Analysis of different drag Models for bubbling fluidized beds using CFD

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1. INTRODUCTION

Gas fluidized beds have found widespread use in the petroleum, chemical, metallurgical and energy industries. Fluidization is a widely employed technology in industry, which involves highly complex fluid-solid flow phenomena. Computational modelling has been successfully applied to single-phase systems and is recently being directed towards the development of predictive tools for multiphase systems, including fluidization.

During this work, Eulerian-Eulerian modelling approaches based on the two fluid model have been assumed as the most suitable choice when simulating the hydrodynamics of gas fluidized beds. In these approaches, fluid-particle interactions are modelled via the inter-phase drag term in the momentum conservation equations, which is semi-empirical in nature.

A proper choice of the empirical drag model as well as of the drag correlation is therefore important. This paper reports on CFD simulations of gas fluidized beds using a commercial code developed by AEA Technology, CFX-4. Eulerian-Eulerian models such as the granular kinetic models (Gidaspow, 1994) and the particle-bed model (Gibilaro, 2001) have been investigated. The particle bed model has been recently implemented in CFX-4 for 2D simulations by Lettieri et al. (2003) and in 3D by Cammarata et al.(2003) . The proposed work will extend the previous work by presenting a sensitivity analysis of different drag models on the simulations of the bubbling fluidization of a Geldart Group B material.

2. EULERIAN-EULERIAN TWO PHASE MODELS

Both the granular kinetic model and the particle bed model are used during this work. Both models assume the gas and solid phases as interpenetrating continua. Mass and momentum equations for each phase are solved using a Eulerian-Eulerian description. The full sets of two-dimensional equations as well as the underlying assumptions for both models are reported in Lettieri et al. (2003). The two approaches mainly differ in the way particle-particle interactions are translated into the continuum formulation of the solid phase. According to the granular kinetic model, the fluid dynamic behaviour of the solid-phase is modelled based on an analogy with the gas kinetic theory. The granular temperature, Θ_s, is introduced to describe the kinetic energy associated with the particle fluctuations, v_s',

$$\Theta_s = \frac{1}{3} < {v'_s}^2 > \tag{1}$$

The granular temperature is estimated by solving an energy balance, which is added to the solid-phase mass and momentum balance equations.

Together with the granular temperature, granular kinetic theory introduces two other parameters: the coefficient of restitution, e_s (where $0<e_s<1$), to account for the non-ideal collisions or inelasticity of the solids, and the radial distribution function, g_o. This function gives a statistical measure of the probability of particle contacting and therefore controls the solids volume fraction so that the maximum packing is not exceeded. It is worth mentioning that all simulations performed adopted a coefficient of restitution $e_s=0.9$ and the radial distribution function given by Ding and Gidaspow (1990).

The particle bed model describes the particle-particle interactions based on the hydrodynamic forces involved in the gas-solid flow (gravity, drag and buoyancy). Direct collisions between particles are not a necessary pre-condition for the particles to exchange momentum. The continuum solid phase is regarded as an elastic fluid, which is capable of opposing imposed deformations. A *particle phase elasticity* is derived and an additional force, the so called particle phase elasticity force, dependent on the solids volume fraction gradient is introduced in the solid phase momentum balance equation (Foscolo et al., 1987). The momentum transfer between particles is modelled only according to the elastic mechanism described above and no viscous stress terms are included in the solid phase momentum equations.

3. INTERPHASE MOMENTUM TRANSFER MODELS

The drag force per unit volume between the dispersed and the continuous phase is modelled as the product of the interphase drag function β and the relative velocities of the two phases, e.g. for the particle phase in the i-th direction.

$$F_i^{(d)} = \beta\left(u_i^{(f)} - u_i^{(s)}\right) \tag{2}$$

where f and s refer to the fluid and solid phase respectively.

The general expression for β, the interphase drag function, is given in Eq. (3) below.

$$\beta = \frac{3}{4} C_D \frac{\rho_f \varepsilon_s \varepsilon_g \left|\left(\mathbf{u}_f - \mathbf{u}_s\right)\right|}{d_p} .E \tag{.3}$$

The empirical function, E, from above is a correction coefficient introduced in order to take into account the presence of high particle concentration in the bed. It is generally modelled as a function of the solids volume fraction.

$$E = \left(1 - \varepsilon_s\right)^n \tag{4}$$

Where ε_s is the solid volume fraction.

It is clear that the computation of the interphase drag term requires knowledge of the drag coefficient, C_D. This term is calculated from correlations based on the particle Reynolds number, Re_p, which is defined as:

$$\mathbf{Re}_p = \frac{\rho_g d_p \left| u_f - u_s \right|}{\mu_g} \tag{5}$$

In this work, the correlations established by Ihme et al. (1972) and Dalla Valle (1948) for the viscous flow regime are used to test the sensitivity of the two models, i.e. the GKT model and the PBM, to changes in the drag coefficient. The drag coefficients are reported in Table 1. Simulations are also carried out using the Ergun (1952) expression for β to study the effect of the change on the simulated fluid bed dynamics. The last case investigated deals with the exponent n used in the general expression for E, see Eq. (3) above. Simulations are carried out using expressions derived by Wen and Yu (1966) and Di Felice (1994). It should be noted that the same drag coefficients were used during the above study. The exponents used are detailed in Table 2 below.

Table 1: Drag Coefficient Correlations, C_D.

Authors	Correlation	Range
Ihme et al (1972)	$C_D = 24/\mathbf{Re}_p + 5.48\,\mathbf{Re}_p^{-0.573} + 0.36$	$\mathbf{Re}_p < 800$
Dallavale (1948)	$C_D = \left(0.63 + 4.8\,\mathbf{Re}_p^{-0.5}\right)^2$	$\mathbf{Re}_p < 2 \times 10^5$

Table 2: Drag Models Implemented

Wen and Yu(1966**)**	
$n = -1.65$	$\beta = \frac{3}{4} C_D \frac{\rho_f \varepsilon_s \left|(\mathbf{u}_f - \mathbf{u}_s)\right|}{d_p} . (1-\varepsilon_s)^{-1.65}$
Difelice(1994)	
$n = 2 - \gamma$, where γ is given below $\gamma = 3.7 - 0.65 \exp\left(- \frac{\left[1.5 - \log(\varepsilon_g \,\mathrm{Re})\right]^2}{2} \right)$	$\beta = \frac{3}{4} C_D \frac{\rho_f \varepsilon_s \left|(\mathbf{u}_f - \mathbf{u}_s)\right|}{d_p} . (1-\varepsilon_s)^{2-\gamma}$
Ergun (1952)	
$\beta = 150 \frac{\varepsilon_s^2 \mu_f}{(1-\varepsilon_s) d_p^2} + 1.75 \frac{\varepsilon_s \rho_f \left|\mathbf{u_f} - \mathbf{u_s}\right|}{d_p}$	

4. Work in progress

A rectangular geometry of dimensions 600mm ×300mm ×10mm isused in the calculations, with gas entering with a uniform velocity at the distributor plate. The lateral walls, front and rear boundary planes were modelled using no-slip velocity boundary conditions for both phases. Dirichlet boundary conditions are employed at the bottom of the bed to specify a uniform gas inlet velocity. Pressure boundary conditions are employed at the top of the freeboard. This implies Dirichlet boundary conditions on pressure, which is set to a reference value of 1.015 ×10^5 Pa and Neumann boundary conditions to the gas flow. The fluidization conditions used for the numerical simulation are summarised in Table 3 below.

The results presented relate to the bubbling fluidisation of a sand-like Geldart Group B material having a particle diameter of 300μm.

Table3. Test conditions used for numerical simulations

Solid Density =2500kg/m^3,	Solid Viscosity =1×10^{-8} Pa sec
Gas Density = 1.2kg/m^3	Gas Viscosity =1.75×10^{-5} Pa sec
Particle Diameter = 350μm	Initial Solid Volume fraction = 0.590
Initial Bed Height = 0.29m	Superficial Velocity = 0.25m/s
Ambient pressure =1 atm	Grid Spacing = 0.005m

The analysis of the results will discuss the influence of using different drag models on bed expansion, average fluid bed voidage and average bubble hold-up. A comparison between simulated bubble size and predictions obtained using the Darton et al. (1977) equation will also be discussed.

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6. References

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