. Eng. Chem. Res. 44 (26).
7. Gidaspow,D., Syamlal, M., and Seo, Y., (1985). Hydrodynamics of Fluidization of Single and Binary Size Particles: Supercomputer Modelling. In Proc of the Fifth Engineering Foundation Conference on Fludization, 5,1-8.
8. Cooper, S., and Coronella, C.J., (2005). CFD simulations of particle mixing in a binary fluidized bed.Powder Technology,151,27-36.
9. Bell, R.A., (2000). Numerical Modelling of Multi-Particle Flows in Bubbling Gas -Solid Fluidized Beds, Swinburne University of Technology,PhD Thesis.
10. Van Wachem,B.G.M., Schouten, J.C., Krishna, R., van den Bleek, C.M., and Sin-clair, J. .,(2001b). CFD Modeling of Fluidized beds with a Bimodal Particle Mixture, AIChE Journal, 47,1291.
11. Di Felice, R.,(1988). Liquid Fluidization of Binary Solid Mixtures, University of Lon- don, PhD Thesis.
12. Mazzei, L., Lettieri, P., Elson, T., Colman, D. (2006). A reveised mono-dimensional particle bed model for fluidized beds, Chem .Eng Sci.61: 1958-1972
13. Syamlal, M., (1987): “ The Particle-Particle Drag Term in a Multiparticle Model of Fludization,”Topical Report, DOE/MC/215353-2372, NTIS/DE87006500, National Technical Information Service, Springfield, VA.
14. Fedors, R.F. and Landel, R.F., (1979). An Emprical Method of estimating the void fraction in Mixtures of Uniform particles of Different sizes , Powder Technology.23,
225-231
15. Mazzei, L., Lettieri, P. (2006) : A numerical alogarthim for the analysis of the bubble dyanmics in 2D Fluidized Beds Simulated by means of CFD Multiphase Codes, Int .Journal of Chemical Reactor Engineering. in press
16. Wu S.Y., and Baeyens, J., (1998). Segregation by size difference in gas fluidized beds, Powder Tech.98, 139-150