Numerical Simulation of 2-D Bubbling Fluidized Bed Dryer Using Sawdust as Bed Material
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Abstract: The hydrodynamics of a fluidized bed dryer for sawdust particles ($d_p = 600 \mu m$) is simulated using Eulerian–Eulerian approach. The governing equations were discretized using a finite volume method with uniform grid size of 0.005 m. For unsteady formulation, 2\textsuperscript{nd} order implicit scheme is used and first order upwind, algorithm is used for discretization of convective terms. In the present investigation, gas and solid phase interaction for the calculation of drag coefficient is calculated with the help of Syamlal–O’Brien drags model. It was focused mainly on the effect of superficial velocity of air on the dynamics of fluidized bed dryer. The fluidization behavior for sawdust particles, distributions of sawdust and air inside the dryer are also discussed.

Keywords: Fluidized bed dryer, Sawdust, CFD modelling, Fluidization

Introduction
Fluidized bed dryer is most popular in the family of dryers for drying of biomass [1]. It is extensively used for drying wet particulate and granular materials [2, 3]. There are several benefits of fluidized beds over other type of biomass dryers due to large contact area between the solids and the gas per unit bed volume, high levels of intermixing of the particulate phase, high transfer coefficients of heat and moisture between solids and gas, high relative velocities between the fluid and the dispersed solid phase. All these factors reduce the drying time without the disadvantage of damaging the materials [3, 4, 5]. This process is comparatively economical and it involves low energy consumption, high throughput [2]. The fluidized bed dryer could thus be a potential compact dryer for biomass drying [6]. In a gas- solid fluidized bed dryer, there are dissimilar phases which all contribute to the taking away of moisture from the wet particles [7]. Generally in fluidized bed drying the mode of operation could be either batch or continuous. For small scale production the mode of batch drying is used while the continuous drying systems are used for large-scale production [8]. Fluidization is mainly influenced by the characteristics of the particles. Geldart (1972) [9] classified particle behaviour in fluidization into four groups, which are widely accepted and used in fluidized bed modelling.

- Group A: Small particles (30–150 $\mu m$), and low density (<1.4 g/cm$^3$). For this group of particles fluidization is easy, smooth and homogeneous. It makes possible operating with low gas flows and controlling the growth and speed of the bubbles.
- Group B: Particles with medium diameter (40–500 $\mu m$) and density (1.4–4 g/cm$^3$). The fluidization is good for high gas flow rates and bubbles tend to grow a lot and appear at the beginning of fluidization.
- Group C: Very small particles with diameter ($d < 30 \mu m$). Fluidization for this group particle is difficult.
- Group D: Dense and larger particles having diameter ($d > 500 \mu m$). Fluidization is difficult and non-uniform, ideal for spouted beds.

Modelling and simulation can be useful for optimizing the any dryer design as well as for designing the new dryers and its operation with minimal temporal and financial costs [10]. The CFD tool is generally used to analyze fluidized bed behaviour, but the modelling present’s numerical instability in the equations because the gas–solid inter phase is transient and known only in some regions [11]. The numerical simulation of the fluidized bed is based on the classical equations of mass conservation, energy and momentum together. These equations illustrate the interactions between the phases. This is based on chemical kinetics in experimental correlations and models derived from the Kinetic Theory of Granular Flows (KTGF) [12]. The investigation on gas–solid flow system is more popular after the development of kinetic theory for gas–solids two-phase flow based on the theory of non-uniform dense gases [13]. The models, which are proposed for the study of hydrodynamics of fluidized bed, can be classified in two broad groups: 1) Computational Fluid Dynamics (CFD) based models; 2) Deterministic (Engineering) models such as two-phase, three-phase models [14]. The two-phase model consists of one dense phase, or emulsion, formed by a large number of particles, and a dilute phase, or bubble, without particles [12]. The CFD based model is based on a continuum
assumption of phases, which provides a field description of the dynamics of each phase called “Eulerian–Eulerian” (two-fluid model) or the Lagrangian trajectory for the study of motion of individual particles coupled with the Eulerian formulation for gas called “Eulerian–Lagrangian” (trajectory model) [14]. The Eulerian multiphase model allows for the modelling of multiple separate yet interacting phases. The phases can be liquids, gases or solids in nearly any amalgamation [15]. The Eulerian–Lagrangian approach handles erratic and time dependent particle sizes in a natural way by tracking each individual particle sizes with its physical properties [10].

In the present article, CFD modelling has been carried out to dynamically simulate the sawdust particles and investigate the distribution characteristics of the particles in a bubbling fluidized bed dyer. In particular, the influences of superficial velocity of air on fluidization behaviour and distribution characteristics of sawdust particles in the bubbling fluidized bed dyer are investigated using Euler-Euler model.

2 Model Descriptions

Modelling is an essential part of CFD analysis for any real experimental set-up. It establishes a basic similarity between real and modelled geometry. The 2-D geometry of fluidized bed dryer has a dimension of 0.4 m height and diameter 0.095 m. For the present study, equivalent 2-D geometry of the bubbling fluidized bed dryer has been made in a GAMBIT 2.4.6.

2.1 Details of the geometry and mesh

After making the geometry part of the fluidized bed dryer, meshing was done for the CFD analysis. The 2-D model of the fluidized bed dryer and computational grid for 2-D simulations is shown in Figure 1. Meshing plays a vital role in CFD analysis. It provides a basic element over which initial and boundary condition can be applied on elemental level and solved by different methods to provide the solution in the whole region. The whole square shaped computational domain is divided into 1520 elements with grid size of 0.005 m.

2.2 Mathematical modelling

The Eulerian–Eulerian model based on the kinetic theory of granular flow is applied to simulate the air–solid turbulent flow. The Euler-Euler (EE) multiphase flow approach has been used for developing the CFD model in which the gas phase and solid particles were treated as inter-penetrating continuums. The general conservation equations have been formulated for mass and momentum for the hydrodynamic study of sawdust. The additional term appears in the momentum equations due to the interacting forces among the gas and solid. The conservation equations for mass and momentum are written as follows:

\[
\frac{\partial}{\partial t} \left( \alpha_g \rho_g \right) + \nabla \cdot \left( \alpha_g \rho_g \vec{v}_g \right) = -\alpha_g \nabla p + \nabla \cdot \tau_{gg} + \alpha_g \rho_g \vec{g} - K_{gg} \left( \vec{v}_s - \vec{v}_g \right) \quad (3)
\]

\[
\frac{\partial}{\partial t} \left( \alpha_s \rho_s \vec{v}_s \right) + \nabla \cdot \left( \alpha_s \rho_s \vec{v}_s \right) = -\alpha_s \nabla p + \nabla \cdot \tau_{ss} + \alpha_s \rho_s \vec{g} - K_{ss} \left( \vec{v}_s - \vec{v}_g \right) \quad (4)
\]

The granular temperature equation is given as:
2 \left( \frac{\partial}{\partial t} (\alpha_s \rho_s \Theta_s) + \nabla \cdot (\alpha_s \rho_s \vec{v}_s \Theta_s) \right) \\
= (-p_g \bar{T} + \bar{r}_s) \nabla \vec{v}_s + \nabla \cdot (K_{gs} \nabla \Theta_s) - \gamma_{gs} + \phi_{gs} \tag{5}

The gas and solid phase stress tensors are given by the equation (6) and (7)

**Gas phase**

\[
\bar{r}_g = \alpha_g \mu_g (\nabla \vec{v}_g + \nabla \vec{v}_g^T) - 2 \alpha g \mu_g \left( \frac{\nabla \vec{v}_g}{\rho} \right) \bar{T} = \alpha_g \mu_g \left( \frac{\nabla \vec{v}_g}{\rho} \right) \bar{T} - 2 \alpha_g \mu_g \left( \frac{\nabla \vec{v}_g}{\rho} \right) \bar{T} \tag{6}
\]

**Solid phase**

\[
\bar{r}_s = \alpha_s \mu_s (\nabla \vec{v}_s + \nabla \vec{v}_s^T) - \alpha_s \lambda_s - 2 \alpha_s \mu_s \left( \frac{\nabla \vec{v}_s}{\rho} \right) \bar{T} = \alpha_s \mu_s \left( \frac{\nabla \vec{v}_s}{\rho} \right) \bar{T} - \alpha_s \lambda_s - 2 \alpha_s \mu_s \left( \frac{\nabla \vec{v}_s}{\rho} \right) \bar{T} \tag{7}
\]

The solid properties can be determined as a function of granular temperature according to the following expression:

\[
p_s = \alpha_s \rho_s \Theta_s + 2 \rho_s (1 + e_{ss}) \frac{\alpha_s^2 \bar{g}_{ss} \Theta_s}{\pi} \tag{8}
\]

\[
\lambda_s = \frac{4}{3} \alpha_s \rho_s d_s g_{ss} (1 + e_{ss}) \frac{\Theta_s}{\pi} \frac{1}{2} \tag{9}
\]

\[
\mu_s = \frac{4}{5} \alpha_s^2 \rho_s d_s g_{ss} (1 + e_{ss}) \frac{\Theta_s}{\pi} \frac{1}{2} + \frac{\alpha_s d_s \sqrt{\Theta_s \pi}}{6 (3 + e_{ss})} \left( \frac{1}{2} + \frac{2}{5} (1 + e_{ss}) (3 e_{ss} - 1) \right) \frac{\alpha_s \bar{g}_{ss}}{\alpha_s \bar{g}_{ss}} \tag{10}
\]

Radial distribution function \( g_{ss} \) is calculated by fallingow expression:

\[
g_{ss} = \left[ 1 - \left( \frac{\alpha_s}{\alpha_s \bar{g}_{ss}} \right)^{1/3} \right]^{-1} \tag{11}
\]

**Syamlal–O’Brien drag model**

Interactions between gas and solid phases have a significant effect on the study of hydrodynamic behaviour of fluidized bed. In Euler–Euler modelling the interaction between gas and solid phase is done by the drag force \( K_{gs} (v_s - v_g) \)[13]. In the present investigation Syamlal–O’Brien drag models is used for gas and solid phase interaction for the calculation of drag coefficient, written as:

\[
K_{gs} = \frac{3 \alpha_s \alpha_g \rho_y}{4 v_{rs} d_s} \left( \frac{Re_s}{v_{rs}} \right) \left| \vec{v}_s - \vec{v}_g \right| \tag{12}
\]

Where; \( v_{rs} \) is the terminal velocity correlation for the solid phase

\[
v_{rs} = 0.5 \left( A - 0.06 Re_s \right) + \sqrt{(0.06 Re_s)^2 + 0.12 Re_s (2B - A) + A^2} \tag{13}
\]

Where;

\[
A = \alpha_g^{4.14} \quad B = \begin{cases} 
0.8 \alpha_g^{1.28} & \text{for } \alpha_g \leq 0.85 \\
0.8 \alpha_g^{2.65} & \text{for } \alpha_g > 0.85
\end{cases}
\]

3. Boundary conditions

The Physical model refers to a 2-D Bubbling Fluidized Bed dryer as shown in Figure 1. The width and height of the dryer bed are 0.095 m and 0.40 m respectively. Air is used as fluidizing medium at the bottom of dryer. At the wall, a no-slip boundary condition is set for the air and a Johnson and Jackson, (1987) [16] slip boundary condition is applied for sawdust given as:

\[
v_{s,w} = -A \frac{\partial v_{s,w}}{\partial n} \tag{14}
\]

Here, the slip co-efficient \( A \) can be expressed in terms of specularity co-efficient \( \Phi \) as given below:

\[
A = \frac{6 \mu_s \alpha_s \bar{g}_{ss} \sqrt{\Phi}}{2 \sqrt{3} \pi \rho g \bar{g}_{ss} \sqrt{\theta}} \tag{15}
\]

The bed of dryer is initially filled to a height of 0.15 m with sawdust having mean particle diameter 600 \( \mu \)m and particle density 600 Kg/m\(^3\). The value of specularity co-efficient and initial volume fraction used for the present simulation is \( \Phi=0.55, 0.6 \) and the atmospheric pressure boundary condition is used at the outlet of the fluidized bed dryer. The coefficient of restitution is set to be 0.95 for sawdust-sawdust particles interaction.

4. Solution procedure

In the present investigation above equations are solved using ANSIS software with FLUENT tool by applying appropriate initial and boundary condition as given in details above. For unsteady formulation, 2\(\text{nd} \) order implicit scheme is used and first order upwind, algorithm is used for discretization of convective terms. Phase couple simple, algorithm for multiphase flow is used for the pressure velocity coupling. All the simulations are run for 20 sec and time-averaging has been done between 2 sec and 20 sec to obtain the time-averaged value. A small time step of 1x10\(^{-3}\) sec was used to avoid the instability.
for each simulation. The grid independency is taken according to Gelderbloom et al. (2003) [17]; in their investigation they showed that the bubble size computed with a grid size of about 10 times particle diameter agreed well with the experimental results.

Results and discussion

5.1 Fluidization and bed expansion of sawdust

In this study, sawdust with average particle size \( d_p = 600 \mu m \) was used as the bed material for analyzing its fluidization. Figure 2 qualitatively shows the fluidization behaviour of sawdust, which was initially patched to have a uniform volume.

Figure 2: Variation in (A) contours of volume friction of sawdust, (B) velocity vector of sawdust, (C) contours of volume friction of air; with respect to superficial velocity of air.
fraction of 0.6 and bed height of 0.15 m. As the air superficial velocity increased, corresponding bed expansion also increased. On increasing the air velocity to the above minimum fluidization velocity, there was observed an increase in the average bubble size which led to increase in the movement of the sawdust particles along the bed height. From Figure 2(C) it was also observed that at higher air inlet velocity, sawdust was fluidized with larger bubbles which create problems of slugging and some particles of sawdust moves away from the bed. The variation in bed expansion of sawdust with respect to air superficial velocity from 0.2 to 0.6 m/sec is shown in Figure 3.

![Figure 3: Variation in bed height with respect to superficial velocity of air.](image)

From Figure 2(B) it was clearly observed that during fluidization solid particles moves upward with air in central position of bed and fall back down through the wall. This movement of sawdust particle is due to the bubble movement from the bottom to the top of the bed which leads to distribution of the sawdust particles along the bed height. At air superficial velocity of 0.5 m/sec good fluidization was seen for sawdust with better solid particle movement from bottom to top. In between 0.5 to 0.6 m/sec superficial velocity of air; rapid bed expansion was seen due to larger size bubble formation. At air superficial velocity of 0.2 m/sec, the maximum bed height was found of 0.174 m while air superficial velocity at 0.6 m/sec the value of maximum bed height was 0.342 m. Figure 5 shows the variation in contours of volume friction and bed expansion of sawdust with respect to time at the superficial velocity of 0.5 m/sec for time interval of 0 sec to 20 sec. From Figure 5 it was clearly observed that during initial period of air inlet bed height rapidly increases, after 8 to 10 sec it becomes flatter with small drop in bed height from maximum height reached before this period and material present in bed getting fluidized. This drop in bed height before starting flatter region is due to blasting of larger bubble which is initially formed. Larger bubble formation takes place during initial period is due to cohesive forces of sawdust. The variation in bed expansion with respect to time for the air superficial velocity of 0.2 to 0.5 m/sec is shown in Figure 4. It is clear from Figure 4 that the bed expansion of sawdust particles in the bed for all air superficial velocity of 0.2 to 0.5 m/sec was given almost similar trend with different value of bed expansion.

5.2 Effect of inlet air velocity on mean volume fraction

The lateral distributions of the time-averaged mean volume fraction of sawdust and air are plotted for different values of the air superficial velocity at height of (a) \( H = 0.15 \) m, (b) \( H = 0.1 \) m and (c) \( H = 0.05 \) m from the inlet as shown in Figure 6. The sawdust particle mean volume fraction is greater near-wall region and less in the central region while air distribution was higher in central region of bed for all air superficial velocity. The time-averaged mean volume fraction of sawdust for the different air superficial velocity along height about central axis is plotted in Figure 7. It was observed that an increase in air superficial velocity, the bed height increased and the mean particle volume fraction in the bed decreased.

5.3 Effect of inlet air velocity on mean static pressure of mixture

The effect of different superficial velocity of air along height at mean static pressure of mixture in the bed is shown in Figure 8. It is observed that mean static pressure of mixture found to be higher near the distributor plate and gradually decreases upto zero towards top region of bed. The mean static pressure of mixture along height was also seen
higher at higher superficial velocity of air. The value of mean static pressure of mixture along height, at higher superficial velocity of air decreases in slow manner which means faster fluidization process, while at low superficial velocity of air decreases fast and become approximately zero which means slower fluidization process.

Figure 5: Variation in contours of volume friction of sawdust with respect to time at the superficial velocity of 0.5 m/sec.
Figure 6: Axial variation in mean volume fraction of Sawdust and Air along lateral distance for the different superficial velocity of air at height of (A) H= 0.15 m, (B) H= 0.1 m and (C) H= 0.05 m from the inlet.
5.4 Effect of inlet air velocity on mean velocity magnitude of sawdust

In most gas–solid fluidized bed dryers, especially in the inlet section, particles always accelerate or decelerate unsystematically instead of similar to a steady state because the drag force of the gas on them rarely balances their effective gravity [15]. Figure 9 shows the variation of mean velocity magnitude of sawdust for the five different superficial velocity of air along height. It is seen that for all the value of air superficial velocity, the mean velocity magnitude of sawdust gradually increases up to certain value of bed height and after that it is decreases. As the value of air superficial velocity increases the value of mean velocity magnitude of sawdust is increases and a small peak is also seen in all cases due to bursting of bubble at the top of bed. The region where bubble bursting occurred, pressure suddenly increases which causes increase in mean velocity magnitude of sawdust at the top region of bed.

6. Conclusion

2-D numerical study of two-fluid model carried out, in order to investigate the effects of particle and gas behaviour for fluidized bed dryer. The Eulerian–Eulerian model based on the kinetic theory of granular flow is applied to describe the air–solid turbulent flow. Different sets of simulations have been performed to investigate the fluidization behaviour of sawdust. The effects of inlet air velocity in mean volume fraction, effect of inlet air velocity in mean static pressure of mixture and effect of inlet air velocity in mean velocity magnitude of sawdust, were investigated. The growth of granular flow regimes, variation in contours of volume friction of sawdust, velocity vector of sawdust, contours of volume friction of air; with respect to superficial velocity of air as well as variation in contours of volume friction of sawdust with respect to time at the superficial velocity of 0.5 m/sec were investigated. Better fluidization of sawdust was seen for the air superficial velocity of 0.5 m/sec from bottom to top. The variation in bed expansion was found of increasing nature during initial period and after 8 to 10 sec and it become flatter with small reduction in bed expansion.

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8. Nomenclature

\( d \) – Diameter  
\( g \) – Gravity  
\( g_{0,ss} \) – Radial distribution function  
\( I \) – Unit tensor  
\( k_{gs} \) – Momentum exchange coefficient between gas and solid  
\( p \) – Pressure  
\( t \) – Time  
\( \vec{v} \) – Velocity vector

Greek letters

\( \alpha \) – Volume fraction  
\( \rho \) – Density  
\( \tau \) – Shear stress tensor  
\( \mu \) – Shear viscosity  
\( \lambda \) – Bulk viscosity  
\( \Theta \) – Granular temperature  
\( \phi \) – Specularity coefficient

Subscripts

\( v \) – Gas phase  
\( q \) – Phases (Gas or Sawdust)  
\( s \) – Sawdust (Solid phase)

9. References


