Influence on the fluidization pattern of a freely bubbling fluidized bed with different modes of air supply

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Abstract

Bubbling fluidized bed (BFB) reactors are extensively used in several process applications like gasification, pyrolysis, drying, and combustions due to their excellent mixing properties and good temperature control. The bubble dynamic and particle movement in the reactor is primarily responsible for uniform heat and mass transfer and mixing. The properties of bubbles in BFB are governed by the gas distribution inside the reactor or supply of the fluidizing gas. This work investigates the influence on the fluid dynamic behaviour of the BFB reactor at different fluidizing gas injection systems using the Computational Particle Fluid dynamic model. Three different modes of fluidizing gas injection include uniform injection, air injection via twenty-five nozzles, and air supply via side nozzles along the reactor height in a gasification reactor of 10.04 cm diameter. Air is used as the fluidizing gas and silica sand as the bed material. The CPFD model is developed in Barracuda Virtual reactor 20.01. The CPFD model is validated against the experimental data obtained from the Electrical Capacitance Tomography (ECT) sensors. The result depicts the better fluidization quality of the bed with uniform air supply as flow boundary and air injection via twenty-five nozzles located at the bottom of the reactor. With air injection via two side nozzles along the reactor height, the bed is fluidized with large bubbles and particle entrainment in the freeboard zone of the reactor. A method is proposed to improve the fluidization quality of the bed while using side nozzles as inlet flow Boundary Conditions (BC) for air injection. The proposed method includes addition of four nozzles along the reactor wall instead of two which improves the fluidization quality of the bed in terms of smaller size bubbles without particle entrainment in the freeboard region.

Keywords: Gas distribution, flow boundary conditions, bubbling-fluidized bed, bubble properties, CPFD

1. Introduction

Fluidized bed reactors are widely used for several industrial applications like waste to energy conversion, chemical synthesis, granulation, drying of pharmaceutical products and raw agricultural products, chemical looping, catalyst regeneration, biomass gasification, pyrolysis etc. (Jaiswal et *al*., 2020; Bandara et *al*., 2021; Singh and Gbordzoe, 2017; Chang et *al*., 2013). The efficiency of the fluidized bed reactors largely depends on the gas distribution inside the reactor since the gas distribution influences the conversion process and the fluidization regime under which the reactor is operated. For instance, during the gasification of biomass or wastes using a bubbling fluidized bed reactor, the carbaneous feedstock is converted into higher calorific value gases in the presence of limited amount of oxidizing agent (Jaiswal et *al*., 2020). The amount of oxidizing medium present for the feedstock conversion depends on how well the fluidizing gas is distributed across the reactor cross-section. Similarly, the hot bed material which is in a continuous motion in such reactors acts as the thermal flywheel and provides the required heat for thermal degradation of the feedstocks. The particle motion is governed by the bubbles rising in the bed and the properties of the bubbles for example bubble rise velocity, bubble diameter and bubble frequency is determined by the gas distribution inside the reactor. Additionally, the mixing phenomena of large biomass particles with bed material and the operating regime of the reactor is determined by the gas distribution (fluidizing gas) to the reactor. The fluidizing gas can be supplied to the particle bed through the distributor or nozzles (Basu *et al*., 2006). The most common method has been the use of a distributor plate that allows to distribute the fluidizing gas uniformly, supports the bed material, provides good gas-solid mixing, prevents channeling, and minimizes dead zones in the reactor (Depypere and Dewettinck, 2004). While using the air distributor or distributor plate, the supply gas velocity has to overcome the distributor plate resistance and lift entire mass of the particles against the gravity in order to fluidize. Alternatively, the fluidizing gas can be supplied to the reactor via orifice or nozzles which can be on the side of the reactor wall or at the bottom of the reactor. It is essential to characterize the fluidized bed behaviour for the specified fluidizing gas flow boundary conditions for a smooth operation of the reactor (Sasic and Johnsson, 2005). Each of the fluidizing gas supply methods, with or without distributor, nozzles, orifice has its own limitation and advantages. For instance, the use of a distributor plate increases the auxiliary power consumption required to pump the gas through the reactor. Additionally, the distributor plate has to be selected depending on the reactor types and process as the distributor plays a critical role in the reactor performance (Raza et *al*., 2021). In addition, there are challenges in operating the reactor with the distributor plate, which requires cleaning and maintenance due to clogging of the pores of the distributor plate by the fine particles and sintering. The blockage of the pores in the distributor plate can lead to local de-fluidization and dead zones in the reactor. In this regard, operating the fluidized bed reactor without a distributor can be of great advantages as it can save the operational cost, and the construction and design cost of the distributor plate. At the same time, it helps to avoid the problem associated with the air distributor. Several studies have been conducted on design of the distributor plate and the influence of the distributor plate on the reactor performance (Geldart and Baeyens 1985; Saxena et *al*., 1979). However, the number studies on operation of a bubbling fluidized bed reactor without an air distributor and comparing the fluid dynamics behaviour of the bed operated with and without distributor are scarce (Agu et *al*., 2018).There are a few studies on operation of a fluidized bed reactor without an air distributor where large size particles are used at the lower region of the bed and the bed is fluidized by passing the gas through the bed of stationary large particles (Agu et *al*., 2018).

In a single study, reported on the operation of reactor without an air distributor Agu *et al*. has studied the bed behaviour with different types of particles and gas velocity based on the pressure and solid circulation using CPFD simulation. However, no information is provided about the grid size and number of computational cells in the article (Agu et *al*.,2018). The grid number or computational domain in such simulations have significant impact on the fluid dynamics behaviour of the bed and the pressure and flow boundary conditions. Additionally, there are no studies in the literature that mention the influence on the bubble properties in a fluidized bed reactor operated without a distributor plate. Therefore, more work is required to characterize the fluid dynamics behaviour of the bed without an air distributor and compare the fluid dynamics behaviour of the bed with a distributor and different modes of gas supply to the reactor.

The objective of this work is to investigate the fluid dynamics behaviour of the bed without air distributor and compare with uniform air supply methods. The uniform air supply methods include use of an air distributor and nozzles. A cold flow model of the fluidized bed reactor with different air injection methods are simulated using a CPFD model developed in Barracuda VR 20.01. For the case without air distributor the bed is fluidized with air supply from two nozzles (holes) located at the opposite side of the reactor wall. The fluid dynamics behaviour of the bed in terms of solid fraction fluctuation and bubble properties are reported briefly. The results from the CPFD simulations are compared with experimental data obtained from a cold fluidized bed equipped with ECT sensors and an air distributor. A method to improve the fluidization quality of the bed without an air distributor is proposed.

2. Material and methods

2.1. Experimental set up

A cold fluidized bed reactor equipped with ECT sensors, and a data acquisition system are used for the experiments. The reactor is 10.04 cm in internal diameter and 150 cm in height. The reactor column is fitted with an air distributor at the bottom and is open at the top. The air distributor is 3mm thick with a 10.04 cm internal diameter. It has a porosity of 40% with a flow area of 36.6 cm^2 which allows the fluidizing gas to pass through the bed uniformly. The reactor is equipped with twin-plane ECT sensors that are located at 15.7 cm and 28.7 cm from the air distributor. Each of the sensors consists of 12 electrodes mounted on the outer wall of the reactor that allows to capture raw data in the form of matrix or images. The online images are extracted from the capacitance measurements using the Linear Back Projection algorithm. The cross-section of each sensor is divided into 32*32 square pixels of which 812 are the effective pixels that lie within the bed. Each pixel holds a normalized relative permittivity value between 0 and 1, which represents the solid-gas fraction. The details of the reactor set up can be found elsewhere (Agu et *al*., 2019). The reactor is filled with sand material and fluidized by using compressed air. The fluctuation of solid volume fraction is measured from the transient data for each gas velocity. The raw data from the experiments were processed in MATLAB to obtain the solid fraction fluctuation, bubble properties, and bed flow dynamics behaviour.

2.2. Simulation set up

The CPFD simulations in this work have been carried out using a commercial software Barracuda VR which consists of numerical codes specially designed for applications in the multiphase flow systems like industrial fluidized bed reactors.

Initially, the reactor is filled with bed material with a static bed height of 30 cm. The properties of bed material like density (2650 kg/m3), and particle size distribution, sphericity (0.86), close pack volume fraction (0.63), are defined similarly to that of the experiment. The particle size distribution was measured from the sieve analysis with mean diameter 423µm.

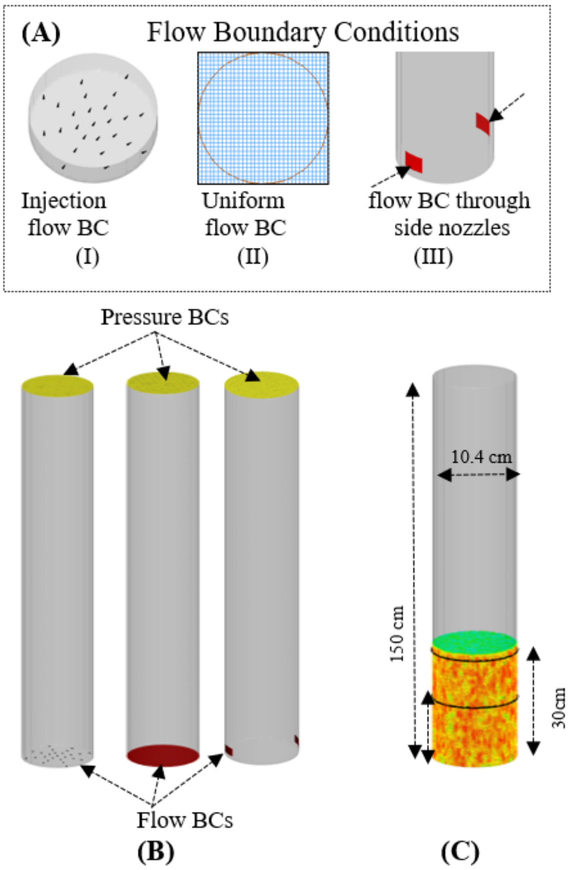


Figure 1: (A-B) flow and pressure boundary conditions, (C) flux planes, initial bed height and reactor dimension.

For the simulation of gas-particle flows, 3D multiphase particle-in-cell approach is used where solid particles are modeled as discrete Lagrangian methods and the fluid is modeled as Eulerian grid of cells. To create a virtual reactor, a CAD geometry equal to the experimental column is imported to Barracuda VR. A uniform grid of total 102400 cells were defined which provides the control volume for all fluid fields calculations. The reactor is operated at atmospheric pressure. Therefore, the pressure boundary condition is defined at the top of the reactor. Three different flow boundary conditions were set up for different modes of gas supply to the reactor as shown in Fig. 1(A). The fluidizing gas is supplied to the reactor using injections points, uniformly distributed along the reactor cross-section (grids or distributor) and through side nozzles (holes). 25 injection points were defined along the reactor cross-section shown in Fig. 1A(I). The mass flow rate through each nozzle (injection points) was equally distributed. Similarly, for the second case, uniform flow boundary conditions are defined at the bottom of the reactor. The reactor cross-section is divided into 1024 cells. The 812 effective pixels or grids that cover the reactor cross-section as shown by the red circle in Fig. 1A(II)) is used as the flow BC. For the third case, two nozzles or holes located on the opposite side of the reactor wall are defined as the flow boundary conditions (as shown in Fig. A(III)). The size of the nozzles is 0.5 cm in diameter. Also, two planes are set up at heights 15.7 cm and 28.7 cm along the height of the reactor column to measure the transient raw data for each of the flow boundary conditions. The drag model used in fluidized bed simulations is an important factor that determines the force acting on a particle by the flow of fluid around it. In this work, the Wen-Yu and Ergun blended drag model is used where the Wen-Yu drag model is suitable for the dilute phase and the Ergun model is used for the dense phase. The details of the drag model, and the governing momentum and force equation can be found elsewhere (O'Rourke and Snider, 2012; Sinder, 2001; Andrews and O'Rourke, 1996; Weber et *al.*, 2013).

3. Results and Discussion

For maximum conversion efficiency, it is important to operate a BFB reactor above minimum fluidization velocity and within a bubbling regime. In this work employs methods to identify the fluidization quality of the bed based on solid volume fraction fluctuation and bubble properties measurement. In addition, influence of different air supply modes on the fluid dyncmics behaviour of the bed are presented.

3.1. Model Validation

The CPFD model is validated by comparing the solid volume fraction fluctuation measured from the experimental data and CPFD simulations at different gas velocities as shown in Fig. 2. As the air was supplied through the static bed, initially the bed expanded as it reached the superficial gas velocity 0.065 m/s. The bed exhibited into the fluidization regime at superficial gas velocity 0.075 m/s. The results depict that the CPFD model followed a similar trend to that of minimum fluidization velocity and bed expansion. However, the solid fraction fluctuation during bed expansion was higher as predicted by the CPFD model compared to the experimental data. This may be due to the difference in the initialization (packing) of the bed material in the CPFD model compared to the experimental conditions.

Further, the bubbles rising in the bed at the superficial gas velocity of 0.1 m/s at different time steps are compared to check the model robustness. The bubbles rising in the bed at a pre-defined plane in the simulation setup and ECT sensors in the experimental set up were captured for both the

CPFD simulation data and the experimental measurement as shown in Fig. 3. The size of the bubbles and the path along which the bubbles move upward (along the center of the bed) in the bed were similar for both experimental tests and the CPFD simulations. However, the number of bubbles predicted by the CPFD model is higher at a lower superficial gas velocity (0.1 m/s). Due to increase in area of the flow boundary conditions in case of CPFD simulation, smaller size bubbles appeared in the CPFD simulation compared to experimental data.

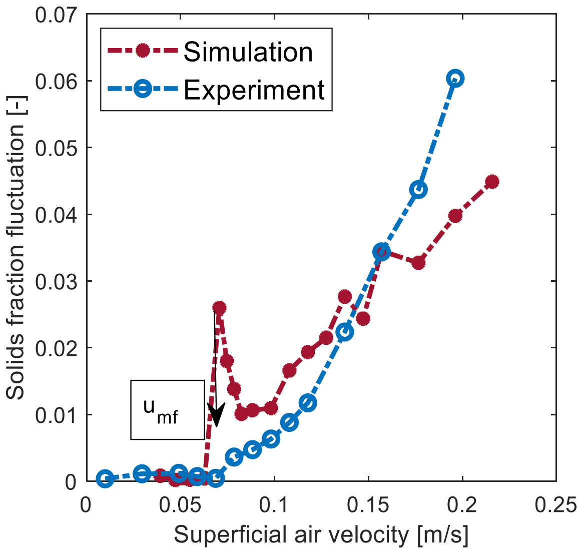


Figure 2: Solid fraction fluctuation at different superficial gas velocities obtained from CPFD simulation and experiment.

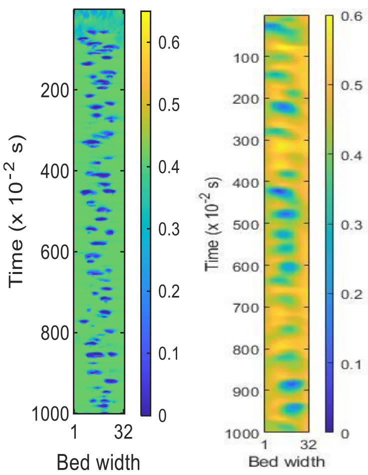


Figure 3: Comparing the rising bubbles in the bed obtained from the CPFD simulation (left side) and experiment (right side) as it reaches the plane at 15.7 cm at the superficial gas velocity 0.1 m/s.

3.2. Influence on solid fraction fluctuation

For a given superficial gas velocity, the variation of the gas-solid fraction in the bed can be used to characterize the fluidization pattern of the bed. The solid fraction fluctuation across the cross section of the bed at height 27.5 cm and at the superficial gas velocity 0.15 m/s were captured over the measurement of 60s with time steps of 0.001. The average solid fraction fluctuation over the measurement period for different flow boundary conditions (injections BC, nozzle BC, uniform BC) from the CPFD simulation and experimental measurements with air distributor are compared in Fig. 4. The result illustrates that for the experimental measurements with an air distributor the solid fraction was lower towards the center of the bed, and it increases near to the wall of the reactor. A similar trend of solid fraction fluctuation was predicted by the CPFD model with the injection flow BC and the uniform flow BC. However, fluctuation in the solid fraction for both the cases were not smooth compared to the experimental measurements because of smaller bubbles in the bed with the uniform and injection flow BCs. Multiple smaller size bubbles were formed in case of the CPFD simulations due to an increase in flow boundary area as compared to that of the experiment. The decrease in the solid fraction at the lower region of the bed near to the center reveals a higher gas fraction in the region. The gas in a fluidized bed rises from the lower region of the bed to the upper region in the form of bubbles. Therefore, the depression of the solid fraction near to the center of the reactor and the increase near to the wall of the reactor illustrate that the bubbles rise upwards following the path near to center of the reactor. For the flow BCs with two nozzles near to the wall of the reactor, the solid fraction fluctuation was lower near to the wall of the reactor where the nozzles wre defined and near to the center of the bed. The fluidizing gas passes in the form of bubbles where some of the bubbles pass near to the wall while some follow the path near to the center of the bed.

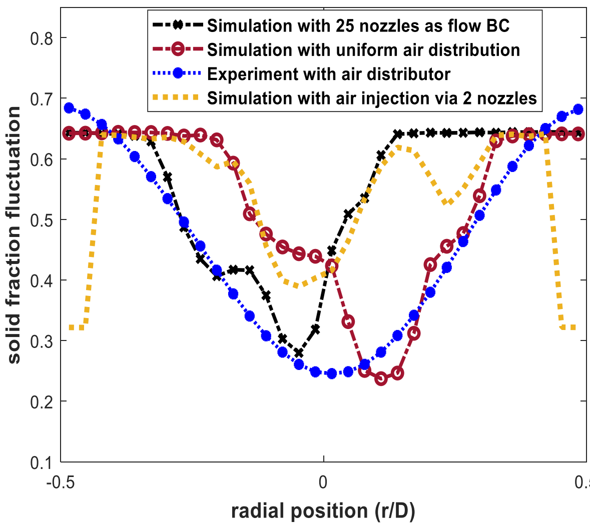


Figure 4: Radial distribution of the solid volume fraction for different flow boundary conditions at superficial gas velocity 0.15 m/s.

3.3. Influence on bubble properties and bed fluid dynamics behaviour

The bubble properties in a BFB reactor significantly influences the fluid dynamics behaviour of the reactor and its performance. The appearance and the movement of the bubbles in the bed is governed by the flow of fluidizing gas through the bed. For the same mass flow rate of the fluidizing gas with different flow boundary conditions, the bubble properties were measured. The influence on the bubble properties and fluidization quality of the bed with different flow boundary condition are compared. The bubbles in BFD can be distinguished from the dense phase by using a bubble solid threshold value. In this work the bubble solid threshold value of 0.2 is used to identify bubbles in the bed. For the CPFD simulations, the bubbles were identified as the zones (object) and the volume of the bubbles were measured by counting (with algorithm written in MATLAB) the number of the cells the bubbles occupied. The bubbles diameter were than calculated from equivalent spherical area.

Fig.5 compares the bubble frequency with respect to the bubble diameter for different flow boundary conditions over the measurement of 30 s. With the uniform flow boundary conditions, the result shows that different bubble sizes within the range of 2.5 cm< d_b <5 cm appeared in the bed with dominant frequency of the bubbles with 3 cm. Here, d_b is the bubble diameter in cm. While for the flow with two nozzles, the bubble diameters are comparatively large for the same mass flow rate (1.5 kg/ hr.) of the air as compared to that of uniform and injection flow BCs. With the injection flow (using 25 nozzles) the bubbles in the bed are smaller and uniform in size as shown in Fig.5(b). The smaller and uniform size bubbles in the bed means better fluidization quality of the bed which contributes to uniform heat transfer and better mixing. On contrary, the large size bubbles in case of flow BC via two nozzles (as shown in Fig.5(c)) can bypass the bed if such bubbles rise in the bed near to the wall of the reactor. Such large bubbles can grow into slugging bubbles as they rise in the bed and turn the bed into a slugging regime. Additionally, with large size bubbles in the bed, the bubble rise velocity in the bed increases significantly which may transform the bed into turbulent regime. When the bed is in the turbulent regime during gasification of biomass in BFB reactor, the fine particles can entraine in the freeboard region which may contaminate the product gases. The entrainment of fine particles in the freeboard region when the air is supplied to the reactor via two side nozzles is shown in Fig.7 F(a). Also, due to increase in gas velocity and bypassing of the fluidizing gas through the side of the bed (near to the wall of the reactor), less oxidizing medium is present for thermochemical conversion of feedstock in the bed where biomass is present. As a result, more oxidizing medium is present in the free board region which can convert the carbon monoxide into carbon dioxide. Therefore, in order to supply the fluidizing gas to the reactor without a distributor (with side nozzles) it is essential to optimize the flow behaviour of the fluidizing gas to the reactor. This can be achieved by increasing the number of nozzles (or holes) along the reactor wall. In this work, the CPFD model was used to simulate a case with four different nozzles along the reactor wall. By increasing the number of side nozzles, entrainment of the bed particles were prevented as shown in the Fig.7(b). Similar, the bubble size is reduced (as shown in Fig.5(d)) and the bubble frequency is increased with large number of smaller size bubbles in the bed. With addition of two more side nozzles, better fluidization quality of the bed is achieved (as shown in Fig.6 and Fig.7.

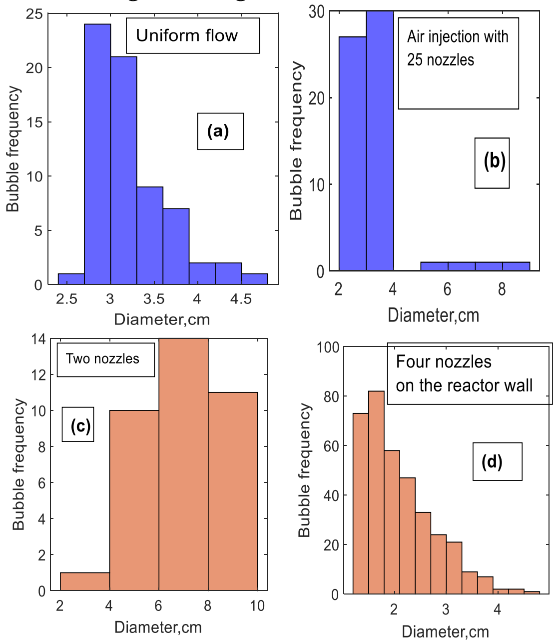
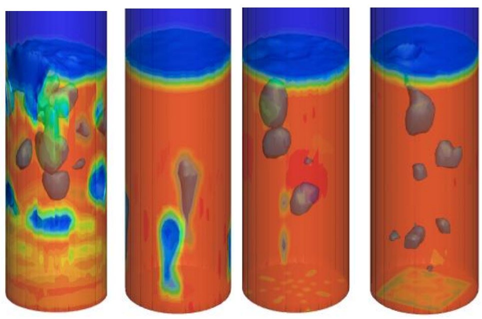


Figure 5: Bubble frequency vs bubble diameter for different air supply systems.



(a) (b) (c) (d)

Figure 6: (a-d), Iso-surface of the bubbles in the bed obtained during, air injection via four nozzles, two side nozzles, 25 nozzles at the bottom of the reactor, and uniform flow BC respectively.

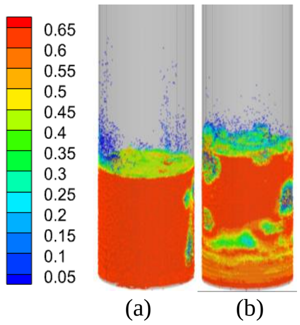


Figure 7: Particle volume fraction of the bed with (a) two side nozzles and (b) four side nozzles.

4. Conclusion

Bubbling fluidized bed reactors are extensively used for several industrial application due to uniform heat and mass transfer. The advantages of a such reactor can only be achieved with proper distribution of the fluidizing gas to the reactor. This work investigates three different methods to supply fluidizing gas to the reactor including uniform flow BC, air injection with twenty-five nozzles and air supply via side nozzles along the reactor wall. A CPFD model has been developed in Barracuda VR 20.01 and the model is used to investigate the fluidization quality of the bed in terms of solid fraction fluctuation and bubble properties. The CPFD model is validated against experimental data obtained from a cold fluidized bed equipped with ECT sensors and air distributor. The result depicts that a better fluidization quality of the bed is achieved with uniform air supply as flow boundary and air injection via twenty-five nozzles located at the bottom of the reactor. With air injection via two side nozzles along the reactor height, the bed fluidized with large bubbles and particle entrainment in the freeboard zone of the reactor. A method was proposed to improve the fluidization quality of the bed while using side nozzles as inlet flow BC for air injection. The proposed method includes addition of four nozzles along the reactor wall instead of two. With the addition of two nozzles, fluidization quality of the bed was improved in terms of smaller bubble sizes without particle entrainment in the freeboard region.

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