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Effect of Steam Inlet Velocity and Solid Initial Bed Height to the Hydrodynamics of Fluidized Bed Gasifier

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Abstract: Fluidized bed reactor is commonly utilized in industry due to their excellent solid mixing, heat and mass transfer properties. Fluidized bed reactor is best to be operated in bubbling flow regime. This study highlights the model development for hydrodynamics study on the effect of steam inlet velocity and solid initial bed height to the solid fluidization in the fluidized bed gasifier using Eulerian-Eulerian multiphase model coupled with kinetic theory granular flow approach. From the simulation, it is observed that 3.5 U_{mf} steam inlet velocity gives the best solid fluidization in the gasifier. The increase in solid initial bed height does not give significant improvement to the fluidization in the gasifier. However, at high initial bed height the expended bed become unsteady. Therefore, for this work solid initial bed height of 10 cm or below is much more suitable as it gives good fluidization in the gasifier while still maintaining the stability of the extended bed.

Key words: Computational fluid dynamic, fluidized bed gasifier, hydrodynamics study, steam inlet velocity, solid initial bed height

INTRODUCTION

It is widely known that combustion of fossil fuels contributes to the build-up of Carbon Dioxide (CO_2) in the atmosphere which contributes to the global warming issue all around the world. In 2005, it is reported that the concentration of CO_2 was 379 ppm, approximately 180-300 ppm more than the equilibrium concentration for the last 650 000 years (Florin and Harris, 2008). Hence, the search towards cleaner and renewable alternative energy is attracting more attention.

Hydrogen is one of the potential alternative energy that could be used to replace the existing fossil fuels. Hydrogen is expected to become a prominent energy carrier for stationary and mobile power generation applications such as in transport, industrial, commercial and residential applications (Florin and Harris, 2008; Mahishi *et al*., 2008; Wu *et al*., 2006). Clean and renewable hydrogen can be produced through the biomass gasification process. Biomass gasification is the process of converting solid biomass into gaseous fuel mainly hydrogen (H_2), carbon dioxide (CO_2), carbon monoxide (CO) and methane (CH_4) (Foscolo *et al*., 2007; Pfeifer and Hofbauer, 2008; Rapagna *et al*., 1997) by heating it in a gasification medium such as air, oxygen or steam (McKendry, 2002).

Biomass gasification process usually took place in various types of fluidized bed reactor due to its excellent solid mixing, heat and mass transfer properties (Mahishi *et al*., 2008; Wu *et al*., 2006; Rapagna *et al*., 1997; Marquard-Mollenstedt *et al*., 2004). For maximum performance of solid mixing, fluidized bed reactor for biomass gasification is best to be operated in bubbling flow regime (Foscolo *et al*., 2007). The presence of bubble in the fluidized bed reactors will ensure that the particles are circulated throughout the bed so that the properties and process condition could be considered uniform. 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 (Busciglio *et al*., 2009; Lim *et al*., 1995, 2009). Deeper knowledge of the fluidized bed hydrodynamics would provide the base for development of a fully predictive model.

Fluid flow, heat and mass transfer, chemical reaction, solid and fluid interaction and other related phenomena that might involve in the fluidized bed can be model and simulate using Computational Fluid Dynamic (CFD). Computational fluid dynamic is a design and analysis tool that uses computers to simulate momentum, mass and energy transfer equation into virtual fluidized bed reactor model. CFD modeling technique is becoming widespread

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in the biomass thermochemical conversion equipment such as fluidized bed, fixed bed, combustion furnace and many others (Wang and Yan, 2008).

The recent development of mathematical modeling of particulate solids behavior together with increasing computation power enables researcher to simulate the behavior of fluidized powdered biomass and to link fundamental particle properties directly to the powder behavior and predict the interaction between particles and gaseous or liquid fluids. In this case, CFD modeling provides a fundamental tool to support engineering design and research in multiphase system (Busciglio *et al*., 2009).

This study highlights the model development for hydrodynamics study on the effect of steam inlet velocity and solid initial bed height on the solid fluidization in the fluidized bed gasifier using Eulerian-Eulerian multiphase model coupled with kinetic theory of granular flow approach (Bokkers *et al*., 2004; Papadikis *et al*., 2009). This model was validated using experimental and simulation data from Busciglio *et al*. (2009) and Tamidi *et al*. (2009). The validated model was used to obtain the most optimum condition for fluidized bed gasifier for biomass gasification.

APPROACH AND METHOD

CFD is the science of predicting fluid flow, heat transfer, chemical reaction and other related phenomena by solving numerical set of Navier-Stokes equations (Wang and Yan, 2008). The results of CFD analysis are relevant for conceptual studies of new design, detail product development, troubleshooting and redesign. Besides, CFD modeling is also cost saving, timely, safe and easy to scale-up (Wang and Yan, 2008). CFD analysis complements testing and experimentation because CFD can reduce the total effort required in the experiment design and data acquisition (Lim *et al*., 1995). CFD codes turn computers into a virtual laboratory and perform equivalent numerical experiment conveniently providing insight, foresight and return on investment.

Various numerical techniques have been employed in the solution of the CFD model equation and the most widely use numerical technique is discretization method including finite difference, finite element and finite volumes method. Finite volume is now the most commonly approached used in CFD code for its ease in the understanding, programming and versatility. The most routinely used commercial codes include ANSYS Fluent, ANSYS CFX, CFD2000 and many others (Busciglio *et al*., 2009; Wang and Yan, 2008). For this project, ANSYS Fluent v.6.3 was used for simulation of biomass gasification in fluidized bed reactor.

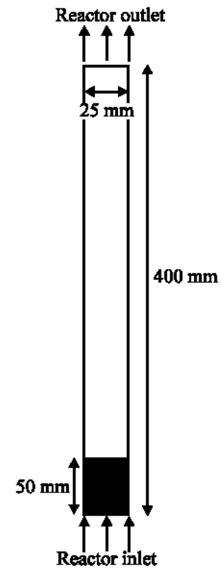


Fig. 1: Domain dimension for biomass gasifier

Table 1: Simulation parameters

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

Working domain: Figure 1 shows the dimension of the working domain in the fluent. It is the representation of a 2D fluidized bed gasifier or reactor. Reactor inlet was specified as velocity inlet of steam 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 stationary wall.

Simulation parameters: Table 1 shows the simulation parameters that are incorporated in the simulation during solver setting. Water vapor or steam is the primary phase in the system. Steam enters the reactor through the reactor inlet at a specified velocity. The secondary phase in the system is the carbon solid particle that is chosen to represent biomass in the reactor with modified parameter to suit the properties of biomass.

The properties of the solid phase are likely to have the sand like behavior of Geldart B Group. There is also gravitational force of 9.81 m sec^{-2} is acting in downward direction of the gasifier.

Table 2: Value of variable

Parameter	Value	Unit
Steam inlet velocity	0.084 ($3U_{mf}$)	m sec^{-1}
	0.098 ($3.5U_{mf}$)	
	0.112 ($4U_{mf}$)	
	0.140 ($5U_{mf}$)	
Solid initial bed height	50 (2d)	mm
	75 (3d)	
	100 (4d)	
	125 (5d)	
	Control variable	
Initial bed height	50	mm
Bed width	25	mm
Inlet velocity	0.084	m sec^{-1}

Variables: The main objective of this simulation is to study the solid fluidization in the fluidized bed gasifier with respect to the change in inlet velocity of steam and solid initial bed height. The variables value of each parameter is shown in Table 2. The control variables are the initial condition of the system.

Mathematical model: 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 balance equation, mass and momentum balance equations.

Volume fraction: For Eulerian-Eulerian model, the volume fractions are assumed to be continuous functions of space and time and their sum is equal to one since the volume of one phase can never be occupied by the other phase as shown below (http://www.scribd.com/doc/20480734/Fluent-Training):

$$\varepsilon_g + \varepsilon_s = 1 \tag{1}$$

Continuity equation: The continuity equation for gas and solid phases are given by (http://www.scribd.com/doc/20480734/Fluent-Training):

$$\frac{\partial}{\partial t}(\varepsilon_g \rho_g) + 1.(\varepsilon_g \rho_g \vec{v_s}) = 0 \tag{2}$$

$$\frac{\partial}{\partial t}(\varepsilon_s \rho_s) + 1.(\varepsilon_s \rho_s \vec{v_s}) = 0 \tag{3}$$

Momentum equation: The momentum balance equations for each phase are derived based on the assumption of there are no mass transfer between the two phases and no lift force, external body force and virtual mass force acting on the secondary phase of the system. The momentum balance equations for each phase are as follow (http://www.scribd.com/doc/20480734/Fluent-Training):

$$\frac{\partial}{\partial t}(\varepsilon_g \rho_g \vec{v_g}) + 1.(\varepsilon_g \rho_g \vec{v_g}\vec{v_g}) = 1.\overline{\overline{S_g}} + \varepsilon_g \rho_g \vec{g} - I_g \tag{4}$$

$$\frac{\partial}{\partial t}(\varepsilon_s \rho_s \vec{v_s}) + 1.(\varepsilon_g \rho_g \vec{v_s}\vec{v_s}) = 1.\overline{\overline{S_g}} + \varepsilon_g \rho_g \vec{g} - \vec{I_g} \tag{5}$$

The interphase momentum change is further defined by the granular kinetic theory in order to estimate the rheological properties for the solid phase. Standard drag models are also employed to estimate the momentum exchange between phases at the particle boundaries.

The fluid-solid momentum exchange coefficient in the interphase momentum change equation for dense fluidized bed can be further described by Gidaspow *et al.* (1992) drag function as follow:

For $\varepsilon_g > 0.8$:

$$F_g = \frac{3}{4}\frac{\varepsilon_s \varepsilon_g \rho_g}{d_p} C_{ds} |\vec{v_s} - \vec{v_g}| \varepsilon_s^{-2.63} \tag{6}$$

For $\varepsilon_g \leq 0.8$:

$$F_g = 150\frac{\varepsilon_s(1-\varepsilon_g)\mu_g}{\varepsilon_s d_p^2} + 1.75\frac{\rho_g \varepsilon_g |\vec{v_s} - \vec{v_g}|}{d_p} \tag{7}$$

Where:

$$C_{ds} \frac{24}{\varepsilon_s}\left[1 + 0.15(\varepsilon_g \, Re_s)^{0.687}\right] \tag{8}$$

$$Re_s = \frac{\rho_g d_p |\vec{v_s} - \vec{v_g}|}{\mu_g} \tag{9}$$

For granular flows in the compressible regime where the solid volume fraction is less than its maximum allowed value, a solid pressure is calculated independently and used for the pressure gradient term in the granular-phase momentum equation. Because a Maxwellian velocity distribution is used for the particles, a granular temperature is introduced into the model and appears in the expression for the solid pressure and viscosities (Lun *et al.*, 1984). The solid pressure is composed of a kinetic term and particle collisions term:

$$P_s = \varepsilon_s \rho_s \theta_s (1 + 2g_0 \varepsilon_s (1 + e_s)) \tag{10}$$

Radial distribution function is a correction factor that modifies the probability of collisions between grains when solid granular phase becomes dense. For one solid phase, the equation for radial distribution function is as follow (http://www.scribd.com/doc/20480734/Fluent-Training):

$$g_0 = \left[1 - \left(\frac{\varepsilon_s}{\varepsilon_s^{max}\square}\right)^{\frac{1}{3}}\right]^{-1} \tag{11}$$

Solid stress tensor contains shear and bulk viscosities arising from particle momentum exchange due to translation and collision. In this simulation, frictional component of viscosity is assumed negligible. The solid shear stress equation with collisional and (Gidaspow *et al.*, 1992) kinetic viscosity is as follow:

$$\mu_s = \frac{4}{5}\varepsilon_g\rho_g d_p g_0(1+e_s)\left(\frac{\theta_s}{\pi}\right)^{0.5} + \frac{10\rho_s d_p\sqrt{\theta_s\pi}}{96(1+e_s)\varepsilon_s g_0}\left[1+\frac{4}{5}g_0\varepsilon_s(1+e_s)\right]^2 \tag{12}$$

Solid bulk viscosity accounts for the resistance of the granular particles to compression and expansion. In Fluent, the solid bulk viscosity has the following form from Lun *et al.* (1984).

$$\lambda_s = \frac{4}{3}\varepsilon_s\rho_s d_p g_0(1+e_s)\left(\frac{\theta_s}{\pi}\right)^{0.5} \tag{13}$$

Kinetic Theory of Granular Flow (KTGF): The transport equation derived from kinetic theory (http://www.scribd.com/doc/ 20480734/Fluent-Training):

The diffusion coefficient is further described by Gidaspow *et al.* (1992) as follow:

$$k_{\theta s} = \frac{150\rho_s d_p\sqrt{\theta\pi}}{384(1+e_s)}\left[1+\frac{6}{5}\varepsilon_s g_0(1+e_s)\right]^2 + 2\rho_s\varepsilon_s^2 d_p g_0(1+e_s)\sqrt{\frac{\theta_s}{\pi}} \tag{14}$$

The collisional dissipation energy represent the rate of energy dissipate within the solid phase due to collisions between particles. This term is represented by the equation derived by Lun *et al.* (1984):

$$\gamma_{\theta s} = \frac{12(1-e_s^2)\rho_s g_0}{d_p\sqrt{\pi}}\varepsilon_s^2\theta_s^{\frac{3}{2}} \tag{15}$$

RESULTS AND DISCUSSION

Effect of steam inlet velocity to solid fluidization in the gasifier: Figure 2 shows the instantaneous expended bed height at different steam inlet velocity. From this graph, it is observed that at lowest inlet velocity which is at 3 U_{mf}, the expended bed height is the lowest and the expended bed height increase as the steam inlet velocity increase. However, at high steam inlet velocity (5 U_{mf}), the expended bed height is not stable and keeps changing with respect to time. This is because at higher inlet velocity, bigger bubbles are formed and the eruption of this bigger bubble affects the stability of expended bed height. The most stable expended bed height is observed at steam inlet velocity of 3.5 U_{mf}.

The fluidization of solid particles by steam at inlet velocity of 3.5 U_{mf} is shown in Fig. 3a. From this figure, it revealed that the solid fluidization is good at the bottom of the bed as the solid volume fraction is about 0.438-0.613 (green and yellow color). However, at higher bed height, the solid phase become denser and solid volume fraction become higher. This is probably because at higher bed height the steam had not enough energy to support and fluidized the solid particles that have higher potential energy due to elevation and at this point, the gravitational effect is much dominant compared to the lifting energy by the steam.

Figure 3b shows the flow regimes inside the fluidized bed reactor for steam inlet velocity of 5 U_{mf}. As the steam inlet velocity increase, the steam has more energy to fluidize the solid particle and the flow regime become more vigorous. However, the bubbles formed in the bed is too big and the void region increase in the bed as the volume fraction of solid is low as 0.35-0.438 (green color). This condition is not favorable especially for gasification reaction to occur inside the reactor because the solid particles and the steam do not mix very well.

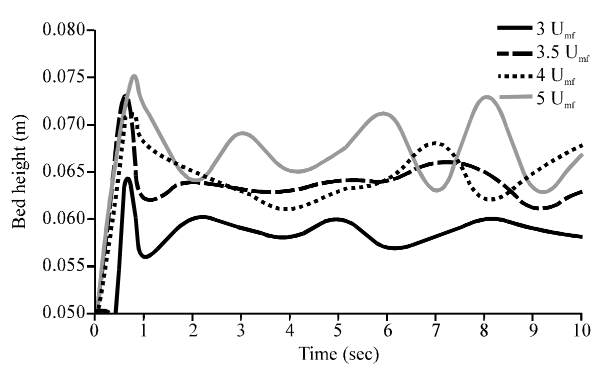


Fig. 2: Expended bed height with respect to different inlet velocity

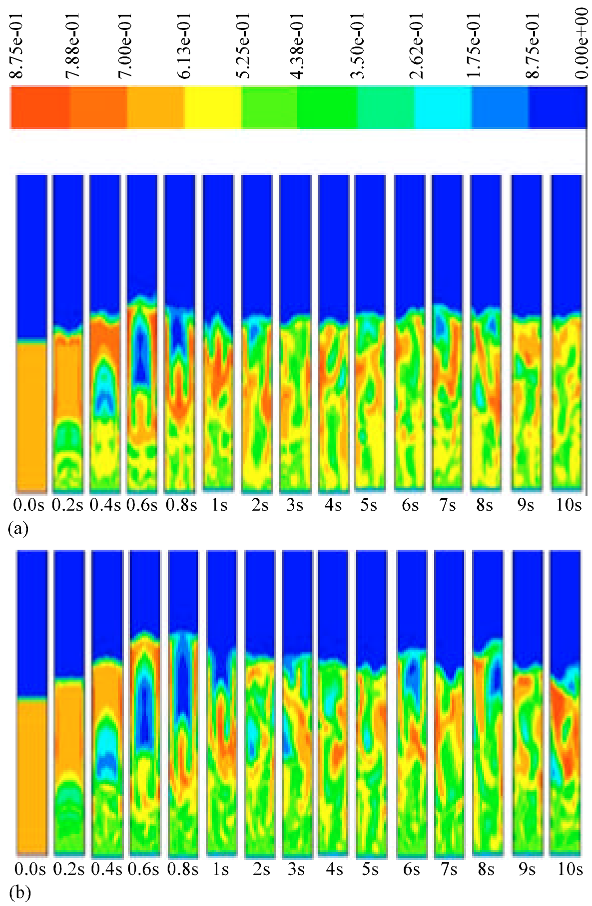


Fig. 3: Contours of solid volume fraction (a) $u_g = 3.5\ U_{mf}$, (b) $u_g = 5\ U_{mf}$. s: sec

Overall, as the steam inlet velocity increase, the steam will have more energy to fluidize the solid particles. However, increasing the steam inlet velocity will increase the void region in the bed, which is in agreement with literature (Lim *et al*., 2009) and at the same time increases the steam operating cost. Thus, operating steam at lower inlet velocity is much favorable. Therefore, the inlet velocity of 3.5 U_{mf} is the optimum steam inlet velocity as it gives the best solid fluidization in the reactor with average solid volume fraction of 0.438-0.613 throughout the bed which is comparable with literature that gives the optimum inlet velocity of 4 U_{mf} (Lim *et al*., 2009). This steam velocity is sufficient in order to fluidize the solid particles in the fluidized bed reactor at bed height of 5 cm.

Effect of solid initial bed height to solid fluidization in the gasifier: Figure 4 shows the instantaneous expended bed height for solid particles that has been gasified by steam at different initial bed height. It is observed that, the higher the initial bed height, the higher the instantaneous

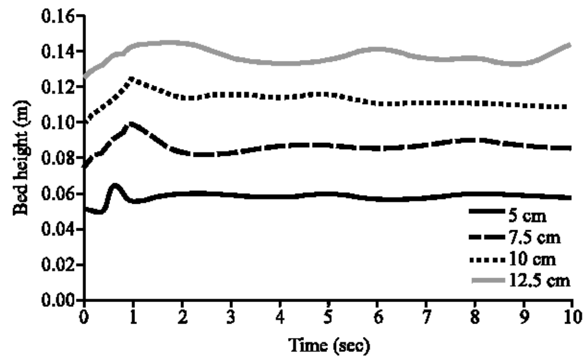


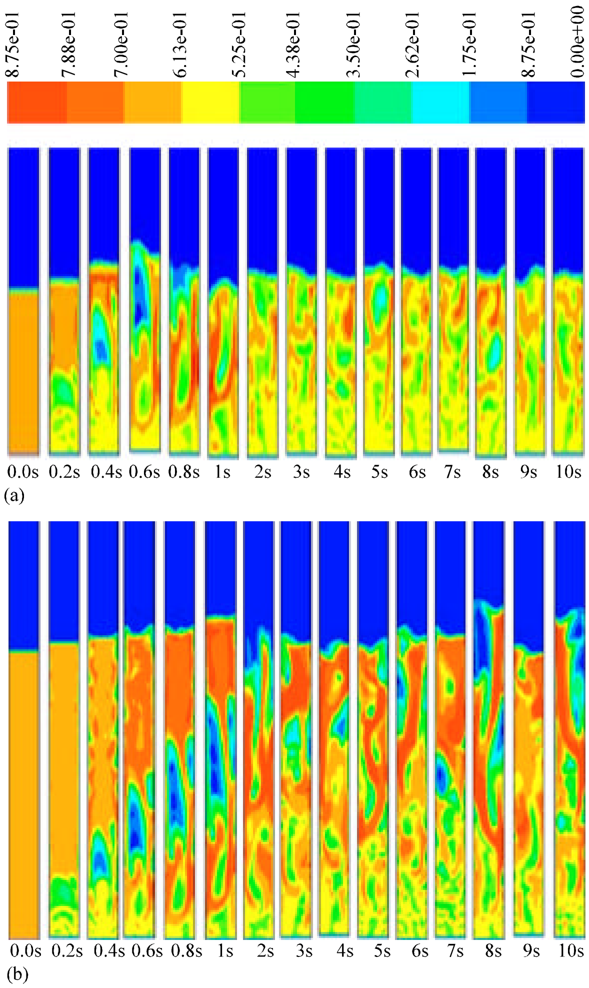
Fig. 4: Expended bed height with respect to different solid initial bed height

expended bed height in the fluidized bed reactor. All expended bed heights are almost stabilized except for 12.5 cm initial bed height. The expended bed for 12.5 cm initial bed height keeps fluctuating with time as the result of bigger bubble eruption at the surface of the bed. From the graph, it is observed that 5 cm initial bed height gives the most stabilized extended bed height which is in good agreement with literature (Busciglio *et al*., 2009; Lim *et al*., 2009; Papadikis *et al*., 2009).

Figure 5a shows the flow regime of gasified solid particles at initial bed height of 5 cm. For the first 0.8 sec, the flow developed from jetting flow regime (0.2 sec) to bubbling flow regime (0.4 sec) and to slugging flow regime (0.6 sec). As the slug reaches the surface of the extended bed height, it erupted and the solid particles flow downward due to gravitational effect (0.8 sec). Stabilized bubble flow is observed after 2 sec with small and homogeneously dispersed bubbles form at the bottom of the bed and the bubbles is getting bigger as it reaches the surface of the bed.

Figure 5b shows the contours of solid volume fraction for solid particles at initial bed height of 12.5 cm. The flow development from 0.2 sec until 0.8 sec is the same as previous ones which is from jetting flow, to bubbling flow and to slugging flow. However at higher solid bed height, bigger bubbles tend to form through the bed. These big bubbles turn into slug as it moves towards the surface of the bed. Formation of big bubbles and slug shows not a very good fluidization condition as the bubbles are not homogeneously dispersed and the condition could not be considered as uniform. This observation is found to be similar with other literature (Busciglio *et al*., 2009; Lim *et al*., 1995).

Overall, the increase of solid initial bed height does not significantly improve the solid fluidization in the gasifier at constant steam inlet velocity. At high solid initial bed height, which is 12.5 cm, more fluctuations in



Fg. 5: Contours of solid volume fraction for solid initial bed height of (a) 5 cm, (b) 12.5 cm. s: sec

the expended bed and formation of slug are observed. This is because solid at higher initial bed height, is much more difficult to be fluidized by the steam. The bubbles form in the bed also tends to be bigger due to coalescing with other bubbles in the bed, which will increase the voidage between the solid particles. The result obtained from simulation shows that initial bed height of 10 cm or lower is much suitable for the gasifier.

CONCLUSIONS

As a conclusion, the difference in steam inlet velocity and solid initial bed height shows significant changes to the hydrodynamics of the fluidized bed reactor. As the steam inlet velocity increases, the steam will have more energy to fluidize the solid particles. However, increasing the steam inlet velocity will increase the voidage in the bed and at the same time increase the steam operating cost. Therefore, the inlet velocity of $3.5U_{mf}$ is the most optimum steam inlet velocity as it gives the best solid fluidization in the gasifier with average solid volume fraction of 0.477-0.513 throughout the bed. Solid initial bed height also may affect the hydrodynamic inside a fluidized bed reactor. With the range used in this work, the increase in solid initial bed height does not give significant improvement on the solid fluidization in the bed. However, there is a trend where the expended bed height tends to fluctuate more at very high initial bed height, which is 12.5 cm. From the simulation, it is observed that solid initial bed height of 10 cm or lower gives the best solid fluidization in the gasifier.

The hydrodynamics model developed is capable to predict the solid fluidization behavior in the fluidized bed. Thus, this provides the insight to the researcher on what is actually happening inside the gasifier during fluidization that is difficult to be observed from experimental setup. The results obtained from this model also can be utilized in various applications such as in design of gasifier, optimization of reaction condition and many others.

NOMENCLATURE

I_g : Interphase momentum change
$\bar{\bar{S}}_g$: Stress-strain tensor
C_d : Drag coefficient
d_p : Particle diameter (m)
d : Column diameter (mm)
e_s : Coefficient of restitution for particle collision
$\vec{g}$: Gravitational force (m sec^2)
g_0 : Radial distribution function
ke_i : Diffusion coefficient
$\bar{v}$: Velocity (m s^{-1})
F : Interphase momentum exchange coefficient
P : Pressure (Pa)
Re : Reynolds number
U_{mf} : Minimum fluidization velocity (m sec^{-1})
u : Superficial velocity (m sec^{-1})

Greek symbols

$\gamma_{\theta i}$: Collisional dissipation energy
ε_i^{max} : Maximum packing limit of species i
θ_s : Granular temperature
μ_g : Shear viscosity
ϕ_{gs} : Energy exchange between gas and solid phase
ε : Volume fraction
ρ : Density (kg m^{-3})

Subscripts

g : Gas phase
s : Solid phase

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