Numerical study of mixing and heat transfer of SRF particles in a bubbling fluidized bed

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Abstract

In this article, numerical investigations on mixing and heat transfer of solid refused fuel (SRF) particles in a bubbling fluidized bed are carried out. The numerical model is based on the Eulerian–Eulerian approach with empirical submodels representing gas–solid and solid–solid interactions. The model is verified by experimental data from the literature. The experimental data include SRF vertical distribution in SRF–sand mixtures of different sand particle sizes ($d_{pm} = 654, 810$ and $1110 \mu\text{m}$) at different fluidization velocities ($u/u_{mf} = 1.2\text{--}2.0$). We proposed magnification of drag force exerted by the gas on SRF particles based on Haider and Levenspiel (Powder Technol 58(1):63–70, 1989) drag coefficient. The proposed model shows good agreement with the experimental data at high fluidization velocities ($u/u_{mf} = 1.5\text{--}2.0$) and poor predictions at low fluidization velocities ($u/u_{mf} = 1.2\text{--}1.5$). Heat transfer results showed that the present model is valid and gives good agreement with the experimental data of wall–bed heat transfer coefficient.

Keywords Mixing · Heat transfer · SRF · Fluidized bed · Fluidization velocity

List of symbols

A, B, C, D	Coefficients (–)
C_D	Drag coefficient (–)
C_p	Particle specific heat ($\text{J kg}^{-1} \text{K}^{-1}$)
d_p	Particle diameter (μm)
ε	Void fraction (–)
g	Gravity acceleration (ms^{-2})
h	Heat transfer coefficient ($\text{Wm}^{-2} \text{K}^{-1}$)
K_{gs}	Momentum exchange coefficient ($\text{kgm}^{-3} \text{s}^{-1}$)
λ	Thermal conductivity ($\text{Wm}^{-1} \text{K}^{-1}$)
μ	Viscosity (Pas)
Nu	Nusselt number $= \frac{hd_p}{\mu} (-)$
Pr	Prandtl number $= \frac{\mu c_p}{\lambda} (-)$
ρ	Density (kgm^{-3})
Re	Reynolds number $= \frac{\rho_g d_p u_g - u_p }{\mu_g} (-)$
T	Temperature ($^{\circ}\text{C}$)

u	Velocity (ms^{-1})
ψ	Particle shape factor (–)

Subscripts and superscripts

D	Drag
g	Gas
m	Mean
mf	Minimum fluidization
p	Particle
s	Solid

Introduction

Fluidized beds offer excellent mixing and heat transfer characteristics which make them efficient in thermal conversion applications, i.e., combustion/gasification of low-grade coal, biomass, and solid refused fuel (SRF) [38]. Practically, fuel particles are of irregular shapes and they enter the fluidized bed at a size range of 5–10 mm [27]. The overall fuel concentration within the fluidized bed combustor is very low compared to the bed material (2–5 mass%) [27]; however, the big differences in the density and size ratios can significantly affect the homogeneity of the bed (segregation phenomenon) [32]. In general, the segregation occurs during fluidization of binary beds containing two or more solid components of different sizes and/or densities. The heavier

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particles (jetsam) tend to settle down to the bottom of the bed, while the lighter particles (floats) float to the upper layers of the bed [26]. This heterogeneous phenomenon can have significant consequences on other processes such as heat transfer, mass transfer, and chemical reactions [30, 33]. Thus, many studies have been concerned with the segregation phenomenon of the binary fluidized bed in both the bubbling [6, 9, 10, 19, 29] and the circulating fluidized beds [28, 49, 51]. Most of the binary fluidized bed studies were concentrating on moderate particle sizes of Geldart B class [11], while few of them studied large particles of Geldart D class [11].

The binary mixture Geldart B-D which is relevant to the fluidized bed combustors did not get enough interest. In the meantime, the particles of the bed material and the fuel material are of non-spherical, irregular shape in addition to the heterogeneous physical properties of some fuel materials such as solid refused fuel (SRF) and municipal solid waste (MSW). A considerable number of experimental and theoretical studies were performed to investigate hydrodynamics and combustion/gasification of SRF/MSW [4, 5, 7, 20, 22–25, 31, 36, 39, 43–45, 47], but few studies have been concerned with simulation of this complex binary mixtures (irregular shape and large size) particles in fuel-sand bubbling fluidized beds [41, 42]. Although, some trials have been performed to simulate the non-spherical particles [15, 18, 46, 50], in fact, the shape of SRF particles cannot be treated as regular particles (see Fig. 1).

The objective of the present study is to shed the light on mixing and heat transfer of this heterogeneous binary mixture of SRF–sand. The Eulerian–Eulerian multi-fluid model (MFM) is used to simulate the segregation of the fuel particles (SRF) at different fluidization velocities. Three gas–solid drag models of Gidaspow [13], Syamlal–O’Brien [40], and Gibilaro et al. [12] are compared. Because of the heterogeneity of SRF particles, the numerical simulations are performed on representative particles of average density and sizes, but with taking into account effect of irregular-shaped large particles (sphericity ratio) on gas–solid

Figure 1 A photograph showing heterogeneity of solid refused fuel (SRF)

momentum interaction. The heat transfer between gas and solid phases is simulated using the Gunn [16] model.

Computational model

The computational model used in the present study is based on the Eulerian–Eulerian approach in which gas and solid phases are treated as a continuum. This multi-fluid model is proposed in the present study simulations because of its fast computation compared to the other multiphase models. Also, the MFM proved reliability in fluidized bed simulations.

Governing equations

The governing equations of fluid flow are applied to both gas and solid phases as in [1] as shown in Table 1:

Gas–solid drag model

The gas–solid drag model is the inherent part of fluidization modeling as it represents the momentum exchange between the gas (fluidization agent) and the solids (the bed particles). In the present study, we are comparing the performance of the top commonly used drag models: the Gidaspow [13] model and the Syamlal–O’Brien [40] model, in addition to the Gibilaro et al. [12] model. The term K_{gs} in the gas and solids momentum conservation equations (Table 1) is the gas–solid momentum exchange. The major force causing momentum interaction is drag force which can be estimated from the empirical drag closures as follows:

Table 1 Governing equations of the present model

Continuity equation of the q th phase

$$\frac{\partial(\epsilon_q \rho_q)}{\partial t} + \nabla \cdot (\epsilon_q \rho_q U) = 0$$

Momentum equation of the q th phase

$$\frac{\partial(\epsilon_q \rho_q U_q)}{\partial t} + \nabla \cdot (\epsilon_q \rho_q U_q U_q) = -\epsilon_q \nabla p - \nabla \cdot \bar{\bar{\tau}}_q + \epsilon_q \rho_q g + (\delta_{qs} J_{gs} - \delta_{qs} J_{gs}) + \delta_{ik} \zeta_{ik} (U_{sk} - U_{si})$$

Energy equation of the q th phase

$$\frac{\partial(\epsilon_q \rho_q H_q)}{\partial t} + \nabla \cdot (\epsilon_q \rho_q U_q H_q) = \nabla \cdot (\epsilon_q \lambda_q \nabla T_q) + (\delta_{qs} J_{gs} - \delta_{qs} J_{gs})$$

$$\delta_{qi} = 1 \text{ if } q = i, \delta_{qi} = 0 \text{ if } q \neq i$$

Stress tensor of the gas phase

$$\bar{\bar{\tau}}_g = \epsilon_g \mu_g [\nabla U_g + \nabla U_g^T - \frac{2}{3} (\nabla \cdot U_g) \bar{\bar{I}}]$$

Stress tensor of the solid phase

$$\bar{\bar{\tau}}_s = (-p_s + \epsilon_s \mu_{b,s} \nabla \cdot U_s) \bar{\bar{I}} + \epsilon_s \mu_s [\nabla U_s + \nabla U_s^T - \frac{2}{3} (\nabla \cdot U_s) \bar{\bar{I}}]$$

$$\mu_s = \mu_{s,col} + \mu_{s,kin} + \mu_{s,fr}$$

$$\mu_{s,kin} = \frac{10 \rho_s d_p \sqrt{\Theta_s \pi}}{96 \epsilon_s (1 + e_{ik} g_{ik})} \left[1 + \frac{4}{5} g_{ik} \epsilon_s (1 + e_{ik}) \right]$$

$$\mu_{s,col} = \frac{4}{5} g_{ik} \epsilon_s \rho_s (1 + e_{ik}) \sqrt{\Theta_s / \pi}$$

$$\mu_{s,fr} = \frac{p_s \sin \theta_f}{2 \sqrt{I_2 D}}$$

- Gidaspow [13] model

$$K_{gs} = \begin{cases} 150 \frac{\varepsilon_s(1-\varepsilon_g)\mu_g}{\varepsilon_g d_p^2} + 1.75 \frac{\rho_g \varepsilon_s |u_g - u_p|}{d_p}, & \text{for } \varepsilon_g \leq 0.8 \\ \frac{3}{4} C_D \frac{\varepsilon_s \varepsilon_g \rho_g |u_g - u_p|}{d_p} \varepsilon_g^{-2.65}, & \text{for } \varepsilon_g > 0.8 \end{cases} \quad (1)$$

where C_D is the drag coefficient of a single sphere and is given as in [35] as follows:

$$C_D = \begin{cases} \frac{24}{\varepsilon_g Re_p} [1 + 0.15(\varepsilon_g Re_p)^{0.687}], & \text{for } Re_p \leq 1000 \\ 0.44, & \text{for } Re_p > 1000 \end{cases} \quad (2)$$

$$\text{where } Re_p = \frac{\rho_g |u_p - u_g| d_p}{\mu_g}$$

- Syamlal–O’Brien [40] model

$$K_{gs} = \frac{3}{4} \frac{C_D Re_p \varepsilon_s \rho_g |u_g - u_p|}{v_{r,s}^3 d_p} \quad (3)$$

where

$$C_D(Re_p) = [0.63 + \frac{4.80}{\sqrt{Re_p/v_{r,s}}}]^2 \quad (4)$$

$$v_{r,s} = 0.5[A - 0.06Re_p + \sqrt{(0.06Re_p)^2 + 0.12Re_p(2B - A) + A^2}] \quad (5)$$

$$A = \varepsilon_g^{4.14}, B = \begin{cases} 0.8\varepsilon_g^{1.28}, & \text{for } \varepsilon_g \leq 0.85 \\ \varepsilon_g^{2.65}, & \text{for } \varepsilon_g > 0.85 \end{cases} \quad (6)$$

- Gibilaro et al. [12] model

$$K_{gs} = \left[\frac{17.3}{Re_p} + 0.336 \right] \frac{\rho_g |u_g - u_p|}{d_p} \varepsilon_s \varepsilon_g^{-1.8} \quad (7)$$

Figure 2 shows a comparison among drag models used in the present study at different voidage and Reynolds numbers. The Syamlal–O’Brien model has the highest drag force values, while the Gibilaro et al. model shows relatively lower drag than the Gidaspow model.

Effect of particles’ sphericity on gas–solid drag

SRF particles have irregular shapes; in order to simplify the modeling of such a complex system, we proposed studying SRF particles overall sphericity ratio on the prediction accuracy. Two non-spherical drag coefficient models were compared, namely the Haider and Levenspiel [17] model and the Dioguardi et al. [8] model given by the following equations:

Figure 2 A comparison among the Gidaspow [13], the Gibilaro et al. [12], and the Syamlal–O’Brien [40] drag models at different voidage and Reynolds numbers

Haider and Levenspiel [17] drag coefficient

$$C_D = \frac{24}{Re_p} [1 + A Re_p^B] + \frac{C}{1 + \frac{D}{Re_p}} \quad (8)$$

Dioguardi et al. [8] drag coefficient

$$C_D = \frac{24}{Re_p} \left[\frac{1 - \psi}{Re_p} + 1 \right]^{0.25} + \frac{24}{Re_p} [0.1806 Re_p^{0.6459}] \psi^{-Re_p^{0.08}} + \frac{0.4251}{1 + \frac{6880.95}{Re_p} \psi^{-5.05}} \quad (9)$$

where A , B , C , and D are fitting coefficients which depend on particle sphericity ratio, while ψ is the shape factor which is the ratio of particle's sphericity to circularity.

We implemented user-defined functions (UDFs) in ANSYS Fluent R18.2 solver to include particles sphericity modification on gas–solid drag force calculations as shown in Fig. 3. This figure indicates the variation on Gidaspow drag force calculated using different sphericity ratio drag coefficients. It is obvious the increase in the drag force due to the decrease in the sphericity ratio.

Simulation method

The finite volume method is used to convert the fluid flow governing equations of both gas and solids phases into linear algebraic numerical equations. The discretized equations are solved by the commercial CFD code ANSYS Fluent R18.2. The 2D fluidized bed is constructed and meshed using ANSYS Design modeler and ANSYS ICEM subprograms, respectively. While, the results are post-processed using ANSYS CFD Post and MATLAB R2017a. The mesh size is set in the range below 10 times the particle size which was found sufficient to achieve good accuracy [2]. Figure 4 represents a schematic of dimensions and boundary conditions applied in the present model.

The bed is initially well mixed with 0.86 mass% SRF overall concentration. The no-slip and partial-slip wall boundary conditions are assigned to gas and solid, respectively. The specular factor is set to 0.6, while particle–wall restitution coefficient is set to 0.9. A summary of simulation parameters and numerical variables used in the present simulations is listed in Table 2.

Results and discussion

Model validation

Spherical Geldart D systems

Literature studies reported that the gas–solid drag model is the most significant element in modeling of binary fluidized beds [3, 49, 51]. Moreover, the Gidaspow model showed good validation in the case of Geldart B binary systems. But, there are no sufficient validation studies regarding Geldart

Figure 3 Effect of particles' sphericity on gas–solid drag force using Gidaspow model

D binary systems. Thus, we chose Geldart D binary systems from the literature for preliminary validation. Figure 5 shows jetsam concentration profiles predicted using different drag models in comparison with experimental data from the literature of Geldart D binary system. The results showed that the Gidaspow and the Syamlal–O'Brien models can give overall good agreement with the experimental data of Geldart D systems of different density ratios. The Gibilaro et al. model cannot give reliable predictions that qualify it to the next phase of model testing.

Figure 4 The bed geometry and the boundary conditions used in the present study

Non-spherical SRF particles system

The typical sphericity of SRF was not measured because such material is heterogeneous in density, size, and shape. For simplicity, we assumed SRF as mono-sized, mono-density, and mono-sphericity particles in the CFD simulations. The mass-weighted average size was estimated from the SRF cumulative size distribution in a previous publication [41]. A validation test is carried out to find out the optimum overall sphericity ratio of SRF particles. In all simulation cases, computations are executed for a total

time period of 30 seconds with time-averaging over the last 10 seconds.

Figure 6 presents results of implementing particles' sphericity ratio on numerical predictions. It can be clearly noticed that SRF particles should not be treated as spherical particles and the optimum overall sphericity ratio of SRF particles which can give the best prediction results is within the range of 0.123–0.23 (see Fig. 6a). But there is a large SRF particle entrainment in case of decreasing sphericity below 0.23 (See Fig. 7). Also, the Syamlal–O'Brien drag model with modified drag coefficient gives poor predictions as it overpredicts drag force and a high-SRF particles entrainment occurs as a result (see Fig. 7d). Treating SRF as spherical particles results in homogeneous distribution of SRF particles over the fluidized bed domain as shown in Fig. 7. Increasing the drag force due to low particles' sphericity results in accumulation of SRF particles in high concentration spots near bed walls. At very low sphericity ratio < 0.23 , the drag force is high enough to entrain the particles outside the column.

SRF mixing in a bubbling bed

A comparison between CFD model predictions and the experimental data of flotsam relative mass fraction at different axial zones within the bed (bottom, middle, and top) with variation of dimensionless fluidization velocity is shown in Fig. 8. The data show a decreasing trend of SRF concentration at bed bottom and bed top at lower fluidization velocity ratios ($ur < 1.6$), while SRF concentration increases from 0.10 to 0.34 at bed body (middle). At elevated fluidization velocities ($ur = 1.6–2.0$), SRF concentration rises at top layer and continues decreasing at bottom layer with a smaller decreasing rate. On the other hand, SRF concentration reaches a saturation state and stabilizes around a constant value of 34%. The experimental results show closer ratios of SRF content within bed bottom, middle, and top in the range of 20–40% at smaller fluidization velocities ($ur < 1.5$). Above this velocity ($ur > 1.5$), bottom composition decreases with a quadratic rate and the top layer concentration increases with an inverse trend to that of the bottom

Table 2 A summary of the present model settings and parameters

Particle viscosity	KTGF	Particle–particle restitution coefficient	0.9 (sand– sand), 0.7 (sand– SRF), 0.6 (SRF–SRF)
Collision viscosity	Gidaspow et al. [14]	Particle–wall restitution coefficient	0.9
Bulk viscosity	Lun et al. [21]	Gas–wall boundary	No-slip
Friction viscosity	Schaeffer [34]	Particle–wall boundary	Partial-slip, specular factor 0.6
Angle of friction	30°	Convergence criteria	1e–3
Max. friction packing	0.61	Max. Iter/time step	50
Granular temperature model	Algebraic	Time step	0.001 s
Max. packing	0.63	Total simulation time	30 s

Figure 5 Jetsman concentration prediction using different gas–solid drag models in comparison with experimental data of Geldart D binary systems

layer. This is clearly shown by SRF velocity vectors where motion of particles is observed of overall migration from bottom to top layer. The CFD predictions are carried out on sand of medium jetsam particle size ($d_p = 810 \mu\text{m}$) and flotsam SRF particles of equivalent spherical mean diameter ($d_p = 3520 \mu\text{m}$). Two CFD solutions are presented in this figure: the first one is considering SRF as spherical particles (i.e., sphericity ratio = 1.00), while the other solution

Figure 6 Effect of SRF particles' sphericity ratio on CFD predictions in comparison with experimental data from the literature (fluidization velocity ratio = 2.0, $d_p = 810 \mu\text{m}$)

includes using UDF for modifying drag coefficient (i.e., SRF particles of sphericity = 0.23) [49]. It can be clearly noticed from Fig. 8 that the CFD model which uses modified drag coefficient gives overall acceptable agreement with the experimental data. The lower sphericity ratio solution predicts the same concaving curve of the experimental data of SRF relative concentration at bed top as well as the convex shape curve of SRF relative concentration at bed body (middle). The CFD model of higher sphericity ratio gives better result in bed bottom regime at low fluidization velocity ratios ($ur < 1.5$), while over this range the lower sphericity solution is in good agreement with the experimental data. The figure shows also that the CFD model using spherical SRF particles (i.e., sphericity ratio = 1.00) cannot be reliable and gives overall poor prediction results.

The bed material (sand) particles' mean size is an important parameter in fluidization process. Thus, we tested three sand samples of three different mean sizes (654 μm , 810 μm , and 1110 μm). Figure 9 shows a comparison between SRF relative mass fractions at a moderate fluidization velocity in three bed locations: bottom, middle, and top using different particles' sizes. The experimental results show a significant influence of bed material on SRF concentration within the bed. The biggest

Figure 7 Effect of particles sphericity ratio on SRF void fraction contours in a bubbling bed ($ur = 1.6$, $d_p = 810 \mu\text{m}$)

mean-size bed ($d_{pm} = 1110 \mu\text{m}$) has the highest segregation behavior, where SRF concentration is minimum at bed bottom and maximum in bed middle. The opposite trend occurs in the smallest bed of mean size ($d_{pm} = 654 \mu\text{m}$). The present model gives overall acceptable predictions with the experimental data.

Figure 8 Flotsam velocity vectors and relative mass fraction at three different bed axial regimes (bottom, middle, and top)

Figure 9 Comparison between SRF relative mass fraction at different axial bed locations (bottom, middle, and top) at fluidization velocity ratio, $ur = 1.6$ and using different bed material sizes

Thermal field analysis

Heat transfer is one of the greatest advantages of fluidized bed due to high thermal exchange potential among fluidization gas and solid bed material. Generally, in fluidized bed applications, the solid bed material is of high thermal conductivity (low conductive resistance) which makes the overall heat transfer coefficient simply independent of solid particles material. The most widely used closure for gas–solid heat transfer is the Gunn's correlation [16] which is valid for voidage > 0.35 and Reynolds number up to 10^5 . The Gunn's correlation is given as follows:

$$Nu = \frac{h_{gs} d_p}{\lambda_g} = (1 - 10\epsilon_g + 5\epsilon_g^2)(1 + 0.7Re_p^2 Pr^{\frac{1}{3}}) + (1.33 - 2.4\epsilon_g + 1.2\epsilon_g^2)Re_p^{0.7} Pr^{\frac{1}{3}} \quad (10)$$

where Prandtl number is defined as $Pr = \mu_g C_{pg} / \lambda_g$ and C_{pg} is the specific heat of the gas phase.

Thermal model validation is important part before performing any numerical investigations. Thus, we used two different experimental cases from the literature representing wall–bed and gas–solid heat transfer mechanisms. Table 3

Table 3 Thermo-physical properties of the experimental validation cases from the literature

Property	Yusuf et al. [48]		Simsek et al. [37]	
	Gas	Solids	Gas	Solids
Diameter/ μm	—	491	—	12600
Density/ kg m^{-3}	1.2	2485	1.205	1440
Viscosity/ Pa s	$1.8\text{e-}5$	—	$1.82\text{e-}5$	—
Specific heat/ $\text{J kg}^{-1} \text{K}^{-1}$	994	737	1005	1350
Thermal conductivity/ $\text{W m}^{-1} \text{K}^{-1}$	0.026	1.0	0.0243	0.16

shows the conditions of the two exciting experimental systems.

Figure 10 shows a comparison between present thermal model predictions and experimental data from the literature of gas–solid and wall–bed heat transfer. This figure demonstrates that the macroscopic wall–bed heat transfer phenomenon can be well predicted by using the present MFM model. However, this MFM cannot predict correctly the mesoscopic gas–solid heat transfer and a higher-resolution simulation approach is needed such as the discrete element model (DEM) which gives good agreement with the experimental data. Based on this validation, we limit the numerical study to investigate wall–bed heat transfer of SRF–sand binary system.

Figure 10 Thermal model validation with experimental cases from the literature **Figure 11** Effect of SRF sphericity ratio on wall–bed heat transfer coefficient predictions

No studies in the literature have concerned with studying effect of particles sphericity on heat transfer coefficient in bubbling beds. Thus, we extended the numerical study to include this important aspect. In order to study the thermal field, we assumed air enters the bed with a temperature of $27\text{ }^{\circ}\text{C}$ and heat is supplied from the side walls ($T = 127\text{ }^{\circ}\text{C}$). Accordingly, simulations were extended for another 10 seconds and the resultant wall–bed heat transfer coefficient is recorded for each fluidization velocity. Figure 11 shows a comparison among model heat transfer coefficient predictions using different SRF sphericity ratios. It is clear from this figure that the wall–bed heat transfer coefficient decreases with increasing fluidization velocity. Also, heat transfer results predicted using low sphericity model is lower than that of the default spherical drag model.

Conclusions

Simulation of mixing and segregation in binary fluidized beds is a challenging problem, especially when it comes to heterogeneous material like SRF. In this study, Eulerian–Eulerian multi-fluid model was set up to simulate mixing behavior of large SRF particles in a bubbling fluidized bed. SRF of irregular-shaped Geldart D particles was modeled numerically in a binary mixture with Geldart B sand particles. The experimental data included the effect of the fluidization velocity ratio and effect of bed particles' sizes on the SRF relative concentration. The results concluded that SRF particles cannot be simulated as spherical particles, while assuming them as disks of overall sphericity ratio of 0.23 can give acceptable agreement with experimental data. The fluidization velocity has a significant influence on mixing of SRF particles. SRF concentration was found decreasing at bed bottom and bed top layers at low

fluidization velocities ($u/u_{mf} < 1.6$), while increasing at bed body (middle layer). At elevated velocities ($u/u_{mf} > 1.6$), SRF concentration in top layer was increasing, and decreasing with same rate from bottom layer. The biggest mean-size bed material ($d_{pm} = 1110 \mu\text{m}$) was found of the highest segregation behavior.

Regarding the thermal field, the present model was validated by experimental data from the literature of wall–bed and gas–solid heat transfer. The model showed good agreement with wall–bed experimental heat transfer data from the literature. Finally, wall–bed heat transfer coefficient decreased with increasing fluidization velocity.

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