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Model Development for Hydrodynamic Study of Fluidized Bed Gasifier for Biomass Gasification

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Abstract: Fluidized beds are widely employed in industrial operation due to their excellent solid mixing, heat and mass transfer properties. Deeper knowledge of the fluidized bed hydrodynamics would provide the base for development of a fully predictive model. This study highlights the model development for hydrodynamic study and the effect of particle size to the solid fluidization in fluidized bed gasifier using Eulerian-Eulerian multiphase model coupled with kinetic granular theory in CFD software, Ansys Fluent v6.3. The result obtained has been compared with literature data and had proven that the model is capable of accurately model the hydrodynamic of fluidized bed gasifier. Different particle size will give different hydrodynamic flow in the gasifier and particle size in the range of 250-300 μm is observed to give the best solid fluidization behavior in the gasifier. This model can be used further to study the effect of other parameters such as steam inlet velocity and solid initial bed height on the hydrodynamic of the fluidized bed reactor.

Key words: Fluidized bed reactor, Eulerian-Eulerian multiphase model, hydrodynamic, particle size

INTRODUCTION

Gasification technologies are expected to play a key role in expending the use of biomass as a major renewable energy source (Pfeifer and Hofbauer, 2008). Gasification of biomass offers an efficient and economical process to provide gaseous fuels. Biomass is well known for its readily available and exists in diverse forms such as wood, energy crops, agricultural, forest waste products and many others (Foscolo *et al*., 2007). Biomass is characterized as low energy density material thus, practical applications are required in order to turn biomass into gaseous, liquid or solid-derived fuels. The conversion of solid biomass to a gaseous fuel significantly increases its potential. The gas fuels can be used in several applications such as co-firing, electricity generation and many others (Pfeifer and Hofbauer, 2008).

Gasification of biomass converts solid biomass into gaseous product (H_2, CO, CO_2, H_2O, CH_4 and light hydrocarbon), condensable tars, nitrogen compounds (NH_3, HCN) and solid products through reactions with gaseous media such as air, steam or oxygen (Pfeifer and Hofbauer, 2008). It is reported that through biomass gasification, a hydrogen rich product gas with reduced CO and CO_2 concentration can be produced (Mahishi *et al*., 2008; Florin and Harris, 2008; Wu *et al*., 2006; Marquard-Mollenstedt *et al*., 2004; Ko *et al*., 2001). Hydrogen rich gas produced from biomass gasification can be utilized in fuel cell units for electricity production, as hydrogen source for refinery hydrotreating operation, ammonia production and methanol and Fisher-Tropsch synthesis (Rapagna *et al*., 1997).

Fluidized beds are widely employed and most suitable for biomass gasification due to their excellent solid mixing, heat and mass transfer properties (McKendry, 2002). Although it's rather simple in its conception, the application of a fluidized bed process still faces some challenges. This is because the process is strongly influenced by the operating conditions and understanding the mechanism governing the complex flow involved in fluidized bed still remains an open technical and scientific issue (Busciglio *et al*., 2009).

Deeper knowledge of the fluidized bed hydrodynamics would provide the base for development of a fully predictive model. For maximum performance of solid mixing, fluidized bed reactor for biomass gasification is the best to be operated in bubbling flow regime. When the fluidizing gas velocity exceeds the minimum fluidization velocity, u_{mf} which is the minimum velocity

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that is required to fully support the solid particles, particles free voids or bubbles will be formed (Lim *et al*., 2009). The presence and the rising up of bubble in the fluidized bed reactors will ensure that the particles are circulated throughout the bed so that the properties and process conditions could be considered uniform.

The fluidization quality of a bed is highly dependent on the distribution of bubbles and bubble dynamics (Bokkers *et al*., 2004). Ideally, good quality of fluidization must have high population of bubbles, bed should be large but bubbles should be small in size, homogeneously occupy the bed and have low rise velocities. It is appears clearly that studies on bubbles are important for investigating fluidized bed fluid dynamic (Busciglio *et al*., 2009; Lim *et al*., 2009). One of the factors that might affect the bubble characteristics and bubble dynamics in fluidized bed is the particle size (Bokkers *et al*., 2004; Lim *et al*., 1995). Different particle size will have different interparticle stresses that play a key role in bubble wakes especially when bubbling fluidization occurs (Lim *et al*., 1995). Therefore, in this study, the effect of particle size to solid fluidization in the fluidized bed gasifier for biomass gasification is studied.

MATERIALS AND METHODS

Multiphase model: A phase is a class of matter with definable boundary and a particular dynamic response to the surrounding flow or potential field. Phases are generally identified by solid, liquid or gas. Fluidized bed for biomass gasification reaction involves multiphase system consisting solid and gas. The powdered biomass exists in the solid form and the gasifying agent which in this case steam, is in gaseous form. Currently, there are two approaches for the numerical calculation of multiphase flows which are the Eulerian-Lagrange approach and the Eulerian-Eulerian approach. The comparison between these two approaches is as stated in Table 1.

From the comparisons made in Table 1, the Eulerian-Eulerian approach is much more suitable for the CFD simulation of fluidized bed gasifier. This is because the assumptions made in Eulerian-Lagrange approach where the volume fraction of the dispersed second phase is low (<10%) is not true for biomass gasification system in the fluidized bed.

The general idea in formulating multiphase model is to treat each phase as an interpenetrating continuum and therefore to construct integral balances of continuity, momentum and energy for both phases. Since the solid phase has no equation of state and lacks variables such as viscosity and normal stress, appropriate assumptions need to be made in order to obtain a complete momentum balance. Simulations of the bubbling behavior of the fluidized bed were performed by solving equations of motion of a multiphase system. Kinetic theory of granular flow also needs to be applied for the conservation of the solid's fluctuation energy (Papadikis *et al*., 2009).

In this study, the Eulerian-Eulerian Multiphase Flow Model coupled with granular kinetic theory has been used for the simulation of fluidized bed gasifier. For the present case of two-phase flow, the model has to solve several equations related to scalar continuity equation, mass and momentum balance equations. The model which includes all the governing equations was developed in Fluent v6.3 software. The model was used to run several simulations according to parameters similar to Busciglio *et al*. (2009).

Table 1: Comparison between two different numerical approaches for multiphase flows (Bokkers *et al*., 2004)

	Eulerian-Lagrange	Eulerian-Eulerian
Description	- Fluid phase is treated as a continuum by solving the time-averaged Navier-Stokes equations - Dispersed phase is solved by tracking large number of particles, bubbles or droplets through the calculated flow field. The dispersed phase can exchange momentum, mass and energy with the fluid phase	- Different phases are treated mathematically as interpenetrating continua - Multiphase model available: Volume of Fluid model (VOF), mixture model and Eulerian model
Assumptions	- The dispersed second phase occupies low volume fraction (<10%), even though high mass loading is acceptable. - Particle or droplet trajectories are computed individually at specified intervals during the fluid phase calculation.	- Volume of a phase cannot be occupied by the other phases - Volume fractions are assumed to be continuous functions of space and time with their sum is equal to one.
Application	- Spray dryers, coal and liquid-fuel combustion, some particle-laden flow. - In appropriate for liquid-liquid mixture, fluidized beds or any application where the volume fraction of the second phase is not negligible	- VOF: Stratified flow, free-surface flows, filling, sloshing, motion of large bubbles in liquid, steady or transient tracking of any liquid-gas interface - Mixture: particle-laden flows with low loading, bubbly flows, sedimentation and cyclone separators. - Eulerian: bubble columns, risers, particle suspension and fluidized beds

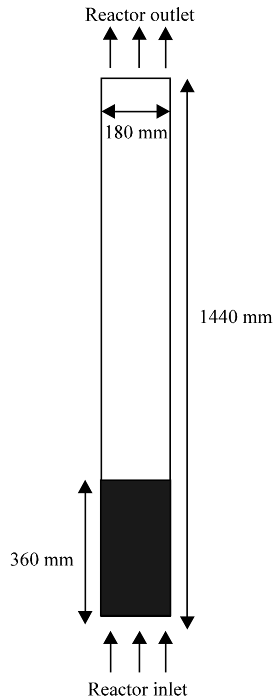

Fig. 1: Domain dimension for model validation (Busciglio *et al*., 2009)

The result from the simulations was then compared with the data from Busciglio *et al*. (2009) who had simulated the hydrodynamics of fluidized bed reactor using Ansys CFX-10.0 and their computational results had been compared with the experimental data. Once proven, the model was then used to study the effect of different particle size to the solid fluidization behavior in a fluidized bed gasifier for biomass gasification.

Initial condition, boundary condition and solution: All calculations are done by Fluent v6.3 on a 2D fluidized bed domain. For model validation, the domain dimension is similar to Busciglio *et al*. (2009) as shown in Fig. 1. Reactor inlet was specified as velocity inlet of air and reactor outlet was specified as pressure outlet which mean only compressible fluid can escape the reactor. The wall of the reactor was specified as no-slip, stationary wall. The simulation parameters are same as Busciglio *et al*. (2009) as shown in Table 2.

After validation, the model was used to simulate the effect of different particle size to the solid fluidization in the fluidized bed gasifier. The 2D fluidized bed domain of the biomass gasifier is shown in Fig. 2. The reactor inlet was specified as velocity inlet of steam and the other boundary conditions remain constant. The simulation parameters for biomass gasifier are as stated in Table 3.

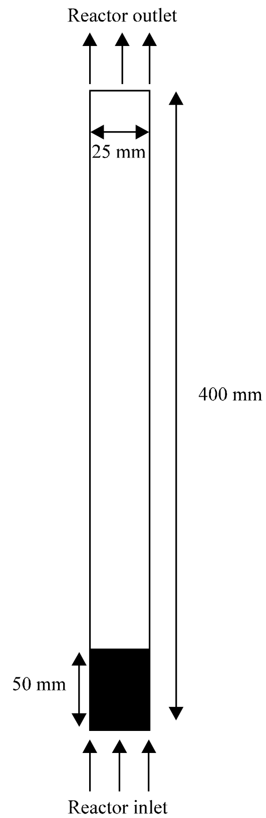

Fig. 2: Domain dimension for biomass gasifier

Table 2: Simulation parameters for model validation (Busciglio *et al*., 2009)

Property	Value	Remarks
Biomass density (ρ_s)	2500 (kg m^{-3})	Carbon (s)
Biomass particle diameter (d_p)	231 (μm)	Fixed
Biomass viscosity (μ_s)	1.72×10^{-5} (kg m^{-1} sec)	Constant
Air density (ρ_g)	1.225 (kg m^{-3})	Air
Air viscosity (μ_g)	1.7894×10^{-5} (kg^{-1} sec)	Constant
Superficial gas velocity	0.0891 (m sec^{-1})	Constant
Restitution coefficient	0.9	Busiglio *et al.* (2009)
Initial solid packing	0.65	Busiglio *et al.* (2009)
Maximum solid packing	0.8	Fixed
Static bed height	360 (mm)	Design required
Bed width	180 (mm)	Design required
Gravitational force	9.81 (m sec^{-2})	Constant

Table 3: Simulation parameters for biomass gasifier

Property	Value	Remarks
Biomass density (ρ_s)	2000 (kg m^{-3})	Carbon(s)
Biomass particle diameter (d_p)	250 (μm)	Fixed
Biomass viscosity (μ_s)	1.72×10^{-5} (kg m^{-1} sec)	Constant
Min fluidization velocity (u_{mf})	0.028 (m sec^{-1})	Constant
Water vapor density (ρ_g)	0.5 kg m^{-3}	Water-Vapor
Water vapor viscosity (μ_g)	1.9×10^{-5} (kg m sec^{-1})	Constant
Restitution coefficient	0.9	Busciglio *et al.* (2009)
Initial solid packing	0.615	Busciglio *et al.* (2009)
Maximum solid packing	0.95	Fixed
Bed width	25 (mm)	Design required
Gravitational force	9.81 (m sec^{-2})	Constant

RESULTS AND DISCUSSION

Model validation: Figure 3 shows the contours for volume fraction of solid inside the fluidized bed gasifier before and after the solid particles being fluidized by air as the gasifying agent. By comparing Fig. 3a and b, it is observed that the bubble diameter simulated by Fluent is much bigger compared to the bubbles computed by Busciglio *et al*. (2009). This is because fluctuation energy of solids particles or the granular temperature of solid is calculated by neglecting the loss of energy due to convection and diffusion. As the particles have higher energy than in actual case, it moves more rapidly when gasified by air. This rapid movement cause the smaller bubbles to coalesce more rapidly and forming larger bubbles.

Even though the bubbles produced is much bigger, the bubble distribution inside the fluidized bed gasifier does not differ much with the one computed by Busciglio *et al*. (2009). Figure 3a bubbles formation was reached after 2 sec of fluidization. This observation is the same as Busciglio *et al*. (2009). From the comparison of Fig. 3a and b, it appears that the code is able to correctly simulate the bubble formation inside the fluidized bed reactor. However, the bubbles simulated in the code are slightly bigger than literature study which might affect the average bed height of the system. It is predicted that bed height simulated by the code should be slightly higher due to the bigger bubble eruption on the surface of the expended bed.

Figure 4 shows the comparison of average bed height computed by the code with other data obtained from literature study Busciglio *et al*. (2009). From the graph in Fig. 4, it is observed that the difference between average bed heights computed by the code in Fluent are insignificant compared to the data obtained from literature. As expected, the average bed height computed by Fluent is slightly higher compared to the experimental and CFX data. The overall difference is less than 5%.

Figure 5 shows the instantaneous bed height computed by Fluent compared with instantaneous bed height from Busciglio *et al.* (2009) for air velocity, U, equals to $3.4U_{mf}$. From the comparison, it is observed that the instantaneous bed height computed by Fluent is slightly higher compared to literature data. It is expected that the bubbles formation in Fluent is larger than bubbles formed in experimental set up and computational in CFX. Thus, bigger bubble eruption at the surface of extended bed causes the bed height to be higher. Overall, the average difference between the computed instantaneous bed height in Fluent and the data from literature is less than 10% and it shows that the model can be used further to study the effect of other parameters to the solid fluidization in the fluidized bed gasifier.

Effect of Particle Size to Solid Fluidization in the Gasifier: Figure 6 shows the instantaneous expended bed height for different solid particle size. From this Fig. 5, it is observed that 100 μm particles give the highest

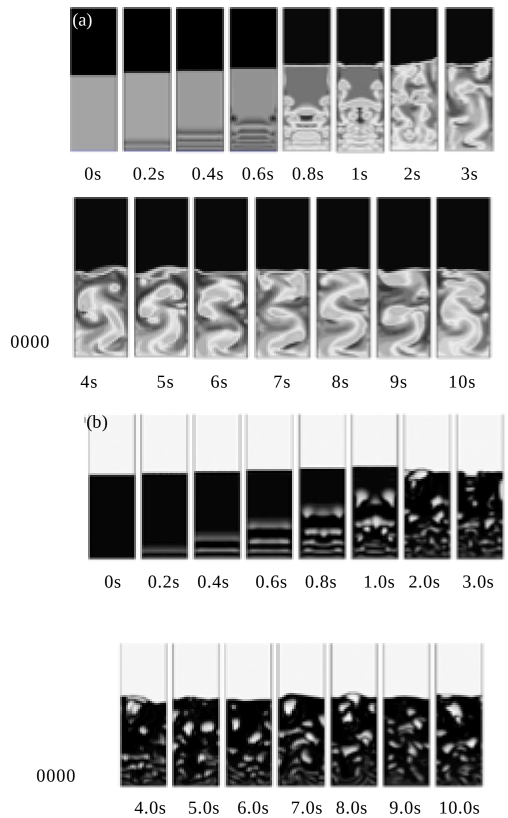

Fig. 3: Contours for solid volume fraction at U=1.7 U_{mf}; (a) Result obtained from simulation and (b) Result from Busciglio *et al*. (2009)

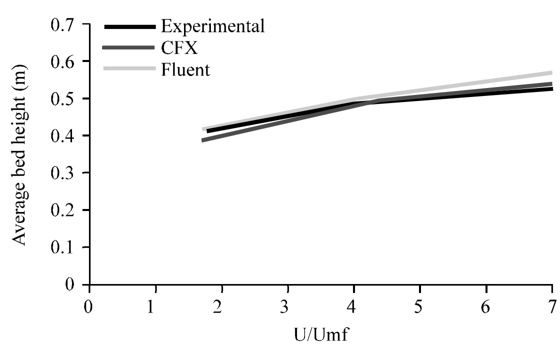

Fig. 4: Comparison of simulation result from Fluent with other result from literature study of Busciglio *et al*. (2009)

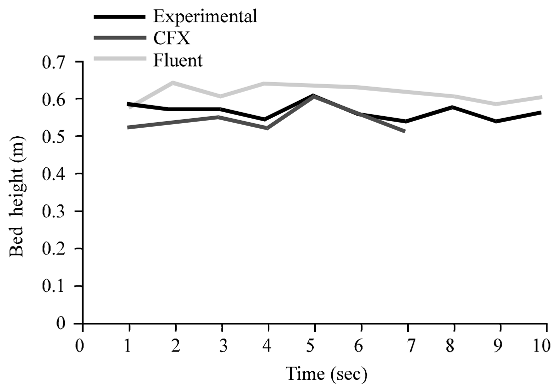

Fig. 5: Comparison of instantaneous bed height of 3.4 U_{mf} obtained from simulation with other result from literature study (Busciglio *et al*., 2009)

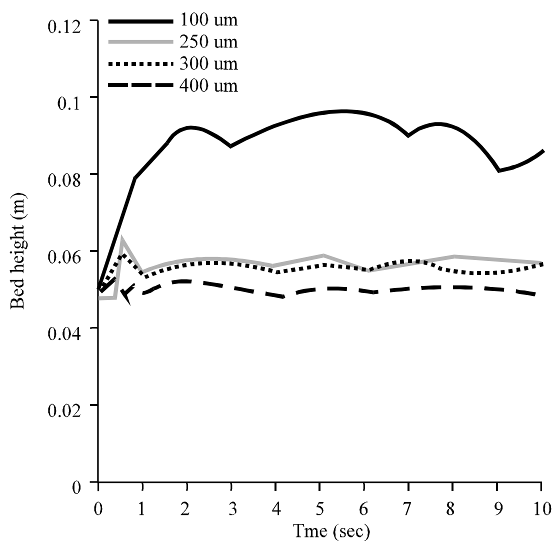

Fig. 6: Expended bed height with respect to different particle size

expended bed height. This is because small particles are lighter and easier to be fluidized by the steam. However, high expended bed shows that the voidage between the solid particles are too big which is undesirable especially for biomass gasification reaction because the interaction between the solid particles is almost none.

Four hundred micometer particles give the lowest expended bed height because these big particles are heavy and difficult to be fluidized by the steam. Low expended bed also shows that the solid packing is too dense and not very suitable for gasification reaction to occur. Solid diameter in the range of 250 to 300 μm gives almost the same expended bed height. This shows that at this range of diameter, the solid fluidization is almost the same as it is being gasified by the steam. Furthermore, the expended beds for both particle diameters are stable with respect to time. Therefore, range of particle diameter of 250 to 300 μm seems to be suitable for biomass gasification.

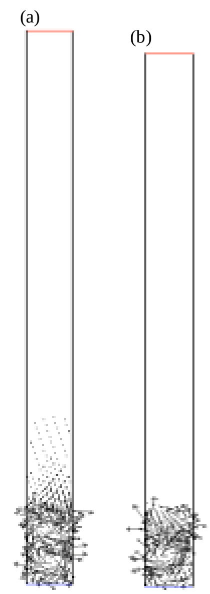

Fig. 7: Vector of solid particles in fluidized bed gasifier (a) 250 μm particle and (b) 300 μm particle

Figure 7 shows vectors of solid particles in the fluidized bed gasifier for 250 μm particle and 300 μm particle. This vector shows the movement of the solid particle inside the bed. From the vector, it is observed that the solid particles are circulated well in the gasifier without any solid particle flows out of the reactor. It is observed that 250 μm particle float higher than 300 μm particle. This is because 250 μm is much lighter than 300 μm particle and tend to float together with the steam during gasification. However, when the solid particle reaches a certain height, at which it loses its momentum energy, the particle will fall back to the bed due to the effect of gravitational force.

Further test was also done using bigger particle diameter in order to observe its effect to the solid fluidization in the gasifier. However, at particle diameter greater than 500 μm, solid segregation occurs in the gasifier as shown in Fig. 8. Solid segregation is a condition where some particles tend to move upwards and float outside of the reactor while the others travel in the opposite direction (Lim *et al*.,1995). This condition shows poor mixing in the fluidized bed gasifier and not suitable for reaction to occur. From the simulation, it is observed that suitable range of particle diameter for biomass gasification is from 250 to 300 μm..

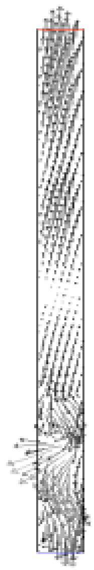

Fig. 8: Particle segregation in the gasifier at particle diameter greater than 500 μm

CONCLUSION

As the conclusion, Eulerian-Eulerian Multiphase Flow Model coupled with granular kinetic theory that has been used in Fluent v6.3 is capable to accurately model the hydrodynamics of the fluidized bed gasifier. This is proven by the comparison with CFX simulation and experimental data that gives almost the same bubble distribution and the percentage difference is less than 5% for average bed height and less than 10% for instantaneous bed height. The model was then used in order to study the effect of particle diameter to the solid fluidization in the gasifier and it is observed that the suitable range of particle diameter for biomass gasification is within 250 to 300 μm.

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