Investigation of CO₂ capture using solid sorbents in a fluidized bed reactor: cold flow hydrodynamics

Tingwen Li^{1,2}, Jean-François Dietiker^{1,3}, William Rogers¹, Rupen Panday^{1,4}, Balaji Gopalan^{1,3}, Gregory Breault^{1,4}

1. National Energy Technology Laboratory, Morgantown, WV 26505, USA

2. AECOM, Morgantown, WV 26505, USA

3. West Virginia University Research Corporation, Morgantown, WV 26506, USA

4. REM Engineering Services, Morgantown, WV 26506, USA

Abstract

Both experimental tests and numerical simulations were conducted to investigate the fluidization behavior of a solid CO₂ sorbent with a mean diameter of 100μm and density of about 480 kg/m³, which belongs to Geldart's Group A powder. A carefully designed fluidized bed facility was used to perform a series of experimental tests to study the flow hydrodynamics. Numerical simulations using the two-fluid model indicated that the grid resolution has a significant impact on the bed expansion and bubbling flow behavior. Due to the limited computational resource, no good grid independent results were achieved using the standard models as far as the bed expansion is concerned. In addition, all simulations tended to under-predict the bubble size substantially. Effects of various model settings including both numerical and physical parameters have been investigated with no significant improvement observed. The latest filtered sub-grid drag model was then tested in the numerical simulations. Compared to the standard drag model, the filtered drag model with two markers not only predicted reasonable bed expansion but also yielded realistic bubbling behavior. A grid sensitivity study was conducted for the filtered sub-grid model and its applicability and limitation were discussed.

Key words: computational fluid dynamics, fluidized bed, two fluid model, fine particles, flow hydrodynamics

1. Introduction

There has been increasing concerns on the global climate change caused by the greenhouse gas (GHG) emission of which the major source is energy related [1]. Within the energy sector, CO₂ emissions resulting from the oxidation of carbon in fuels during combustion dominates the total GHG emissions. The capture of carbon dioxide at the point of emission from coal- or gas-burning power plants becomes an attractive route to reducing CO₂ emissions into the atmosphere and is gaining widespread interest [2,3]. Among various technologies, CO₂ capture using solid sorbents is being recognized as a very promising technology and has attracted intense attention from both academic and industrial fields in the last decade [4,5]. Solid sorbent based CO₂ capture offers many advantages such as the potential to reduce energy consumption, improve regeneration, enable greater capacity, and offer good selectivity, as well as ease in handling [6].

Solid sorbents for carbon capture comprise a diverse range of porous, solid-phase materials, including mesoporous silicas, zeolites and metal-organic frameworks [7]. The significant porosity and the large surface area required for gas-solid contact lead to light density and small particle size for solid sorbents. In addition to the development of novel sorbent material with high performance, it is also important to investigate the flow hydrodynamics of solid sorbents in fluidized bed reactors to facilitate the technology development. The solid sorbent particles typically can be classified as Geldart Group A (or A/B) powders. Small particles classified as Group A powders are of great importance in various chemical processes owing to advantages such as ease of fluidization, large surface area and many other superior properties.

Numerous experimental work has been reported to study the fluidization behavior of group A particles. As a complementary tool to experiment, computational fluid dynamics (CFD) has been widely used to investigate gas-solid flow in multiphase systems [8]. So far two major approaches are being used to model the gas-solid fluidization system, i.e. the Eulerian–Eulerian (EE) method (also called the two-fluid model or TFM), and the Lagrangian–Eulerian (LE) method. The former treats both gas and solid phases as interpenetrating continuum with appropriate constitutive correlations. Additional constitutive closures are usually developed based on the kinetic granular theory to model the solids phase as a continuum [9]. The latter treats the gas phase as a continuum, but tracks the solid phase on the particle level by solving the trajectory of each individual particle (e.g. discrete element method) or a swarm of particles (e.g. particle-in-cell method). Both EE and EL models have been applied to model the solid phase in fluidized-bed reactors involving Geldart A particles, with varying success. As discrete element method (DEM) simulations of group A particles are typically limited to extremely small systems, they usually focus on fundamental aspects such as minimum fluidization velocity and the effect of cohesive force [10]. For validation against real fluidized bed systems of fine particles, most simulations were done based on the EE approach.

Despite the massive reported successes in EE modeling of the hydrodynamics in fluidized beds of groups B and D particles, group A particles pose a great challenge. It has been widely reported that the TFM failed to accurately predict the hydrodynamics of Geldart A particle in bubbling fluidized bed. Ferschneider and Mege [11] are about the first to report that the TFM substantially over-predict the bed expansion of Geldart A particle in bubbling fluidized bed. Similar behavior has been observed in many studies [12–15]. It has been generally accepted that the interphase drag force plays the most important role in affecting the bed expansion. To overcome the over-prediction of bed expansion in numerical simulations, several studies have reported adjusting the interphase drag. McKeen and Pugsley proposed to reduce the interphase drag force by a constant factor to match the bed expansion [12]. Zimmermann and Taghipour were able to overcome the over-prediction of bed expansion of FCC fluidized bed by calibrating the drag model based on the measured minimum fluidization velocity [13]. These methods lack generality and do not work for all cases. Some modelers adjusted the drag force by considering the agglomeration effect during fluidization of fine powders [16,17]. Some recent studies suggest that the over-prediction of bed expansion is mainly attributed to the small heterogeneous structures i.e. small bubbles, cluster, or agglomerates which are not resolved in the numerical simulation with a relatively coarse grid [18]. Hence, to overcome this limitation, one has to use sufficiently fine grid and small time step to resolve the heterogeneous structures including steep gradients in field variables with small spatial and temporal scales [19]. Otherwise, one has to account for the effect of unresolved structures through appropriate sub-grid models. The unresolved small structures, mainly in the form of clusters or bubbles depending on solids concentration, can be accounted for through sub-grid models that are developed based on fine-grid simulations [20] or based on an energy minimization multi-scale (EMMS) model [21]. Many sub-grid drag modifications have been developed to account for the effect of small unresolved scales on the resolved mesoscales in dense gas-particle flows [22,20,23]. A comparative analysis of sub-grid drag modifications for dense gas-particle flows in bubbling fluidized beds conducted by Parmentier et al. [24] for coarse grid simulation a bubbling fluidized bed of group B particles showed promising results. Specifically, Wang et al. [25] developed a bubble-structure-dependent drag model by dividing the local heterogeneous flow into three sub-systems, bubble phase, emulsion phase and inter-phase and demonstrated better performance comparing to the conventional drag models for the simulation of bubbling fluidization of group A particles. Overall, the general validity of those models to group A particles needs further study.

The current work focuses on studying the flow hydrodynamics of solid CO₂ sorbents under typical fluidization conditions experimentally and computationally. One type of CO₂ sorbent, called 32D, developed at National Energy Technology Laboratory (NETL) with a mean diameter of around 100µm and density of about 480 kg/m³ was tested in a cylindrical fluidized bed. The system was then modeled by TFM for validation. Careful grid study and systematic sensitivity study were carried out in the simulations. Difficulties in predicting the bed expansion have been encountered and the filtered sub-grid drag model is then adopted. Specifically, the latest filtered sub-grid drag model developed by Sarkar et al. [26] is employed. Validation against the measured pressure drops for different operating conditions are reported.

2. Experimental setup

A fluidized bed system was constructed to investigate the fluidization behavior of CO₂ sorbents. Figure 1(a) shows a photo of the fluidized bed system. The system mainly consists of a 1.8m tall cylindrical column made of plexiglass, 10cm in diameter, and gas plenum and pipelines. Multiple sintered metal porous plates were used as distributor plate to maintain sufficiently high pressure drop and uniform gas flow distribution. The system was designed for both cold flow test and reacting flow test for CO₂ capture. For cold flow test, nitrogen was used to fluidize the bed while a mixture of CO₂ and N₂ was used for the reacting flow tests under both fixed and fluidized bed regimes. A flow controller was used to control the gas flow rate. The experimental facility was grounded to control the accumulation of static electricity.

Instrumentations for pressure drop and temperature measurements were installed. The current study focuses on the cold flow tests, only the experimental instrumentations for pressure measurement are schematically shown in Figure 1(b). Three pressure drops were measured for the system. PDT 850 measured the pressure drop between 10.16 cm and 20.26 cm above the distributor. PDT 812 measured the pressure drop between 4.46cm and 10.16cm above the distributor. The pressure drop through the distributor plate to 4.46cm above it was measured by PDT 820. Most pressure data were measured using the Rosemount pressure transmitters with 1Hz data acquisition rate. The gauge pressure in the plenum was measured by PT810 which characterizes the overall pressure drop through the distributor and bed. For some tests, a high speed Setra pressure transmitter with 1000Hz data acquisition rate was also used at PDT 812 for comparison with the Rosemount pressure transmitters and possible analysis of pressure signals. However, the high frequency pressure data were not used in the current study for model validation. In addition to the measurement through installed instrumentations, video of the system was also taken to allow image analysis to extract the bed expansion and qualitative observation.

The minimum fluidization velocity was measured in this system which is consistent with the value determined from the characterization procedure. The matrix for cold flow hydrodynamic tests of 32D sorbent consisted of three bed heights of 10, 15 and 20 cm and four gas velocities of 1, 3, 5 and 7 U_{mf}. Seven repeats were conducted for tests with a gas velocity greater than the minimum fluidization velocity. While for tests with minimum fluidization gas velocity, only four repeats were conducted.

Figure 1.

NETL 32D Sorbent

The solid CO₂ sorbent investigated in the current study is NETL-32D which is an amine-impregnated mesoporous sorbent using the polyethylenimine (PEI), N1-(3-trimethoxysilypropyl) diethylenediamine (TMPED) and silica gel (PQ Corporation CS2129). The silica substrate was obtained from PQ Inc. with an average particle size of 100 microns, a surface area of 280 m²/g, and pore volume of 2.4 mL/g [27]. A picture of the sorbent particles is shown in Figure 2. The detailed procedure of the immobilization process can be found in [28,29]. The stability and CO₂ capture performance of the final immobilized sorbents have been tested by both fixed-bed reactor and thermogravimetric methods [30].

Figure 2.

To characterize the physical properties of 32D sorbents, the methodology developed by Tucker et al. [31] was used which combines historical methods, industry practice, and standards from several international organizations. The particle size distribution was characterized as shown in Figure 3. The particle size ranges from 20µm to about 200µm and the mean particle size is about 100µm. Other physical properties including particle shape, density (particle, skeletal and bulk), and porosity were characterized. The minimum fluidization velocity was measured in a 6.35cm diameter fluidized bed by measuring the pressure drop when varying the gas flow rate. The details of the procedure of particle characterization considered in the present study are reported in Tucker et al. [31]. Table 1 summarizes the physical properties of 32D sorbents used in the experimental test. The particle has a porosity of 0.71, which leads to very low particle density of 480kg/m³. According to the Geldart's classification of particles, the 32D sorbent belongs to Group A.

Figure 3.

Table 1.

3. Numerical model

The open-source code MFIX was used to simulate the experimental system. The governing equations solved in the simulations are summarized in Table 2. Mass and momentum conservation equations were solved for the gas and solids (particulate) phases, with appropriate closure relations [32,33]. Constitutive relations derived from granular kinetic theory were used for the solids phase [9,34]. More information on MFIX, as well as detailed documentation on the model equations and the numerical implementation, can be found on the MFIX website, <https://mfix.netl.doe.gov>. In most simulations, equations based on the standard kinetic granular

theory with an algebraic form of granular temperature were solved. The Schaeffer friction model was used for the frictional regime of dense flow [35]. The Gidaspow drag model was used for the interphase momentum exchange [9].

Table 2.

Cartesian grid was used in the simulation with uniform grid size in all directions. Both 2D and 3D simulations were carried out. The 2D simulation was simply a vertical slice of the column. For the 3D simulation of the cylindrical column, the Cartesian grid was used with the cut-cell method to resolve the cylindrical wall [36]. The simulation was initialized with a static bed height and a uniform gas flow was then fed through the bottom distributor. A no-slip wall boundary condition was used for both gas and solid phases. For most simulations, a second-order superbee scheme was used for discretizing the governing equations and the implicit first order Euler scheme with variable time step was used for the temporal discretization.

The default parameters used in the simulation are summarized in Table 3. Most parameters are based on the particle characterization. For properties which were not available from the measurement, values based on the literature were used [37]. A necessary sensitivity study was conducted for parameters that are of great uncertainties.

Table 3.

4. Experimental results

This section summarizes the experimental observation and measurements. One unique feature of Group A powders is that the bed expands homogeneously before reaching the minimum bubbling velocity, U_{mb} , the velocity at which the first bubble appears. This behavior was observed in the current experimental system with 32D sorbents as the gas velocity was gradually increased from minimum fluidization velocity to higher value. Vigorous bubbling was observed to start at a gas velocity of about $4 U_{mf}$. In the original design, one sintered metal porous plate was used as the distributor. To make sure the gas flow through the distributor was uniform, the pressure drop through the distributor was estimated from the measurement of PDT 820 and PDT 812, which are 83, 265, 447, and 635 Pa for the gas velocity of 1, 3, 5 and $7U_{mf}$, respectively. Compared to the pressure drop due to bed weight, for example, 423 Pa for the 15 cm static bed height, the pressure drop through the distributor is believed sufficient to maintain uniform gas flow for most tests [38]. However, the flow uniformity might be questionable for low gas velocities of 1 and $3U_{mf}$, especially for the high static bed height of 20 cm. To guarantee the

uniformity of gas flow distribution mainly for low flow rate tests, four sintered metal porous plates were stacked to further increase the pressure drop through the distributor.

Figure 4 shows a series of images taken from the system under different superficial gas velocities. The static bed height was 15 cm. As gas velocity was increased from $1U_{mf}$ to $2U_{mf}$, the bed exhibited almost homogenous expansion with only tiny bubbles visible at the bed surface and the bed surface remained stationary. At $3U_{mf}$, the bed further expanded with large bubble occasionally observed. The bed heights can be easily measured which are 16.5 cm and 19 cm for $2U_{mf}$ and $3U_{mf}$. At $4U_{mf}$, the bed further expanded. Some large bubbles started bursting and the surface began to wave as shown in

Figure 4(d). Further increasing to $5U_{mf}$ and $7U_{mf}$ led to vigorous bubble bursting at the bed surface which resulted in strong fluctuations and made the bed height measurement difficult. As the bed was vigorously bubbling, the expanded bed height decreased slightly compared to that of $4U_{mf}$. In addition, large bubbles with size about 0.5~1 cm can be observed rising along the wall which was not seen for low gas velocities.

Figure 4.

Table 4 summarizes the pressure drops measured by PDT 812 between 4.46 and 10.16 cm above the distributor for different operating conditions. The mean pressure drops are averages of 7 repeats. The standard deviation was also calculated and is shown in the table. From the pressure drop measurements, it can be seen that the pressure drop slightly increases with the static bed height which might be attributed to the slight drop in gas velocity due to hydrostatic pressure. For different superficial gas velocities, the pressure drop varies marginally and no consistent trend can be observed except for the tests with $1U_{mf}$. For experimental runs at minimum fluidization velocity, the pressure drop shows considerable variation with the static bed height and is lower than that at higher flow rates. One possible reason would be a weak gas by-passing takes place in the system at the incipient of fluidization, which tends to mitigate with higher static bed height and gas velocity. It should be noted that the changes in pressure drop for different operating conditions are quite small. Overall, the pressure drop measured by PDT 850 does not show much sensitivity to the operating conditions tested in the current study.

Table 4.

Table 5 reports the pressure drops measured by PDT850 between 10.16 and 20.26 cm above the distributor for different operating conditions. Due to the relative position of the top pressure

probe with respect to the bed surface, the pressure drop measurements are very different for different static bed height. For a static bed of 10 cm at $1U_{mf}$, both probes are above the bed, and no pressure drop was recorded. Besides that, no consistent trend can be observed for different gas velocities. The pressure drop measurements tend to suggest that the solid holdup does not change much for the velocities tested. For the static bed height of 20cm, all pressure probes are immersed in the bed, the pressure gradients calculated from PDT 812 and PDT 850 are very close to each other, suggesting the axial distribution of solid holdup is quite uniform.

Table 5.

5. Simulation results

Both 2D and 3D simulations were conducted in the current study. The 2D simulation was largely relied on in the current study for qualitative investigation on some key factors affecting the simulation results. Only a few 3D simulations were conducted to verify the conclusion drawn from the 2D simulations when possible.

Grid study

The suggested rule-of-thumb to obtain grid convergence is a grid size about 10-particle-diameter for simulating the gas-solid flows in circulating fluidized bed risers [20,39,40]. For bubbling fluidized beds of group B particles, several studies reported a slightly smaller grid size for grid independent results. For example, a grid size of 5~6 particle diameter was found to be sufficient for 221 μm alumina particles in a 5 cm diameter fluidized bed [41]. Wang et al. [18] investigated the effect the grid resolution on the bed expansion of Geldart A particle of 75 μm diameter and density of 1290 kg/m^3 . They found that a grid size of 2 to 4 particle diameter and a time step of $1\text{e-}6$ s were needed for the grid independent result of bed expansion to match the DEM simulation. For a micro-fluidized bed reaction of FCC particles with a mean diameter of 53 μm and density of 1400 kg/m^3 , Liu et al.'s [15] grid study reported that a grid size of 1 particle diameter was needed to reach the grid convergence. Cloete et al. [42] studied the effect of particle size on grid independence and found it unexpectedly large. According to their study, the ratio between grid size and particle size for grid independent simulation decreases considerably as the particle size is reduced. All these studies tend to indicate that a grid independent simulation result requires extremely fine grid for Group A particles. For the 32D particles with quite different density from most particles reported in the literature, a systematic grid study was deemed necessary.

The experimental test with a static bed height of 15 cm and a superficial gas velocity of $5U_{mf}$ was first simulated in the grid study. The grid sizes of 2, 1, 0.5, 0.25 mm were used for 2D simulations which corresponded to 20, 10, 5 and 2.5 times of the particle diameter. All simulations were conducted for 40 s and the results in the fully developed state were analyzed.

Figure 5 shows snapshots of voidage distribution inside the bed at the end of simulation ie. 40s predicted by different grid resolutions and mean voidage distribution averaged from 20 to 40s. The figure clearly demonstrates the strong effect of grid size on the flow hydrodynamics. Homogeneous expansion tends to be predicted by the grid sizes of 2 and 1 mm with only a few small bubbles existing in the system. More and more bubbles are predicted as the grid size is reduced to 0.5 and 0.25 mm. Because of the existence of bubbles, the gas tends to bypass through the bubbles and lead to denser emulsion phase and lower bed height with grid refinement.

Figure 5.

As there exist considerable quantitative differences between 2D and 3D simulations [43–47], it is necessary to confirm if the 3D simulations yield a similar grid sensitivity. Two grid sizes of 2 mm and 1.25 mm which correspond to 20 and 12.5 times of particle diameter were used in the 3D simulation. Figure 6 shows a transient snapshot of the voidage distribution and time-averaged over 30s along a vertical slice of the simulation with a grid size of 1.25 mm. The results by 2mm grid are very similar hence not shown here. Consistent with the results obtained in the 2D simulations using 2mm and 1mm grids, the bed tends to expand homogeneously and no obvious bubbling behavior can be seen.

Figure 6.

It is too computationally expensive to conduct the 3D simulation with further grid refinement. To reduce the computational cost, a small system with dimensions of 5×5×30 cm was simulated with the same static bed height and superficial gas velocity. The small system was discretized use the 0.5 mm grid to verify the bubble formation in 3D simulations. Figure 7(a) shows the voidage distribution along a vertical slice. More heterogeneous flow is predicted compared to that shown in Figure 6(a) with a coarser grid. Small bubbles can be tracked by rendering the iso-surface of voidage at 0.8 as shown in Figure 7(b). The predicted bubble behaviors are similar to the 2D simulations with no apparent qualitative difference observed.

Figure 7.

The bed height can be extracted from time averaged results for different grid resolutions by measuring the location of transition from dense bed to the free board near wall, specifically a

value of 0.8 was used for the mean voidage. The extracted bed heights were insensitive to this value due to the sharp concentration gradient at the bed surface. Figure 8 plots the bed heights predicted by different grid sizes for both 2D and 3D simulations. For the simplified 3D simulation with 0.5mm grid, the bed height is shown as an empty symbol in Figure 8 by assuming a negligible effect of cross-sectional area on bed expansion. As the grid is refined, the bed height decreases for both 2D and 3D simulations. Compared to the 2D simulations, the 3D simulations with similar grid size predicted slight higher bed expansions. This is contrary to the difference reported in the literature for bubbling fluidized bed of group B particles in which the 3D simulations usually predicted lower bed height than 2D simulation [41,45,47,48]. Compared to the bed height measured in the experiment which is about 18.7 cm, all simulations over-predict the bed expansion. With grid refinement, the prediction of bed expansion approaches the experimental measurement. However, no grid independent behavior can be observed and the grid study indicates further grid refinement is needed. However, it requires at least two levels of refinement to confirm the grid convergence. As the grid is refined by a factor of 2 the time step reduces as well which further increases the computational cost. The computational cost increased substantially due to the increased cell count, reduced time step, and degradation in parallel performance. For example, it took about 5 hours to complete the simulation with 1 mm grid on 16 CPU cores while it took 15 hours to complete the simulation with 0.5 mm on 64 CPU cores.

Figure 8.

The bubble properties have a significant effect on the overall performance of fluidized bed reactor as the bubble behaviors affect the solids movement and gas-solid contact. Correct prediction of bubbles is critical in modeling the fluidized bed reactors. In numerical simulations, many small bubbles were predicted which typically followed the bulk solids motion. Not much bubble coalescence and breakup can be observed in the simulation which is inconsistent with the known bubbling behavior of typical group A particles. The bubble size information was extracted from the simulation using a threshold voidage of 0.8. Detailed information on the procedure for image processing and bubble statistics can be found in [49]. Figure 9 shows the variation of bubble size along the height predicted by 2D simulations with different grid sizes. The error bar represents the standard deviation for all bubbles extracted at that height. No bubble information was extracted from the 3D simulations as a homogeneous expansion with rare bubbling was predicted due to insufficient grid resolution. The mean bubble size is almost constant along the height. This is consistent with the observation that not much bubble coalescence was predicted. The mean bubble size decreases slightly as the grid is refined, supported by the snapshots shown in Figure 5 where more small bubbles are predicted especially from 0.5 mm to 0.25 mm. For the simulation with a grid size of 1 mm, only 1051 bubbles were tracked in 400 images of 20 s simulation which leads to the large statistical error in the mean bubble diameter. The numbers of bubbles tracked in the simulations of 0.5 mm and 0.25 mm grid were 38779 and 139014, respectively. Clearly, the smaller grid used in the simulation the more bubbles were predicted.

There was no bubble size measurement available from the current experimental system. For comparison, the correlation shown in equation (29) proposed by Darton et al. [50] was used to estimate the bubble size at different heights, which is shown in Figure 9. Compared to the correlation, the 2D simulations fail to capture the bubble growth along the height due to hydrostatic pressure change and bubble coalescence. Hence, the bubble size has been greatly under-predicted noting that a different scale is used for the Darton et al.'s correlation (shown on the right axis). In addition, bubbles with size around 1 cm range were observed from the video captured on the acrylic side wall which confirms the under-prediction of bubble size in 2D simulations.

$$d_b = 0.54 \left(U_g - U_{mf} \right)^{0.4} h^{0.8} / g^{0.2} \quad (29)$$

Figure 9.

Comprehensive sensitivity study

So far, no evidence of grid independent behavior has been observed for both bed expansion and bubble sizes and the simulations over-predict the bed expansion and under-predict the bubble size. A comprehensive sensitivity study has been conducted covering various numerical parameters, physical properties, and constitutive sub-models to seek improvement. The following analyses are based on 2D simulations using the grid size of 0.25 mm.

The critical effect of frictional model in modeling bubbling fluidized bed has been reported in several studies [51,52]. Two frictional models available in MFIx, the Schaeffer model [35] and Srivastava & Sundaresan model, referred as Princeton friction model were tested [53]. To fully take advantage the Srivastava & Sundaresan friction model, the full PDE for the granular temperature was solved instead of the algebraic form of granular temperature used in the grid study. Many models are available in the literature for the interphase momentum exchange in gas-solid systems. The differences caused by different drag models for fluidized bed simulations have been widely reported [13,54,55]. Two of the most widely used standard drag models, the Gidaspow drag model and the Syamlal and O'Brien drag models are compared. The Syamlal and O'Brien drag model was calibrated based on the measured minimum fluidization velocity of the 32D particles [56]. In addition to physical sub-models, various numerical parameters including time step, convergence tolerance, and numerical schemes were tested. It was found that the numerical scheme showed the most impact on the simulation results. Although two additional schemes were tested (minmod and van Leer), only the minmod results are reported here as they showed the most deviation from the superbee scheme results.

The main results of bed heights and bubble properties obtained from the sensitivity study are summarized Table 6 and Figure 10. Taking the results of ASGS as a baseline, the Princeton frictional model has a negligible effect on the bed expansion but slightly increase the bubble size. The difference in bed height introduced by the numerical scheme is not negligible compared to that by different friction models and drag models. The bubble size is less sensitive to the numerical scheme but the number of bubbles predicted is affected by the choice of numerical scheme. The calibrated Syamlal and O'Brien drag model predicted lower bed expansion and slightly larger bubbles and less bubble count. Compared to the experimental measurement of bed height, the calibrated Syamlal and O'Brien drag produces the best agreement. However, without confirmation of the grid independence, it is hard to conclude which drag model is better.

Table 6.

Figure 10.

In addition to the sensitivity studies presented above, parametric studies were also conducted for restitution coefficient, frictional coefficient, specular coefficient for wall boundary condition. None of them showed significant impact on the simulation results as far as the bed expansion and bubble sizes are concerned. The results are not shown here for brevity.

Filtered model simulations

Due to the lack of grid convergence at the finest grid of the current study, a sub-grid model accounting for the unresolved heterogeneous flow structures with the relatively coarse grid is used. Two types of sub-grid models have been developed based on an energy minimization multi-scale (EMMS) model [22,57] or coarse grid filtering of the fine-grid simulations [58,59]. Here, the later model was adopted for which highly resolved TFM simulations of the gas-particle flow of Geldart type A particles in a periodic domain were conducted and results were filtered to deduce correlations for the corresponding filtered TFM equations. The filtered drag model can be expressed as

$$\vec{F} = (1 - H_d) \beta (\vec{V}_g - \vec{V}_s) \quad (30)$$

where $\vec{V}_g$ and $\vec{V}_s$ are the filtered velocities of gas and solid which are the resolved velocities in the coarse grid simulation, β denotes the homogenous drag correlation computed from filtered variables for which the Wen & Yu drag model was typically used in developing the filtered model, and H_d is inhomogeneous index accounting for the unresolved sub-grid structures.

It should be noted that development of such filtered models requires a large number of simulations with extremely fine grid resolution. Hence, the initial efforts mainly relied on the highly resolved 2D simulations of a periodic domain [59,60]. Recently, the effort has been made to develop an accurate filtered model to support modeling the carbon capture process using solid CO₂ sorbents. To improve the accuracy in coarse grid simulations, new filtered constitutive closures have been developed for gas-solids flow from highly resolved 3D simulations. Compared to the filtered model developed mainly from 2D simulations, the latest filtered model accounts for the inherent three-dimensionality of gas-solid flow and is believed more realistic. Furthermore, the high resolution 3D simulations were based on the typical properties of CO₂ sorbents with a diameter of 150 μ m and density of 441 kg/m³ which makes it more appropriate for the 32D sorbents considered in the current study.

In addition to the latest filtered model by Sarkar et al., the filtered model developed by Igci et al. [59] was also tested for comparison. The filtered drag models are summarized in Table 7. The Igci et al. filtered model was mainly developed from 2D simulations with a correction factor accounting for the 3D effect. In Igci et al. model, the filtered drag force is a function of filter size and filtered solids volume fraction which are used as markers to classify the inhomogeneity of sub-filter scale structures. The Sarkar et al. model was completely developed from high resolution 3D simulations. In addition, the filtered drag model includes the filtered gas-solid slip velocity as an important additional marker as suggested in previous studies [23,24,60]. Generally, for the same filtered solid concentration, higher slip velocity suggests higher inhomogeneity hence further drag reduction.

Table 7.

The experimental test of 32D sorbents with a static bed height of 15cm and superficial gas velocity of $5U_{mf}$ was simulated to compare the performance of different filtered drag models. The 3D computational domain was discretized using a uniform grid size of 2 mm. Figure 11 shows some transient snapshots of voidage along a vertical slice by using the Gidaspow drag, the Igci et al. filtered drag, and Sarkar et al. filtered drag. For the simulation using Gidaspow drag model, the bed expands almost homogeneously with no apparent bubble observed. The bed expansion has been substantially decreased when the Igci et al filtered drag model was used. Heterogeneous flow can be clearly seen from the simulation results. However, bubbles with high voidage are barely seen from the simulation. The bed expansion is further reduced by the Sarkar et al. filtered model and bubbling behaviors can be clearly observed.

Figure 11.

To further illustrate the differences among these drag models, bubbles predicted by different simulation are shown in Figure 12 where the iso-surface of voidage of 0.8 is used to define the bubble surface. As can be seen that for the Gidapow drag and Igci filtered drag models, no bubble can be seen inside the bed. While the Sarkar et al. filtered drag model predicts reasonable bubbling behavior inside the bed which is more consistent with the experimental observation. It is believed the one marker approach tends to smear the heterogeneous structure and even though the bed expansion is reasonably predicted. Similar behavior has been reported by Parmentier et al. [24] when comparing different sub-grid drag models. The difference can also be observed in the bed surface from the iso-surface plot which corresponds to the different bubbling behaviors predicted numerically.

Figure 12.

The bed heights are compared in Table 8 for different drag models as well as the experimental measurement. Clearly the unfiltered model over-predicts the bed expansion by almost 40%. Both filtered drag models provide an improved agreement to the measured bed height. The Igci et al. filtered drag model slightly over-predicts the bed expansion while the Sarkar et al. filtered drag model somehow under-predict it. As far as the bubbling behavior is concerned, the Sarkar et al. filter model is believed to compare the most favorably to the experimental observation.

Table 8.

In filtered models, the effective drag is a function of solid concentration, slip velocity as well as the filtered size which is proportional to the numerical grid size to reveal grid independent solutions. A grid study was conducted for the 3D simulations using the Sarkar et al. filtered drag model. In addition to the simulation with 2 mm grid reported above, simulations with grid sizes of 4 mm and 1.25 mm were performed. The transient results were first examined which indicated the bed expansion is not very sensitive to the grid size. The bed heights predicted by 4 mm, 2 mm and 1.25 mm grids are 18 cm, 17.1 cm, and 16.8 cm, respectively, as shown in Figure 8. On the other hand, the prediction of bubble behavior was found to be quite sensitive to the grid resolution as shown in

Figure 13. Clearly, the finer the grid is, the more bubbles are predicted by the numerical simulation.

Figure 13.

Figure 14 compares the mean bubble diameter along the height predicted by different grid sizes. It is quite difficult to extract the 3D bubble information from the simulation mainly because of the complex bubble interactions and movements. Here the mean bubble size was characterized by extracting multiple vertical slices from the 3D simulation then measuring the properties of 2D bubbles. For simulation with a grid size of 4 mm, only a small amount of bubbles were predicted which leads to the significant variation and large standard deviation associated with the mean bubble size. Similar to the results using the unfiltered drag model, the bubble size is affected by the grid size which decreases as the grid is refined. The increase of bubble size with height reduces as the fine grid is used as more and more small bubbles are resolved. In the current analysis, all bubbles are counted. It should be noted that some small bubbles predicted by the fine grid are accounted for by the sub-grid model in the coarse grid simulation. Hence, a better way to quantify the grid effect on the bubble properties should be used for systematic analysis. It should be noted that the bubble size extracted with this approach tends to bias toward a smaller value comparing to the true bubble size as the bubble center is rarely in the slice. A correction factor of $\sqrt{3/2} = 1.2247$ is used to account for this by assuming spherical bubbles when comparing to the real bubble size measured experimentally or predicted by empirical correlations. Figure 15 compares the bubble size predicted by different grid sizes after correction with the correlation by [50] (the same scale is used for both). Compared to the empirical correlation for bubble size, the coarse grid simulations with filtered drag yield a better agreement in contrast to the results shown in Figure 9.

Figure 14.

Figure 15.

So far, only the filtered sub-grid drag model was used in the simulation. Igci and Sundaresan [59] reported a significant contribution to the filtered solid stresses arising from subfilter-scale Reynolds-stress-like velocity fluctuations. A similar filtered model was developed by Sarkar et al. [26] which was incorporated in the later stage of the current study. The filtered solid stress models were tested for the simulations with 2mm grid. The bed heights are 19.5 cm and 17 cm for Igci et al. and Sarkar et al, respectively. Compared to the experimental measurement, the inclusion of the filtered solids stress improves the numerical prediction for Igci et al model. As far as the bubbling behavior is concerned, homogeneous expansion was still predicted with no visible bubble. Only negligible effect can be observed for the Sarkar et al. model after inclusion of the filtered solid stress for the current system.

A wall correction for filtered model was proposed by Igci and Sundaresan [61] for riser flows which further reduces the filtered drag, solids pressure and viscosity based on the distance from

the wall. They have shown that the inclusion of wall correction affected the results predicted by the filtered model simulation quantitatively and qualitatively and was necessary for capturing the experimental data [62]. By including the wall correction in the current simulations, the bed expansion was found to decrease to 16.2 cm. Bubbling behavior was observed which was absent without the wall correction. In this regard, it is suggested that the model implicitly accounts for some wall effect by including the filtered slip velocity as an additional marker [60], hence no wall correction was used in Sarkar et al. filtered model.

In this part of work, we focused on the filtered drag model as it is believed the most critical factor affecting the flow hydrodynamics of bubbling fluidized bed. By comparing the coarse grid simulations with and without filtered sub-grid drag model, the poor performance of standard TFM can be attributed to the over-prediction of interphase drag force. With a 2 mm numerical grid, the simulation with the latest filtered drag model yielded much improved results both qualitatively and quantitatively compared to the standard model and the previous filtered model. Reasonable bed expansion, pressure drop and realistic bubbling behavior were obtained in the simulation which showed a promising comparison with the experimental measurement and observation.

Two reasons for the over-estimation of gas-solid drag force for group A particles have been reported in the literature. One argument is that the strong cohesive force, i.e. van der Waals force, between group A particles forms stable or unstable agglomerate which effectively reduce the drag force [17]. Others claimed that the unresolved small scale structures play a major role in the over-estimation [18]. In coarse grid simulations, both agglomerate and cluster effectively reduce the interphase drag force. The improved results of coarse-grid simulations with the filtered sub-grid drag model indicate the dominant effect of meso-scale structure in drag reduction. On the other hand, the fine grid simulations using standard drag model predicted somehow unrealistic small bubbles compared to the experimental observation. This tends to suggest that the effect of cohesive force should be considered in the fine grid simulation. Further study is needed for confirmation of this speculation. It should be noted that the particle size distribution and the non-spherical shape are not accounted for in the current study. However, they might have considerable impact on the flow hydrodynamics in bubbling fluidized bed which needs attention in the future modeling study.

Comparison with experimental data

Finally, cases of different static bed heights and superficial gas velocities were simulated using the Sarkar et al. filtered drag model. For all simulations, the pressure drop data were extracted from simulation when the flow reaches fully developed state. Specifically, the mean pressure drops between 4.46 and 10.16 cm above the distributor and between 10.16 and 20.26 cm above the distributor were compared against the experimental measurements.

Table 9 summarizes the results of 15 and 20 cm bed heights under three superficial gas velocities of 3, 5 and $7U_{mf}$ where numerical predictions and experimental measurements of pressure drop at different locations are compared. The simulation slightly over-predicts the pressure drop for all cases which is consistent with the tendency of under-predicting the bed height reported above. For different superficial gas velocities, no significant change in the pressure drop is predicted which is in line with the experimental measurements. Overall, the simulations using Sarkar et al. filtered drag model yield reasonable agreement to the experimental data with relative errors less than 10%.

Table 9.

From the pressure drop measurements of PDT812, the solid holdup does not change much for different bed heights and gas velocities. However, differences in the bubbling behavior were observed in the experiment.

Figure 16 shows the bubble size information predicted by the numerical simulations for different operating conditions. As can be seen for the low gas velocity of $3U_{mf}$, the bubble size is smaller than those for higher gas velocities of $5U_{mf}$ and $7U_{mf}$ for both 15 cm and 20 cm static bed heights. No significant differences in bubble size between $5U_{mf}$ and $7U_{mf}$ is predicted. It should be noted as the gas flow is increased, the bubble count predicted in the simulation increases substantially. For 1200 snapshots of vertical slice extract from 30s simulation results for the static bed height of 15 cm, the total numbers of bubbles are 161, 2448 and 4529 for 3, 5 and $7U_{mf}$, respectively. The small amount of bubbles predicted for the case with $3U_{mf}$ leads to a large statistical error in the mean bubble diameter calculation though it is consistent with the visual observation of a limited number of bubbles in the experiment. Overall, for both bed heights, increasing the superficial gas velocity from $5U_{mf}$ to $7U_{mf}$ mainly increases the number of bubbles in the system and has only small impact on the mean bubble size.

Figure 16.

6. Summary and future work

In this study, a carefully designed experimental work was conducted for a light solid sorbent to understand the flow hydrodynamics for CO_2 capture in a fluidized bed reactor. The experimental results were used for validating the CFD model. The NETL 32D solid sorbents tested belong to Geldart Group A which poses great challenges to the CFD modeling. Numerical simulation of the experimental system has been carried out based on the two-fluid model approach. A systematic grid study was performed in both 2D and 3D simulations. The over-prediction of bed

expansion as reported in the literature was observed. Grid refinement did lead to reduced bed expansion. However, no good grid independent results have been reached. In addition, unrealistic small bubbles were predicted by the numerical simulation which was against the experimental observation and some empirical correlations for the bubble size. A sensitivity study on a grid size of 2.5 times of the particle diameter indicated that the bed expansion is affected by drag model and frictional model and numerical schemes. However, no qualitative improvement was observed as far as the bubble properties were concerned. This suggests that some physics is missing in the TFM for the light group A particles which require further fundamental investigation.

The recently developed filtered sub-grid model by Sarkar et al. [26] was employed to conduct relatively coarse-grid simulations of this system. The bed heights and pressure drops of different operating conditions, i.e. static bed height and superficial gas velocities were predicted with a reasonable agreement with the experimental measurement. Comparison of the simulations with standard drag model and previous filtered sub-grid drag model were made. The latest filtered sub-grid model yields superior qualitative and quantitative agreement with the experimental measurement and observation. The filtered model does show very promising results compared to the experimental data for the light group A particles under investigation using a relatively coarse grid. This study is the first part of an effort to simulate the CO₂ capture using solid sorbents in fluidized bed reactors. Given the reasonable confidence from the current modeling exercise, the lessons learnt will be applied to the reacting flow modeling and further results will be reported in the near future.

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References

- [1] IEA, CO₂ Capture and storage: A key carbon abatement option, Agency International Energy, 2008.
https://www.iea.org/publications/freepublications/publication/CCS_2008.pdf.
- [2] D.Y.C. Leung, G. Caramanna, M.M. Maroto-Valer, An overview of current status of carbon dioxide capture and storage technologies, *Renew. Sustain. Energy Rev.* 39 (2014) 426–443. doi:10.1016/j.rser.2014.07.093.
- [3] M. Wang, a. Lawal, P. Stephenson, J. Sidders, C. Ramshaw, Post-combustion CO₂ capture with chemical absorption: A state-of-the-art review, *Chem. Eng. Res. Des.* 89 (2011) 1609–1624. doi:10.1016/j.cherd.2010.11.005.
- [4] J. Wang, L. Huang, R. Yang, Z. Zhang, J. Wu, Y. Gao, et al., Recent advances in solid sorbents for CO₂ capture and new development trends, *Energy Environ. Sci.* 7 (2014) 3478–3518. doi:10.1039/C4EE01647E.
- [5] A. Samanta, A. Zhao, G.K.H. Shimizu, P. Sarkar, R. Gupta, Post-Combustion CO₂ Capture Using Solid Sorbents: A Review, *Ind. Eng. Chem. Res.* 51 (2012) 1438–1463. doi:10.1021/ie200686q.
- [6] S. Sjoström, H. Krutka, Evaluation of solid sorbents as a retrofit technology for CO₂ capture, *Fuel*. 89 (2010) 1298–1306. doi:10.1016/j.fuel.2009.11.019.
- [7] A.-H. Lu, S. Dai, eds., *Porous Materials for Carbon Dioxide Capture*, Springer, 2014.
<http://www.springer.com/us/book/9783642546457> (accessed February 17, 2016).
- [8] J.R. Grace, T. Li, Complementarity of CFD, experimentation and reactor models for solving challenging fluidization problems, *Particuology*. 8 (2010) 498–500. doi:10.1016/j.partic.2010.09.003.
- [9] D. Gidaspow, *Multiphase Flow and Fluidization: Continuum and Kinetic Theory Descriptions*, Academic Press, 1994.

- [10] J.K. Pandit, X.S. Wang, M.J. Rhodes, On Geldart Group A behaviour in fluidized beds with and without cohesive interparticle forces : A DEM study, 164 (2006) 130–138. doi:10.1016/j.powtec.2006.03.007.
- [11] G. Ferschneider, P. Mege, Eulerian Simulation of Dense Phase Fluidized Beds, Oil Gas Sci. Technol. 51 (1996) 301–307. doi:10.2516/ogst:1996026.
- [12] T. McKeen, T. Pugsley, Simulation and experimental validation of a freely bubbling bed of FCC catalyst, Powder Technol. 129 (2003) 139–152. doi:10.1016/S0032-5910(02)00294-2.
- [13] S. Zimmermann, F. Taghipour, CFD Modeling of the Hydrodynamics and Reaction Kinetics of FCC Fluidized-Bed Reactors, Ind. Eng. Chem. Res. 44 (2005) 9818–9827. doi:10.1021/ie050490+.
- [14] T. Li, K. Pougatch, M. Salcudean, D. Grecov, Numerical simulation of horizontal jet penetration in a three-dimensional fluidized bed, Powder Technol. 184 (2008) 89–99. doi:10.1016/j.powtec.2007.08.007.
- [15] X. Liu, C. Zhu, S. Geng, M. Yao, J. Zhan, G. Xu, Two-fluid modeling of Geldart A particles in gas–solid micro-fluidized beds, Particuology. 21 (2015) 118–127. doi:10.1016/j.partic.2014.05.012.
- [16] B. van Wachem, S. Sasic, Derivation, simulation and validation of a cohesive particle flow CFD model, AIChE J. 54 (2008) 9–19. doi:10.1002/aic.11335.
- [17] A.H.A. Motlagh, J.R. Grace, M. Salcudean, C.M. Hrenya, New structure-based model for Eulerian simulation of hydrodynamics in gas–solid fluidized beds of Geldart group “A” particles, Chem. Eng. Sci. 120 (2014) 22–36. doi:10.1016/j.ces.2014.08.042.
- [18] J. Wang, M. Vanderhoef, J. Kuipers, Why the two-fluid model fails to predict the bed expansion characteristics of Geldart A particles in gas-fluidized beds: A tentative answer, Chem. Eng. Sci. 64 (2009) 622–625. doi:10.1016/j.ces.2008.09.028.
- [19] W.D. Fullmer, C.M. Hrenya, Quantitative assessment of fine-grid kinetic-theory-based predictions of mean-slip in unbounded fluidization, AIChE J. (2015) n/a–n/a. doi:10.1002/aic.15052.

- [20] Y. Igci, A.T. Andrews IV, S. Sundaresan, S. Pannala, T. O'Brien, Filtered two-fluid models for fluidized gas-particle suspensions, *AIChE J.* 54 (2008) 1431–1448. doi:10.1002/aic.11481.
- [21] J. Li, The EMMS model — its application, development and updated concepts, *Chem. Eng. Sci.* 54 (1999) 5409–5425. doi:10.1016/S0009-2509(99)00274-2.
- [22] W. Wang, J. Li, Simulation of gas–solid two-phase flow by a multi-scale CFD approach—of the EMMS model to the sub-grid level, *Chem. Eng. Sci.* 62 (2007) 208–231. doi:10.1016/j.ces.2006.08.017.
- [23] J.-F. Parmentier, O. Simonin, O. Delsart, A functional subgrid drift velocity model for filtered drag prediction in dense fluidized bed, *AIChE J.* 58 (2012) 1084–1098. doi:10.1002/aic.12647.
- [24] S. Schneiderbauer, S. Puttinger, S. Pirker, Comparative analysis of subgrid drag modifications for dense gas-particle flows in bubbling fluidized beds, *AIChE J.* 59 (2013) 4077–4099. doi:10.1002/aic.14155.
- [25] S. Wang, J. Chen, Q. Wang, G. Liu, H. Lu, L. Sun, Evaluation of a bubble-structure dependent drag model for the simulation of bubbling fluidization with Geldart A particles, *Powder Technol.* 289 (2016) 44–51. doi:10.1016/j.powtec.2015.11.024.
- [26] A. Sarkar, F.E. Milioli, S. Ozarkar, T. Li, X. Sun, S. Sundaresan, Filtered sub-grid constitutive models for fluidized gas-particle flows constructed from 3-D simulations, in preparation, *Chem. Eng. Sci.* 152 (2016) 443–456. doi:10.1016/j.ces.2016.06.023.
- [27] A. Lee, D. Mebane, D. Fauth, D. Miller, A model for the adsorption kinetics of CO₂ on amine-impregnated mesoporous sorbents in the presence of water, in: 28th Int. Pittsburgh Coal Conf., Pittsburgh, PA, 2011.
- [28] D.J. Fauth, M.L. Gray, H.W. Pennline, H.M. Krutka, S. Sjostrom, a. M. Ault, Investigation of porous silica supported mixed-amine sorbents for post-combustion CO₂ capture, *Energy and Fuels.* 26 (2012) 2483–2496. doi:10.1021/ef201578a.
- [29] S. Choi, M.L. Gray, C.W. Jones, Amine-tethered solid adsorbents coupling high adsorption capacity and regenerability for CO₂ capture from ambient air., *ChemSusChem.* 4 (2011) 628–35. doi:10.1002/cssc.201000355.

- [30] S. Hammache, J.S. Hoffman, M.L. Gray, D.J. Fauth, B.H. Howard, H.W. Pennline, Comprehensive Study of the Impact of Steam on Polyethyleneimine on Silica for CO₂ Capture, *Energy & Fuels*. 27 (2013) 6899–6905. doi:10.1021/ef401562w.
- [31] J.R. Tucker, L.J. Shadle, S. Benyahia, J. Mei, C. Guenther, M.E. Koepke, Improvement in Precision, Accuracy, and Efficiency in Standardizing the Characterization of Granular Materials, in: Vol. 7A Fluids Eng. Syst. Technol., ASME, 2013: p. V07AT08A064. doi:10.1115/IMECE2013-65027.
- [32] T.J. Benyahia, S., Syamlal, M., O'Brien, Summary of MFIx Equations, 2012.
- [33] O.T.J. Syamlal, M., Rogers, W., MFIx documentation: Theory guide, Morgantown, 1993.
- [34] C.K.K. Lun, S.B. Savage, D.J. Jeffrey, N. Chepurniy, Kinetic theories for granular flow: inelastic particles in Couette flow and slightly inelastic particles in a general flowfield, *J. Fluid Mech.* 140 (1984) 223. doi:10.1017/S0022112084000586.
- [35] D.G. Schaeffer, Instability in the evolution equations describing incompressible granular flow, *J. Differ. Equ.* 66 (1987) 19–50. doi:10.1016/0022-0396(87)90038-6.
- [36] J.-F. Dietiker, T. Li, R. Garg, M. Shahn timer, Cartesian grid simulations of gas–solids flow systems with complex geometry, *Powder Technol.* 235 (2013) 696–705. doi:10.1016/j.powtec.2012.11.028.
- [37] C. Lai, Z. Xu, W. Pan, X. Sun, C. Storlie, P. Marcy, et al., Hierarchical calibration and validation of computational fluid dynamics models for solid sorbent-based carbon capture, *Powder Technol.* 288 (2016) 388–406. doi:10.1016/j.powtec.2015.11.021.
- [38] S.B.R. Karri, T.M. Knowlton, Gas Distributor and Plenum Design in Fluidized Beds, in: W.-C. Yang (Ed.), *Handb. Fluid. Fluid-Particle Syst.*, CRC Press, 2003: pp. 209–235. <http://www.crcnetbase.com/doi/abs/10.1201/9780203912744.ch6> (accessed June 22, 2015).
- [39] K. Agrawal, P.N. Loezos, M. Syamlal, S. Sundaresan, The role of meso-scale structures in rapid gas–solid flows, *J. Fluid Mech.* 445 (2001) 151–185. doi:10.1017/S0022112001005663.
- [40] T. Li, A. Gel, S. Pannala, M. Shahn timer, M. Syamlal, CFD simulations of circulating fluidized bed risers, part I: Grid study, *Powder Technol.* 254 (2014) 170–180.

doi:10.1016/j.powtec.2014.01.021.

- [41] T. Li, S. Benyahia, J.-F. Dietiker, J. Musser, X. Sun, A 2.5D computational method to simulate cylindrical fluidized beds, *Chem. Eng. Sci.* 123 (2015) 236–246. doi:10.1016/j.ces.2014.11.022.
- [42] S. Cloete, S.T. Johansen, S. Amini, Grid independence behaviour of fluidized bed reactor simulations using the Two Fluid Model: Effect of particle size, *Powder Technol.* 269 (2015) 153–165. doi:10.1016/j.powtec.2014.08.055.
- [43] K. Kuwagi, A. Kogane, H. Hirano, A. Bin Alias, T. Takami, Non-dimensionalization and three-dimensional flow regime map for fluidization analyses, *Chem. Eng. Sci.* 119 (2014) 199–211. doi:10.1016/j.ces.2014.08.006.
- [44] T. Li, S. Pannala, M. Shahnam, CFD simulations of circulating fluidized bed risers, part II, evaluation of differences between 2D and 3D simulations, *Powder Technol.* 254 (2014) 115–124. doi:10.1016/j.powtec.2014.01.022.
- [45] N. Xie, F. Battaglia, S. Pannala, Effects of using two- versus three-dimensional computational modeling of fluidized beds, *Powder Technol.* 182 (2008) 1–13. doi:10.1016/j.powtec.2007.07.005.
- [46] L. Cammarata, P. Lettieri, G.D.M. Micale, D. Colman, 2D and 3D CFD Simulations of Bubbling Fluidized Beds Using Eulerian-Eulerian Models, *Int. J. Chem. React. Eng.* 1 (2003). doi:10.2202/1542-6580.1083.
- [47] E. Peirano, V. Delloume, B. Leckner, Two- or three-dimensional simulations of turbulent gas – solid flows applied to fluidization, *56* (2001) 4787–4799.
- [48] N. Reuge, L. Cadoret, C. Coufort-Saudejaud, S. Pannala, M. Syamlal, B. Caussat, Multifluid Eulerian modeling of dense gas–solids fluidized bed hydrodynamics: Influence of the dissipation parameters, *Chem. Eng. Sci.* 63 (2008) 5540–5551. doi:10.1016/j.ces.2008.07.028.
- [49] T. Li, J. Grace, X. Bi, Study of wall boundary condition in numerical simulations of bubbling fluidized beds, *Powder Technol.* 203 (2010) 447–457. doi:10.1016/j.powtec.2010.06.005.
- [50] R.C. Darton, R.D. LaNauze, J.F. Davidson, D. Harrison, Bubble growth due to

- coalescence in fluidised beds, *Trans Inst Chem Eng.* 55 (1977) 274–280.
- [51] A. Passalacqua, L. Marmo, A critical comparison of frictional stress models applied to the simulation of bubbling fluidized beds, *Chem. Eng. Sci.* 64 (2009) 2795–2806. doi:10.1016/j.ces.2009.03.005.
- [52] M. Farzaneh, A.-E. Almstedt, F. Johnsson, D. Pallarès, S. Sasic, The crucial role of frictional stress models for simulation of bubbling fluidized beds, *Powder Technol.* 270 (2015) 68–82. doi:10.1016/j.powtec.2014.09.050.
- [53] A. Srivastava, S. Sundaresan, Analysis of a frictional–kinetic model for gas–particle flow, *Powder Technol.* 129 (2003) 72–85. doi:10.1016/S0032-5910(02)00132-8.
- [54] C. Loha, H. Chattopadhyay, P.K. Chatterjee, Assessment of drag models in simulating bubbling fluidized bed hydrodynamics, *Chem. Eng. Sci.* 75 (2012) 400–407. doi:10.1016/j.ces.2012.03.044.
- [55] A. Bakshi, C. Altantzis, R.B. Bates, A.F. Ghoniem, Eulerian–Eulerian simulation of dense solid–gas cylindrical fluidized beds: Impact of wall boundary condition and drag model on fluidization, *Powder Technol.* 277 (2015) 47–62. doi:10.1016/j.powtec.2015.02.056.
- [56] M. Syamlal, T.J. O’Brien, Fluid dynamic simulation of O₃ decomposition in a bubbling fluidized bed, *AIChE J.* 49 (2003) 2793–2801. doi:10.1002/aic.690491112.
- [57] N. Yang, W. Wang, W. Ge, J. Li, CFD simulation of concurrent-up gas–solid flow in circulating fluidized beds with structure-dependent drag coefficient, *Chem. Eng. J.* 96 (2003) 71–80. doi:10.1016/j.cej.2003.08.006.
- [58] L.S. Ferreira, J.O. Trierweiler, Modeling and simulation of the polymeric nanocapsule formation process, in: *IFAC Proc. Vol.*, 2009: pp. 405–410. doi:10.1002/aic.
- [59] Y. Igci, S. Sundaresan, Constitutive Models for Filtered Two-Fluid Models of Fluidized Gas–Particle Flows, *Ind. Eng. Chem. Res.* 50 (2011) 13190–13201. doi:10.1021/ie200190q.
- [60] C.C. Milioli, F.E. Milioli, W. Holloway, K. Agrawal, S. Sundaresan, Filtered two-fluid models of fluidized gas-particle flows: New constitutive relations, *AIChE J.* 59 (2013) 3265–3275. doi:10.1002/aic.14130.

- [61] Y. Igci, S. Sundaresan, Verification of filtered two-fluid models for gas-particle flows in risers, *AIChE J.* 57 (2011) 2691–2707. doi:10.1002/aic.12486.
- [62] Y. Igci, S. Pannala, S. Benyahia, S. Sundaresan, Validation Studies on Filtered Model Equations for Gas-Particle Flows in Risers, *Ind. Eng. Chem. Res.* 51 (2012) 2094–2103. doi:10.1021/ie2007278.

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(a)

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Figure 12. Snapshot of bubbles predicted by different drag models (a) Gidaspow drag model; (b) Igci et al. filtered drag model; (c) Sarkar et al. filtered drag model.

Figure 13. Snapshot of bubbles predicted by Sarkar et al. filtered drag model at different grid resolutions (a) 4 mm grid; (b) 2 mm grid; (c) 1.25 mm grid.

Figure 14. Mean bubble size along the height predicted by different grid resolutions extracted from 2D slices.

Figure 15. Comparison of mean bubble size along the height predicted by different grid resolutions with the empirical correlation.

(a)

(b)

Figure 16. Mean bubble sizes predicted for different operating conditions (a) 15 cm static bed height, (b) 20 cm static bed height.

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Table 1. Physical properties of 32D sorbents.

Sauter mean particle diameter (μm)	100
Sphericity (-)	0.89
Particle density (kg/m^3)	480
Particle Skeletal Density (kg/m^3)	1690
Bulk density (kg/m^3)	300
Particle porosity (-)	0.71
Void fraction (-)	0.45
Minimum fluidization velocity (m/s)	0.0023

Table 2. Summary of MFIX equations solved for the current simulations.

A. Governing equations

(a) Continuity equations

$$\frac{\partial}{\partial t}(\varepsilon_g \rho_g) + \nabla \cdot (\varepsilon_g \rho_g \vec{V}_g) = 0 \quad (1)$$

$$\frac{\partial}{\partial t}(\varepsilon_p \rho_p) + \nabla \cdot (\varepsilon_p \rho_p \vec{V}_p) = 0 \quad (2)$$

(b) Momentum equations

$$\frac{\partial}{\partial t}(\varepsilon_g \rho_g \vec{V}_g) + \nabla \cdot (\varepsilon_g \rho_g \vec{V}_g \vec{V}_g) = \nabla \cdot \bar{\bar{\tau}}_g - \varepsilon_g \nabla P + \varepsilon_g \rho_g g - I_{gp} \quad (3)$$

$$\frac{\partial}{\partial t}(\varepsilon_p \rho_p \vec{V}_p) + \nabla \cdot (\varepsilon_p \rho_p \vec{V}_p \vec{V}_p) = \nabla \cdot \bar{\bar{\tau}}_p - \varepsilon_p \nabla P + \varepsilon_p \rho_p g + I_{gp} \quad (4)$$

B. Constitutive equations

(a) Gas stress tensor

$$\bar{\bar{\tau}}_g = 2\mu_{ge} \bar{\bar{S}}_g \quad (5)$$

$$\bar{\bar{S}}_g = \frac{1}{2}(\nabla \vec{V}_g + (\nabla \vec{V}_g)^T) - \frac{1}{3} \nabla \cdot \vec{V}_g \bar{\bar{I}} \quad (6)$$

$$\bar{\bar{D}}_g = \frac{1}{2}(\nabla \vec{V}_g + (\nabla \vec{V}_g)^T) \quad (7)$$

(b) Solid stress tensor

$$\bar{\bar{\tau}}_p = (-P_s + \eta\mu_b \nabla \cdot \vec{V}_p) \bar{\bar{I}} + 2\mu_p \bar{\bar{S}}_p \quad (8)$$

$$\bar{\bar{S}}_p = \frac{1}{2}(\nabla \vec{V}_p + (\nabla \vec{V}_p)^T) - \frac{1}{3} \nabla \cdot \vec{V}_p \bar{\bar{I}} \quad (9)$$

$$P_s = \varepsilon_p \rho_p \Theta_p [1 + 4g_0 \varepsilon_p \eta] \quad (10)$$

$$\mu_p = \left(\frac{2 + \alpha}{3} \right) \left[\frac{\mu_p^*}{g_0 \eta (2 - \eta)} \left(1 + \frac{8}{5} \eta g_0 \varepsilon_p \right) \left(1 + \frac{8}{5} \eta (3\eta - 2) g_0 \varepsilon_p \right) + \frac{3}{5} \eta \mu_b \right] \quad (11)$$

$$\mu_p^* = \frac{\varepsilon_p \rho_p \Theta_p g_0 \mu}{\varepsilon_p \rho_p \Theta_p g_0 + \frac{2\beta\mu}{\varepsilon_p \rho_p}} \quad (12)$$

$$\mu = \frac{5}{96} \rho_p d_p \sqrt{\pi \Theta_p} \quad (13)$$

$$\mu_b = \frac{256}{5\pi} \mu \varepsilon_p^2 g_0 \quad (14)$$

$$\eta = \frac{1 + e}{2} \quad (15)$$

$$\bar{\bar{\sigma}}_{p,fric} = -P_{s,fric} \left(\bar{\bar{I}} - \frac{\sin \phi}{\sqrt{I_{2D}}} \bar{\bar{D}}_p \right) \quad (16)$$

$$\mu_{fric} = \frac{P_{s,fric} \sin \phi}{2\sqrt{I_{2D}}} \quad (17)$$

$$P_{s,fric} = \begin{cases} 10^{24} (\varepsilon_p - \varepsilon_p^*)^2 & \varepsilon_p > \varepsilon_p^* \\ 0 & \varepsilon_p \leq \varepsilon_p^* \end{cases} \quad (18)$$

(c) Granular temperature

$$\Theta_p = \left[\frac{-(K_1 \varepsilon_p + \rho_p) Tr(\bar{\bar{D}}_p)}{2K_4 \varepsilon_p} + \frac{\sqrt{(K_1 \varepsilon_p)^2 Tr^2(\bar{\bar{D}}_p) + 4K_4 \varepsilon_p [2K_3 Tr(\bar{\bar{D}}_p^2) + K_2 Tr^2(\bar{\bar{D}}_p)]}}{2K_4 \varepsilon_p} \right]^2 \quad (19)$$

$$K_1 = 2(1 - e) \rho_p g_0 \quad (20)$$

$$K_2 = \frac{4}{3\sqrt{\pi}} d_p \rho_p (1 + e) g_0 \varepsilon_p - \frac{2}{3} K_3 \quad (21)$$

$$K_3 = \frac{d_p \rho_p}{2} \left\{ \frac{\sqrt{\pi}}{3(3 - e)} \left[\frac{(3e + 1)}{2} + \frac{2}{5} (1 + e)(3e - 1) g_0 \varepsilon_p \right] + \frac{8\varepsilon_p}{5\sqrt{\pi}} g_0 (1 + e) \right\} \quad (22)$$

$$K_4 = \frac{12(1 - e^2) \rho_p g_0}{d_p \sqrt{\pi}} \quad (23)$$

$$g_0 = \frac{1 - 0.5\varepsilon_p}{(1 - \varepsilon_p)^3} \quad (24)$$

(d) Inter-phase momentum exchange

$$I_{gp} = \beta (\vec{V}_g - \vec{V}_p) \quad (25)$$

$$\beta = \begin{cases} 150 \frac{\varepsilon_p^2 \mu_g}{\varepsilon_g d_p^2} + 1.75 \frac{\varepsilon_p \rho_g |\vec{V}_p - \vec{V}_g|}{d_p} & \text{if } \varepsilon_p > 0.2 \\ \frac{3}{4} C_d \varepsilon_g^{-2.65} \frac{\varepsilon_p \varepsilon_g \rho_g |\vec{V}_p - \vec{V}_g|}{d_p} & \text{if } \varepsilon_p \leq 0.2 \end{cases} \quad (26)$$

$$C_d = \begin{cases} \frac{24}{\text{Re} \cdot \varepsilon_g} (1 + 0.15 (\text{Re} \cdot \varepsilon_g)^{0.687}) & \text{if } \text{Re} \cdot \varepsilon_g < 1000 \\ 0.44 & \text{if } \text{Re} \cdot \varepsilon_g \geq 1000 \end{cases} \quad (27)$$

$$\text{Re} = \frac{\rho_g |\vec{V}_p - \vec{V}_g| d_p}{\mu_g} \quad (28)$$

Table 3. Numerical parameters used in the simulations.

Parameter	Value	Parameter	Value
Diameter (cm)	10	Height (cm)	100
Temperature (K)	297	Pressure (atm)	1
Gas viscosity (Pa.s)	1.8e-5	Gas molecular weight (kg/kmol)	28.8
Particle diameter (μm)	100	Particle density (kg/m^3)	480
Estimated inter-particle restitution coefficient (-)	0.9	Estimated particle-wall restitution coefficient	0.8
Estimated angle of inter-particle friction (deg)	30	Estimated particle-wall frictional angle (deg)	30

Table 4. Pressure drops measured by PDT 812 between 4.46 and 10.16 cm for different static bed heights and superficial gas velocities (unit: Pa).

Hs (cm)\Ug	$1U_{mf}$	$3U_{mf}$	$5U_{mf}$	$7U_{mf}$
10	118.56 $\pm$ 1.6	136.36 $\pm$ 2.3	133.39 $\pm$ 2.1	134.56 $\pm$ 2.3
15	130.61 $\pm$ 3.7	139.31 $\pm$ 2.5	137.41 $\pm$ 4.1	138.49 $\pm$ 2.6
20	133.14 $\pm$ 4.6	140.89 $\pm$ 3.2	141.33 $\pm$ 2.4	140.52 $\pm$ 1.9

Table 5. Pressure drops measured by PDT850 between 10.16 and 20.26 cm for different static bed heights and superficial gas velocities.

Hs (cm)\Ug	$1U_{mf}$	$3U_{mf}$	$5U_{mf}$	$7U_{mf}$
10	0	25.46 $\pm$ 1.2	35.44 $\pm$ 0.6	34.39 $\pm$ 0.6
15	99.46 $\pm$ 2.4	159.05 $\pm$ 2.3	168.55 $\pm$ 0.6	167.76 $\pm$ 0.6
20	216.82 $\pm$ 5.5	249.41 $\pm$ 2.1	246.49 $\pm$ 1.6	249.23 $\pm$ 1.6

Table 6. Bed heights and bubble counts predicted by simulations of different model settings.

Tests	ASGS	PPGS	PPSS	PPGM
Granular temperature	ALG	PDE	PDE	PDE
Friction model	Shaefer	Princeton	Princeton	Princeton
Drag model	Gidaspow	Gidaspow	Syam_O'Brien	Gidaspow
Discretization	Superbee	Superbee	Superbee	Minmod
Bed height (cm)	19.8	20	18.5	22.45
Bubble count	139,014	98,133	56,301	41,059

Table 7. Filtered sub-grid models used in the current study.

Sarkar et al. [26]

$$H_d = \min \left[\left(1 + \frac{b}{|v_{slip}^*|} \right) \bar{\phi}_p^{\left(c + \frac{d}{|v_{slip}^*|} \right)}, 0.97 \right] \quad (31)$$

$$\Delta_{filter}^* = \text{from } 1.3495 \text{ to } 6.7476$$

$$a = 0.9506$$

$$b = 0.1708$$

$$c = 0.049 \left(\frac{1}{\Delta_{filter}^*} - 1 \right)$$

$$d = 0.3358$$

$$v_{slip}^* = \frac{|\mathbf{u}_s - \mathbf{u}_g|}{v_t}$$

$$\Delta_{filter}^* = \frac{g \Delta_{filter}}{v_t^2}$$

Igci et al. 2013 [59]

$$H_d = f(\Delta_{filter}^*) H_{3D}(\varepsilon_p) \quad (32)$$

$$f(\Delta_{filter}^*) = \frac{(\Delta_{filter}^*)^{1.6}}{(\Delta_{filter}^*)^{1.6} + 0.4}$$

$$H_{3D}(\varepsilon_p) = H_{2D}(\varepsilon_p) g g(\varepsilon_p)$$

$$H_{2D}(\varepsilon_p) = \begin{cases} 2.7 \varepsilon_p^{0.234} & \varepsilon_p < 0.012 \\ -0.019 \varepsilon_p^{-0.455} + 0.963 & 0.012 \leq \varepsilon_p < 0.014 \\ 0.868 \exp(-0.38 \varepsilon_p) - 0.176 \exp(-119.2 \varepsilon_p) & 0.014 \leq \varepsilon_p < 0.25 \\ -4.59 \times 10^{-5} \exp(19.75 \varepsilon_p) + 0.852 \exp(-0.268 \varepsilon_p) & 0.25 \leq \varepsilon_p < 0.455 \\ (\varepsilon_p - 0.59)(-1501 \varepsilon_p^3 + 2203 \varepsilon_p^2 - 1054 \varepsilon_p + 162) & 0.455 \leq \varepsilon_p \leq 0.59 \\ 0 & 0.59 < \varepsilon_p \leq 0.65 \end{cases}$$

$$g g(\varepsilon_p) = \begin{cases} \varepsilon_p^{0.24} (1.48 + \exp(-18 \varepsilon_p)) & \varepsilon_p < 0.18 \\ 1 & \varepsilon_p \geq 0.18 \end{cases}$$

$$\Delta_{filter}^* = \frac{g \Delta_{filter}}{v_t^2}$$

Table 8. Bed heights predicted by different drag models and experimental measurement.

Tests	Gidaspow	lgci et al.	Sarkar et al.	Exp.
Bed height (cm)	25.8	21.3	17.1	18.7

Table 9. Numerical predictions and experimental measurement of pressure drop for different tests.

Condition\result	Exp. (PDT812)	Num.	Exp.(PDT850)	Num.
15cm, 3U _{mf}	139.3 ± 2.5	142	159.1 ± 2.3	180
15cm, 5U _{mf}	137.4 ± 4.1	146	168.5 ± 0.5	175
15cm, 7U _{mf}	138.5 ± 2.6	145	167.8 ± 0.5	175
20cm, 3U _{mf}	140.9 ± 3.2	145	249.4 ± 2.1	259
20cm, 5U _{mf}	141.3 ± 2.4	145	246.5 ± 1.6	268
20cm, 7U _{mf}	140.5 ± 1.9	146	249.2 ± 1.6	264