CFD Modeling of Cold-Flow Fluidized Beds and Validation with X-ray Computed Tomography

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The hydrodynamics in a 152 mm ID bubbling fluidized bed filled with 500-600 μm glass beads have been simulated using the computational fluid dynamics (CFD) code Fluent 6.3, and the results are compared to experimental data obtained using pressure measurements and 3D X-ray computed tomography. The grids in a Cartesian coordinate system used in simulations are made by 2D (8×8 mm, 4×4 mm, 2×2 mm, and 1×1 mm) and 3D (4×4×4~6 mm). 4 mm grid resolution is sufficient and excellent to use on this scale reactor study to analyze of the time averaged, time and spatial averaged local gas holdup throughout the glass bead bed in 2D and 3D simulations. Throughout the whole fluidized bed, there are two larger symmetric circles of glass beads, in which the gas holdup about 0.46-0.52 is larger than in the middle of the fluidized bed about 0.44-0.46. The 3D simulations with Syamlal-O'Brien and Gidaspow drag models better predict the local gas holdup variation throughout the whole fluidized bed, comparing with the experimental data. The results of simulations with Wen-Yu drag model are generally over predicted. The injector air path extends further into the bed along the axial height from quarter to half of radial long, where the maximum gas holdup is decreasing from 0.9 to 0.6 in the Y-Z slice throughout the fluidized bed with side air injection. But the simulations could not exactly predict the gas path. Because the side air has little effect on the X-Z slice, there remain somewhat symmetric.

Introduction

Fluidized beds are commonly used by the chemical, mineral, pharmaceutical, and energy industries because of low pressure drops, uniform temperature distributions, and the high heat transfer rates. In the biofuel industry, fluidized bed reactors are central components of combustion, pyrolysis, and gasification. The local gas holdup (i.e. solid voidage) distributing in the whole fluidized bed is an extremely important hydrodynamics parameter determining the uniform mass and energy distributions and gasification efficiency. Time-averaged data is more meaningful than transient in the design of commercial-scale fluidized beds.

Because the bed material (solid phase) is typically opaque in gas-solid systems it is more difficult to get data in detail for the whole bed. Since fluidization is a dynamic process, invasive monitoring methods can influence the internal flow, thereby reducing the reliability of the measurements. Currently, there are only a few instances of noninvasing monitoring techniques used with fluidized beds. The techniques for multiphase flows include electrical capacitance tomography, ultrasonic computed tomography, gamma densitometry tomography, X-ray fluoroscopy (radiography/stereography), and X-ray computed tomography, detailed by Heindel et al. (2008). Design of commercial-scale fluidized beds traditionally depends on pilot-scale experiments and too expensive, tough and take time. There has been an increasing usage of computational fluid dynamics (CFD) to complement industrial design of fluidized applications. The situation is complex when multiphase flows are involved, because transport equations for mass, momentum, and turbulence properties have to be solved for each individual phase. A few researchers have attempted to study process characteristics in the fluidized bed using CFD methods and some progress has been made, focusing on terminal settling velocities, minimum fluidization and fluidizability, and residence time of biomass particles. Many numerical simulations of fluidized beds in the literature are tow-dimensional (2D) simulations, mostly due to a lack of computer resources, i.e. memory and processor speeds. To date, even with improved computational facilities, three-dimensional (3D) simulations are still expensive as the equations have to be integrated using small time-steps over long periods of time to compute and average the inherently chaotic and transient fluidization processes; however, parallel computing can speed up the simulational process.

The goal of this research is to find a systemic method for the design of commercial-scale fluidized beds from experimental and simulation techniques. The use X-ray computed tomography and radiography to quantitatively obtain the local time-averaged gas holdup in the fluidized bed was discussed. Experimental validation of local gas-holdup values were done with simulations using Fluent 6.3 to model a comparable bubbling fluidized bed in 2D and 3D. Both quantitative and qualitative predicted local gas holdup distributions calculated using the 3D X-ray flow visualization measurements were compared to the model. The different drag models, Syamlal-O'Brien, Gidaspow and Wen-Yu, were compared in detail. The results of operations with or without side air injection in experiment and simulation were analyzed. The study is useful in understanding how gas holdup is distributed throughout a fluidized bed using different research methods. It is meaningful to the design of commercial-scale fluidized beds.

Experimental Setup

Fluidized Bed Reactor

The fluidized-bed reactor was fabricated from three 152 mm ID acrylic tube sections with 12.7 mm thick acrylic flange plates attached to each end. The 3 sections measure 150 mm, 300 mm, and 640 mm in length, respectively, for the plenum, reactor, and freeboard chambers. Between the plenum and bed chamber is a 1.2 mm thick stainless steel perforated aeration plate containing 130 1 mm diameter holes that are approximately equally space in concentric rings to ensure homogeneous gas distribution and a low pressure drop. The aeration plate has an open area ratio of 0.57%. Air is introduced into the system by an inlet in the bottom of the plenum. The air travels through the marble-filled plenum and is dispersed evenly over the bottom of the aeration plate. Rubber gaskets are placed between each flange, sealing the various components and forcing all gas to flow directly through the bed and out the top. A pressure tap in the bottom of the plenum holds a transducer connected to a data acquisition card to record inlet pressure. A second tap on the side of the fluidized bed 25 mm above the aeration plate is used for side air injection. Data regarding side air injection will not be included in this report however.

500-600 µm diameter glass beads, Geldart Group B, are used as the bed material due to its excellent fluidization and heat transfer properties. The key features of the region are exampled as deaerate quickly when gas is shut off, large bubbles, and intermediate solids mixing. Gas and solid properties of the fluidized bed used in the experiments and simulations are listed on Table 1.

X-ray System

Iowa State University's XFloViz facility was used to image the fluidized bed and has been described in-detail in the literature (Heindel et al., 2005, and Heindel et al., 2008). Only a brief outline will be presented here. Two LORAD LPX200 portable X-ray tubes provide the X-rays. Current and voltage can be adjusted from 0.1 to 10.0 mA and 10 to 200 kV, respectively, with a maximum power of 900 W. Low energy radiation is suppressed by 2 1 mm thick copper and 1 aluminum filters placed directly in front of the X-ray sources. Located opposite each X-ray source is an X-ray detector/CCD camera pair. The CCD camera with the image intensifier has a temporal resolution ranging from 10 frames per second (fps) to 60 fps, depending on binning options, and is primarily used for radiographic imaging. The image intensifier is a 406 mm diameter Precise Optics PS164X screen detector with a 35.0 mm output image diameter. A DVC-1412 monochrome digital camera captures the image from the intensifier. Generally, 2×2 binning (640×512 active pixels) at 20 fps is used for radiographic movies in order to maximize picture quality while maintaining good temporal resolution. A different detector/CCD camera pair is primarily used for CT imaging because of its high spatial resolution. This camera is usually located opposite the second source and is connected to a square 440×440 mm cesium-iodide scintillator screen which transforms radiation into visible light. A 50 mm Nikon lens captures images which are digitized by an Apogee Alta U9 system. This system has 3072×2048 pixels and is thermoelectrically cooled to allow long exposure times. Usually, an exposure time of 1 second with 4×4 binning is chosen to minimize acquisition time while maintaining the signal strength. Both cameras and sources are located on a 1.0 m ID rotation ring that can rotate 360º around the fluidized bed. Data from both CCD cameras are acquired by software developed by Iowa State University's Center for Nondestructive Evaluation (CNDE) and a personal computer with 4 GB of RAM. The software allows for control of both cameras, as well as motion control for the rotation ring. Images acquired by the CCD camera with the image intensifier are normalized for pixel non-uniformity (linearization) and for warping effects produced by the image intensifier. Images acquired by the CCD camera with the CsI screen are also normalized for pixel non-uniformity. These corrections are performed using algorithms within the CNDE software. Volumetric reconstruction of the CT images is performed using CNDE's 64-node LINUX cluster.

CFD Modelling

CFD uses numerical methods to solve partial differential equations that model fluid dynamics. In general, CFD can be applied to reactive, multiphase, laminar and turbulent flows. In this work, governing equations for mass, momentum, and energy for each phase (gas and solid) are solved within a geometrical domain that corresponds to the fluidized-bed reactor used in experiments.

Grid

A Cartesian coordinate system is used for grid generation, where hexahedral elements are formed for the 3D geometry and quadrilateral elements for the 2D geometry. For the 2D simulations we have carried out a grid-refinement study by varying grid size by factors of two for four different grids， 8×8 mm, 4×4 mm, 2×2 mm, and 1×1 mm, respectively, in which a square mesh was used. When side air injection is simulated, the side injector diameter is set to 11 mm with an open area 95 mm^2. The distance between the side air injector centerline and the aeration plate is 30 mm. In order to obtain better meshes, the side air injector area is represented by a rectangle in the 3D grid. The same 3D mesh is used in cases with and without side air injection.

Multiphase Model

There are two approaches for CFD modeling of gas-solid flows: the Lagrangian-Eulerian model and the Eulerian-Eulerian. Using the Lagrangian-Eulerian model, trajectories of each particle are tracked by solving individual equations of motion, whereas the gas phase is modeled using an Eulerian framework.

As a consequence, the Lagrangian-Eulerian model requires large computational resources for large systems of particles. With the Eulerian-Eulerian, the base assumption is that gas and solid phases are interpenetration continua. Therefore, the Eulerian-Eulerian model for gas-solid flows is the more commonly used CFD model to predict the dynamic behavior of dense phase fluidized bed reactors.

Numerical Methodology

CFD simulations of the experimental fluidized bed reactor are carried out with Fluent 6.3 on the High Performance Computing (HPC) machines at Iowa State University. The time step for the simulations is calculated based on the convective Courant-Friedrichs-Levy condition, which depends on the ratio of the grid size and maximum velocity in the flow domain. The time step is set in the range of 10^{-3}-10^{-4} s, depending on simulation conditions. The convergence criterion is set to 10^{-4}. Second-order discretization schemes are used for the spatial and temporal derivatives. Simulations were performed from t = 0 to 70 s. Statistics are not saved from the calculations for t < 10 s to make sure the transients die down. Since the experimental data are time-averaged, all comparisons will be based on time-averaged simulations that span 10 to 70 s.

Results

Grid Resolution Study

The velocity distribution through the gas inlet was uniform throughout all CFD simulations. The overall bulk time-averaged gas holdup distributions show no some substantial changes among the different 2D grids. There are three regions divided by +/- 38 mm, the core region and two wall regions. Experimental values are obtained along mutually perpendicular rays at specified Z heights. There are three zones, the aeration zone 0-38 mm, the fluidized zone 38-152 mm, and the surface zone in the upper portion of the bed. The local fluctuations in the experimental data are due, primarily, to noise in the data. For all 2D simulations, gas holdup is generally over predicted slightly in the core region while being under predicted in the wall region. The condition is not to become better by decreasing the grid size. The experimental gas holdup just above the distributor is smaller due to the effect of the plate, where gas jetting is observed and not all holes are active. Because of the uniform inlet profile used in the simulations, no jetting phenomena are predicted with CFD analysis. The 2D-1 case gives the best overall agreement with the experimental data, while the 2D-4 and 2D-2 cases of give better results in the upper portion of the bed. Based on the analysis of the time averaged, time and spatial averaged local gas holdup throughout the glass bead bed, it is concluded that the 4 mm grid resolution is sufficient and excellent to use on this scale reactor study. The grid size in 3D simulations is also chosen as 4 mm in the following study.

Drag Model Study

Throughout the fluidized bed, there are two larger circulation zones of glass beads both experiments and simulations. Those are almost symmetric in simulations. The glass beads are drawn by pressured air to move upwards and concentrated in the middle of the fluidized bed, where there is a relatively small quantity of gas-phase. Therefore the gas holdup is smaller in this region about 0.44-0.46. After the gas carries the glass beads to the top of fluidized bed, it jets out and the glass beads are then circulated back down the reactor along the walls. Some bubbling gas is formed in the two circulation zones; therefore, gas holdup is larger in these regions about 0.46-0.52, where the values by Wen-Yu drag model are larger than the results from Syamlal-O'Brien and Gidaspow drag models qualitatively. And its abrupt change in gas holdup near the top of the bed is higher than that of experiment and the others simulations.

With Side Air Injection

With side air injection, a path of high gas holdup extends from the injector to the bed surface. This air path gradually expands into the bed as it rises up the bed, indicating that horizontal dispersion increases with axial height. But the simulations could not exactly predict the gas path. The radial velocity is under predicted. The side air injection could not break the bulk flux regions throughout the fluidized bed, just following to go through along the wall. Side air injection also produces a non-uniform average bed height above the injector. The highest average bed height occurs near the wall, directly above the injector. The horizontal level Z = 0.25D is just above the side air injection position, where the maximum gas holdup is above 0.9. And simulations can capture the effect and quantitatively agree with the experimental data. The injector air path extends further into the bed along the axial height from quarter to half of radial long, where the maximum gas holdup is decreasing to about 0.6-0.7. The gas holdup is arranged almost from 0.44 to 0.52 throughout the whole fluidized bed in the X-Z slice, because the side air has little effect on the regions, which remain somewhat symmetric.

Conclusions

The Eulerian-Eulerian CFD simulations based on Fluent 6.3 software for a 152 mm ID bubbling fluidized bed containing 500-600 μm glass beads give results in 2D (8×8 mm, 4×4 mm, 2×2 mm, and 1×1 mm), and 3D (4×4×4~6 mm) simulations using three different drag models with or without side air injection. The study is useful in understanding how the gas holdup is distributing in the whole fluidized bed by different research methods including with X-Rays experiments and CFD modelling. It is meaningful to the design of commercial-scale fluidized beds in the continuing process.

The method of the X-ray CT is the good noninvasing monitoring technique to get the whole gas holdup in the opaque fluidized bed. In the whole fluidized bed, there are two larger symmetric circles of glass beads, in which the gas holdup about 0.46-0.52 is larger than in the middle of the fluidized bed about 0.44-0.46.

4 mm grid resolution is sufficient and excellent to use on this scale reactor study to analyze of the time averaged, time and spatial averaged local gas holdup throughout the glass bead bed in 2D and 3D simulations.

The 3D simulations with Syamlal-O'Brien and Gidaspow drag models better predict the time, time and spatial averaged gas holdup variation throughout the whole fluidized bed, comparing with the experimental data. The results of simulations with Wen-Yu drag model are generally over predicted.

The injector air path extends further into the bed along the axial height from quarter to half of radial long, where the maximum gas holdup is decreasing from 0.9 to 0.6 in the Y-Z slice throughout the fluidized bed with side air injection. But the simulations could not exactly predict the gas path. Because the side air has little effect on the X-Z slice, there remain somewhat symmetric.