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CFD Analysis of the Fluidised Bed Hydrodynamic Behaviour inside an Isothermal Gasifier with different Perforated Plate Distributors

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Abstract

The hydrodynamic behaviour of gas-solid mixtures inside Bed Fluidised Beds (BFB) gasifiers has a major impact on the gasification process due to particles - gas and particle - particle contact mechanisms. The Discrete Phase Model DPM with Multiphase Particle-in-Cell method MPPIC was used as a CFD approach to study the hydrodynamic behaviour of an 800 height x 83 mm Ø prototype fluidised bed gasifier with 4 different perforated plate distributors. In terms of bubble forming, pressure drop and superficial velocity, the type D distributor, i.e. triangular, had the best performance among the other types and in turn better bed height and bed movement, thus allowing best fluidisation performance as a consequence of better flow distribution.

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Keywords: Fluidised bed; Gasification; BFB; DPM; MPPIC; Distributor.

1. Introduction

In the light of renewable energy demand, the attention to using biomass as a sustainable source of energy has increased vastly. There are many effective methods used to extract energy from biomass, one being thermal gasification. In a gasification process, the biomass is converted into a synthetic gas (syngas) by hydrodynamical and chemical interactions in the presence of air or oxygen. There are several types of gasifiers depending on many factors such as feeding method, temperature range, and heating method [1].

Fluidised bed gasifiers are one of the best gasification systems. In the fluidized bed gasification, the biomass particles have a better chance to mix with fluid inside the gasifier which makes some of these particles to suspend in the fluid (gas) stream [2]. However, the hydrodynamic mechanism surface–bed, fluid-particle, and particle – particle collisions are still in development due to the complex interaction between particle - particle and particle – surface.

1.1. Particle – Fluid Flow Models

Recently, modeling of the hydrodynamic mechanism of particle – fluid fluidisation has received growing interest.

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Generally, there are two main models widely used to describe this mechanism.

1.1.1. Two - Fluid Model (TFM)

This model is based on a Eulerian - Eulerian continuum approach [3,4]. In this approach, both fluid and solid phases are continuous and fully interpenetrating [5] and both momentum and heat transfer can be exchanged between the phases. Normally, the solution of TFM approach relies on grid-based methods such as finite difference method FDM and finite volume method FVM.
However, using two - fluid model to simulate the hydrodynamic of Bubbling fluidised bed BFB with Geldart A particles gives over-predicted results during bed expansion [6,7,8].

1.1.2. Discrete Particle Model (DPM)

In this model, the Newtonian equations of motion are solved for each individual particle, and an interaction model is applied to handle particle encounters. Meanwhile, Navier – Stokes are equations based on the concept of local averaging as used in CFD to solve the fluid flow continuum, taking momentum transfer between the fluid and the particles into account [9]. This model is also called CFD-DEM or EULER - LAGRANGE approach. According to particles collision models, DPM can be divided into two approaches, hard sphere approach [10,11] and soft sphere approach [12].

1.1.2.1. Hard sphere approach

In this model, collisions between particles and between particles and the walls are assumed to happen immediately. Momentum binary collisions are used to determine particle trajectories.

1.1.2.2. Soft-sphere approach

This model was originally developed by Cundall and Strack [13]. In this approach particles deform due to contact is calculated. Particles remain geometrically rigid, and “deformation” is considered in the force models. Typically, soft sphere approach simulations are more time consuming than hard sphere approach.

1.2. CFD Software Packages for fluidised bed simulation

The technical booming in computing speed and data storage have a major impact on CFD development. Nowadays, over 200 CFD software packages are used to simulate different flow problems. Some of those are commercial such as FLUENT and others are open sources such as OpenFOAM.

Hui and Xinhui have used the commercial software (FLUENT) to simulate the effect of using a distributor in fluidised bed flows [14]. Three-dimensional simulations are carried out for the system with and without the distributor. They used two different distributor arrangements (square and triangular). It was found that using gas distributors at the entrance of the fluidized bed is important to form and maintain a constant pressure in the fluidised bed system. On the other hand, they concluded that the fluid velocity becomes more stable when using triangular arrangement than square. However, there is no information about the fluidised bed such as the model used during the simulation, solid particles properties (material type, particle diameter, and particles number) or total solid weight.
In the same context, spout fluidised beds have been simulated by using the open source software package OpenFOAM [15]. DPM-CFD approaches are used to simulate unsteady and nonhomogeneous gas-particle flows. Grid -particle size ratio effects, porosity estimation and drag closure effects on gas-particle flow patterns and pressure drop fluctuations in spout fluidised bed simulations are investigated in their simulation. They predicted a minimum grid size to particle diameter ratio for fluidised bed simulations and a suggestion for a reasonable porosity, ε, for the particle packing processing.

1.3. Gas Distributor

A gas distributor is the most important part in the fluidised bed gasifier. The role of this part is distributing fluidisation gas uniformly through the solid bed. Hence, gas distributor designs have a major impact on the hydrodynamics behaviour and heat transfer of the fluidised bed. Commercially, there are several types of gas distributors in us, which ranged from simple shapes to extremely complex ones. Perforated plate distributors are the most common and simplest to manufacture.

1.3.1. Perforated Plate Distributor Design

The design of gas distributors is subjected to several criteria such pressure drop across the bed (Δp_b), hole (orifice) dimeter (d_o) and holes arrangement [16]. Gas distributors should provide enough pressure drop to initiate fluidisation uniformly. Hence the pressure drop across a bubbling bed with (H) height, can be calculated via the equation Basu [17]:

$$\Delta p_b = \rho_p(1-\varepsilon).H.g = \rho_p(1-\varepsilon_{mf}).H_{mf}.g \quad (1)$$

Where:

ρ_p – Particle density (kg/m^3).

ε – Voidage (Porosity). $\varepsilon = \frac{void\ volume}{volume\ of\ (particles+voids)}$

ε_{mf} – Voidage for minimum fluidisation conditions.

Another important factor in distributor design is the fractional opening area of the orifices:

$$Fractional\ area\ of\ the\ orifices = N\frac{\pi}{4}d_o = \frac{U.\rho_g}{U_o.\rho_{go}} \quad (2)$$

Where:

$N -$ Number of holes (orifices) per unit area of the distributor (m^{-2}).

$N = {}^{1}/_{P^2}$ for square arrangement pitch.

$N = \frac{2}{\sqrt{3}.P^2}$ for triangular arrangement pitch.

P- The pitch between two holes (m).

U, U_o – Superficial gas velocity and orifice gas velocity respectively (m/s).

1.3.2. Minimum Fluidisation Velocity $\boldsymbol{U_{mf}}$:

The Ergun's equation (3) [18] for pressure drop along the packed bed can be used in conjunctions with fluid drag equations to determine the minimum fluidisation velocity (U_{mf}).

$$\frac{\Delta p}{L} = 150\left(\frac{(1-\varepsilon)^2}{\varepsilon^3}\right)\cdot\frac{\mu\cdot U}{(\emptyset\cdot d_p)^2} + 1.75\frac{(1-\varepsilon)}{\varepsilon^3}\cdot\frac{\rho_g\cdot U^2}{\emptyset\cdot d_p} \quad (3)$$

And

$$Fluid\ drag = \Delta p\cdot A = A\cdot L\cdot(1-\varepsilon)(\rho_p-\rho_g)\cdot g \quad (4)$$

Where:

L – Bed unit height (m).

$\mu -$ Fluid dynamic viscosity (${}^{N\cdot s}/_{m^2}$).

$\emptyset$ - Particle sphericity. $\emptyset = 1$ for spherical particles.

Hence, for minimum fluidisation conditions [19]:

$$U_{mf} = \frac{(d_p)^2\cdot(\rho_p-\rho_g)\cdot g}{150\cdot\mu}\left(\frac{\varepsilon_{mf}^3\cdot\emptyset^2}{1-\varepsilon_{mf}}\right) \quad \text{.... For small particles } Re_{p,mf} < 20. \quad (5)$$

And

$$U_{mf}^2 = \frac{d_p\cdot(\rho_p-\rho_g)\cdot g}{1.75\cdot\mu}\cdot\varepsilon_{mf}^3\cdot\emptyset \quad \text{.... For large particles } Re_{p,mf} > 1000. \quad (6)$$

If ε_{mf} and/or $\emptyset$ are unknown, the U_{mf} can be calculated by:

$$U_{mf} = \frac{\mu}{d_p\cdot\rho_g}\cdot[(C_1^2 + C_2\cdot Ar)^{0.5} - C_1] \quad (7)$$

Where:

Ar - Archimedes number.

$$Ar = \frac{\rho_g\cdot(\rho_p-\rho_g)\cdot g\cdot d_p^3}{\mu^2} \quad (8)$$

C_1 and C_2 are empirical constants can be found experimentally as [20]:

C_1 = 27.2 and C_2 = 0.0408.

2. Methodology

2.1. System setup

To improve the hydrodynamic properties of BFB gasifier, a better air distributor should be chosen to provide suitable air distribution, uniform pressure drops through the bed, best porosity, optimum bubble size and better air - particle contact. An 800-mm height prototype fluidised bed gasifier with 83 mm inner diameter was used to test the performance of different perforated plate air distributors. The tests were carried out for 4 different perforated plate distributors, as shown in Fig. 1.

Type	Hole dia.(mm)	Pitch(mm)	Arrangement	No. of holes
A	2	7.5	Square	97
B	2	7.5	Triangular	109
C	1	6.35	Square	137
D	1	6.35	Triangular	151

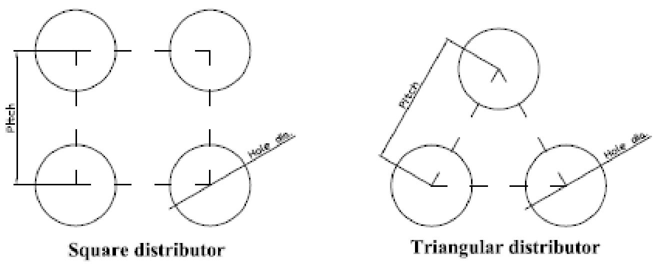

Fig. 1. Specifications of the perforated plate distributors.

2.2. Simulation model

The numerical simulations of this 4 distributor cases were carried out using the open source software (OpenFOAM) version 2.3.0. operating in Linux environment. The Discrete Phase Model (DPM) with Multiphase Particle-in-Cell method (MPPIC) was the approach used for particles collision. Gasifier geometry and mesh generating were done by SALOME-7.7.1 then exported to OpenFOAM as UNV file format to complete the simulation. Table 1. presents the main input data used in the simulation.

Table 1. Simulation parameters.

Parameters	Unit	Value
Silica sand	kg	0.5
Particle size, d_p	μm	478.43 (425-500)
Particle density, ρ_p	kg/m^3	2650
Gas density (air), ρ_g	kg/m^3	1.2
Gas kinematic viscosity, υ	m^2/s	1.56 x 10-5
Particles injection mode	-	Patch injection
Minimum fluidisation velocity, U_{mf}	m/s	0.2
Minimum gas flow rate, q_{min}	CLM	100
Maximum gas flow rate, q_{max}	CLM	145

3. Results

The simulation was done with aid of RAVEN, the supercomputer processing facility of Cardiff University. Data was generated from the simulation of various cases, each averaging 230 GIGABYTES. Results of the internal interaction within the gasifier were produced for 8 seconds with 0.01 seconds time step. All hydrodynamic properties for (sand – air) fluidised bed at any point inside the gasifier were determined. The distributors were tested for two mean superficial velocities of U_m= 0.33 m/s and 0.44 m/s while the minimum fluidisation velocity was U_{mf} = 0.21574 m/s. The simulation shows that all distributors succeed to form bubbling fluidisation phase in the BFB gasifier. Moreover, the main hydrodynamic parameters were studied for each distributor case, these include porosity (Voidage), pressure drops and instantaneous superficial velocity.

3.1. Porosity

As porosity, $\boldsymbol{\varepsilon}$, represents the density of particles in the bed, it can be used as indicator to the bubble forming and size in the fluidised bed. Fig. 2 shows a cross section of the fluidised bed gasifier presenting ($\boldsymbol{\varepsilon}$) for the different time. Bubbles forming in the system can be clearly seen in all types of configurations, especially those at high mean superficial velocities U_m =0.44 m/s. Bubble forming for type D was much more than the other types.

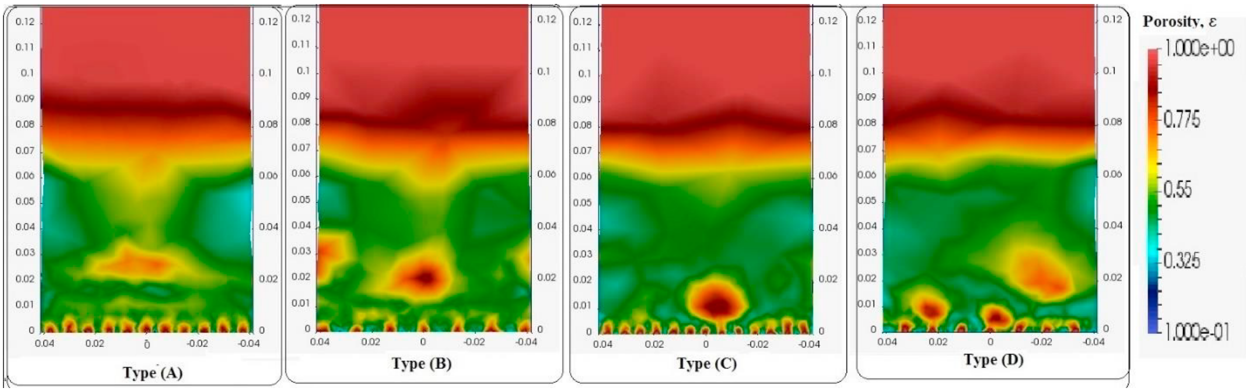

Fig. 2. Porosity contours in the gasifier cross section, U_m=0.44m/s.

3.2. Pressure drops

The pressure drop is an important factor that drives the fluidisation process inside BFB gasifier due to his effect on bed height and movement. Fig. 3. presents the average pressure drop along the gasifier bed for the 4 distributors. All types show proportionally high average Δp for U_m= 0.44 m/s than U_m= 0.33 m/s which means more bed height. However, type D distributor provides the highest average Δp among the other types for the high superficial velocity U_m with bed height to diameter ratio (H/D_o) exceeds 1.

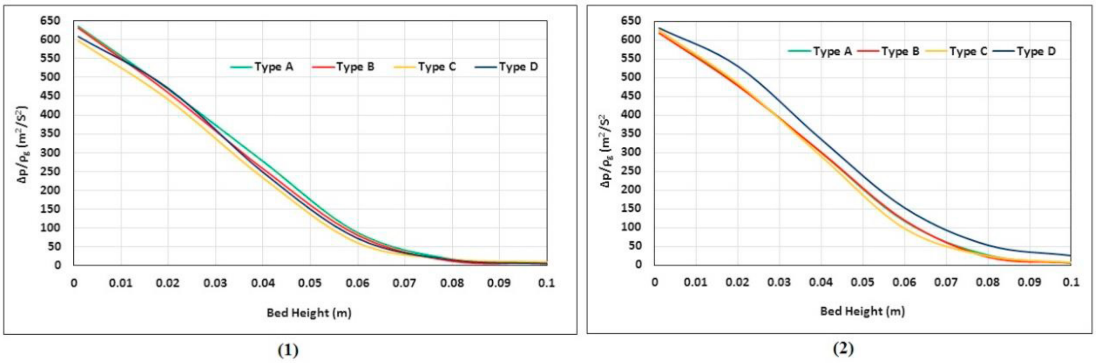

Fig. 3. Average pressure drops across bed height at X=0.0206 m, (1) U_{mf}=0.33 m/s. (2) U_{mf}=0.44 m/s.

3.3. Instantaneous superficial velocity

Another important hydrodynamic property for the fluidised bed which has a major impact on heat transfer between solid particles and gas is the superficial velocity of the gas (air) U_s. In Fig. 4, the behaviour of U_s at a point inside the (air - sand) bed were presented for the four types. Generally, type D distributor provides higher U_s than other types along the time frame used for the simulation for both high and low mean superficial velocities. In comparison with the minimum fluidisation velocity U_{mf} and at bed height H=0.08 m, U_s from type D distributor was higher than U_{mf} for almost the time frame used especially for high mean superficial velocity.

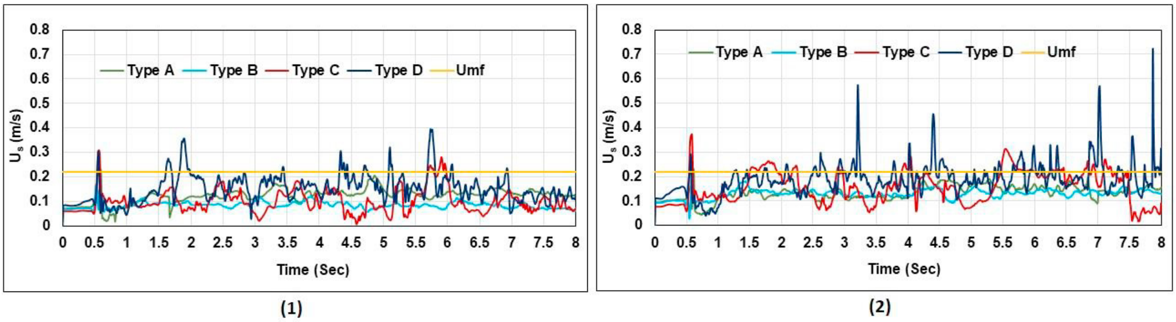

Fig. 4. Instantaneous superficial velocity for the point (0.0309,0,0.08) m in the gasifier domain: (1) U_m= 0.33m/s, (2) U_m= 0.44 m/s.

4. Conclusions

As it is very difficult to measure the hydrodynamic performance inside the BFB gasifier, a DPM with Multiphase Particle-in-Cell method (MPPIC) approach was used to simulate a BFB gasifier with 4 different distributors. This simulation was carried out by the open source software (OpenFOAM) for 8 seconds time frame. The simulation provides a complete image for the hydrodynamic behaviour of the (sand – air) interaction inside the gasifier system. Examining the three main hydrodynamic parameters (Porosity (ε), Pressure drops (Δp) and instantaneous superficial velocity (U_s)) shows that all 4 types succeed increating bubbling fluidisation phase. However, type D distributor with triangular pitch has the best performance among the other types in bubble forming, pressure drop, superficial velocity distribution and in turn better bed height and bed movement thus allowing best fluidisation performance as a consequence of better flow distribution.

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