VERIFICATION OF THE IMPORTANCE OF INTRODUCING PARTICLE SIZE DISTRIBUTIONS TO BUBBLING FLUIDIZED BED SIMULATIONS

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Abstract

Fluidized beds are widely used in industrial operations due to their ability to give good mixing and a high contact area between the phases. The excellent controlling ability of temperature allows good operating conditions for solid catalyzed gas phase reactions and also the ease of the design.

Powders used in industrial fluidized beds have a particle size distribution, and the particle size distribution influence significantly on the flow behaviour. In modelling of fluidized beds a mean particle diameter is often used, and important information about flow behaviour can therefore be lost. The objective of this work is to study the influence of including particle size distribution in the simulation of a 2-D bubbling fluidized bed. Related to this work a series of simulations are performed using the commercial CFD software FLUENT 6.3.

The simulations are run with one, two, three and four particle phases. The particle size distribution is accounted for by including multiple particle phases. The computational results are compared to results from experiments performed by Mr. W.J. Wu at Telemark University College, Norway.

The computational results are compared with each other with respect to the bubble appearance, bubble distribution, bubble velocity, bed expansion and particle segregation. The comparison shows that the results vary significantly depending on the number of particle phases used.

Computational results of bubble velocity, bubble distribution, bed expansion and particle segregation are compared to the experimental data. The results from the simulations with three and four particle phases agree well with the experimental results according to bed expansion and bubble behaviour.

The simulations show the importance of accounting for the particle size distribution in the computational model. By using one particle size, important information of the flow behaviour is lost, and the results deviate significantly from the experiments.

1 Introduction

Fluidization is a well known mechanism in industry for the purpose of mixing the particles. This operation makes the solid to achieve a fluid like behavior while suspending it in a gas or a liquid. The fluid like behavior of solids gives a rapid and easy transportation ability with intimate gas contacting, which is the most important factor that makes the fluidization an important unit operation used in industry.

Fluidized beds are used in industry as heat exchangers due to there unique ability to rapidly transport heat and maintain a uniform temperature, to make granules through solidifying a melt, for the purpose of coating metal objects with plastics and other objects like tablets of drugs, sweets and etc and for growing of particles like table salt. Drying of solids is an other application of fluidization. The fluidized bed dryer is used extensively in a wide variety of industries because of its large capacity, low construction cost, easy operability, and high thermal efficiency [1]. In addition to those, fluidized beds are used in industry in order to carry out synthesis reactions, for cracking the hydrocarbons and for the combustion of low grade coal and oil shale fines, fuels that cannot be burned efficiently in conventional boiler furnaces and for the incineration of solid waste. Carbonization and gasification is also an area with importance of fluidization. The latest area of application of fluidized beds is in operation of bio-fluidization, in other words the cultivation of microorganisms.

There are mainly two types of fluidization systems, the solid-liquid systems and solid-gas systems. Unlike the liquid-solid fluidized beds, the gas-solid fluidized beds have some unusual and useful properties compared to other contacting and mixing methods. The gas-solid fluidized beds looks very much like a boiling liquid and in many ways exhibits liquid-like behavior [2]. Only the gas-solid systems will be considered here.

Previous paragraphs show that the fluidized beds are widely used in a vast range of industrial applications. In addition it is emphasized that the bubbling fluidization (bubbling fluidized beds) is in a competitive position with the circulation fluidization and only the bubbling fluidization will be studied here.

It is important to study about the dynamics and other properties of the bubbling fluidized beds. The efficiency of the bubbling fluidized beds are dependant on the bubble size, bubble frequency, bubble distribution and bubble velocity in the bed. The bubble characteristics are very important in the design of fluidized beds because they govern hydrodynamics and efficiency of the operation for which the bed is used [3]. Particles used in industrial operations are usually consist of a wide range of particle sizes (distribution of particle sizes). It is with a great importance to study how those things dependant on the particle size distribution.

Simulations with satisfactory results are the prime requirement for this type of studies. Common practice is to use the mean diameter of the particles to represent powders in simulations. As the fluid dynamics should be dependant on the particle size distribution, above mentioned practice can lead to loss of valuable information. It is important to check the influence from introducing particle size distributions in simulations.

Researchers have used both the Euler-Euler approach and the Euler-Lagrange approach for fluidized bed simulations depending on the requirements. Halvorsen, B. [4] has used the Euler-Euler approach with MFIX software programme in her simulations of bubbling fluidized beds. Patil et al [5] and [6] have used Euler-Euler approach with two different closure models. Those are the constant viscosity model and a model based on the kinetic theory of granular flow. They have compared the simulated results of the two models with each other and also with the experimental results. Enwald et at [7] have presented a model using Euler-Euler approach as well as the application of the model in the simulations of bubbling and circulating fluidized beds.

Huilin et al [8] has used both approaches separately showing the results as a comparison with the experiments. Details of particle collision information are obtained through tracing particle motions based on Euler-Lagrange approach coupled with the discreet hard sphere model. A CFD model based on kinetic theory of granular flow and Euler-Euler approach is used to simulate flows in bubbling gas-solid fluidized beds.

Boemer et al [9] have developed a computer code to simulate the fluid dynamics of fluidized beds using Eularian approach. Arastoopour, H. [10] has used Eularian approach for the simulations he used to compare the predicted flow parameters with large scale experimental data of fluidized beds.

Huilin et al [11] has used a multi fluid Eulerian CFD model with closure relationships according to the kinetic theory of granular flow to study the motion of particles in a gas bubbling fluidized bed with the binary mixtures. They have concluded that in order to obtain realistic bed dynamics from fundamental hydrodynamic models, it is important to correctly take the effect of particle size distribution and energy dissipation due to non-ideal particle-particle interactions into account.

Different solid phases can be used to represent different particle sizes of a distribution in a simulation. As found from the literature survey, most of the simulations of bubbling fluidized beds have used only one or two solid phases and it is interesting to use more than two particle phases in simulations and check the influence.

A computational study of the influence of particle size distribution on bubbling fluidized beds is performed. The commercial software FLUENT 6.3 version is used to perform the simulations. The results of the simulations are compared with a reference experiment. The simulations used the same dimensions for the particle bed as in the reference experiment.

2 Mathematical Model Used in the Simulations

A combination of models with Eulerian multiphase approach that is finalized by Ariyarathna D.G.A.S.U. (2008) [12] is used to simulate the 2-D fluidized bed with uniform distribution of air in order to check the influence of particle size distribution on simulations. The "Syamlal O'Brien Symmetric" drag model is used to introduce the solid-solid drag forces and the "Syamlal O'Brien" drag model to introduce the solid-fluid drag forces.

The recommended combination of models is presented in Tab. 1.

Some of the important equations are taken from the Fluent User Guide [13] and presented here in this publication. The momentum exchange coeffi-

Table 1: Recommended combination of models.

Drag Model	Syamlal O'Brien
Granular Viscosity	Syamlal O'Brien
Granular Bulk Viscosity	Constant
Frictional Viscosity	Schaeffer
Frictional Pressure	Based-ktgf
Solid Pressure	Ma-ahmadi
Radial Distribution Function	Ma-ahmadi
Rest of the models	use default settings

cient between the fluid and solid phases, $K_{\ell\mathrm{s}}$ and the drag coefficient, C_D are given by the Eq. 1 and Eq. 2.

$$K_{\ell\mathrm{s}} = \frac{\alpha_\mathrm{s}\rho_\mathrm{s}\left(\frac{C_\mathrm{D}\,\mathrm{Re_s}\,\alpha_\ell}{24 v_\mathrm{r,s}^2}\right)}{\tau_\mathrm{s}}, \tag{1}$$

$$C_\mathrm{D} = \left(0.63 + \frac{4.8}{\sqrt{\mathrm{Re_s}/v_\mathrm{r,s}}}\right)^2 \tag{2}$$

Here $\mathrm{Re_s}, \alpha_\mathrm{s}$ and ρ_s are the relative Reynolds number, the phasic volume fraction and the physical density of the solid phase. α_ℓ, τ_s and $v_\mathrm{r,s}$ are the phasic volume fraction of the liquid, the solid phase stress-strain tensor and the terminal velocity for the solid phase respectively.

Granular viscosity, μ_g is presented in the Eq. 3. It is consists with the kinetic and collisional viscosity terms, $\mu_\mathrm{s,kin}$ and $\mu_\mathrm{s,\,col}$ and those are presented in the Eq. 4 and Eq. 5.

$$\mu_\mathrm{g} = \mu_\mathrm{s,kin} + \mu_\mathrm{s,\,col} \tag{3}$$

$$\mu_\mathrm{s,kin} = \frac{\alpha_\mathrm{s} d_\mathrm{s}\rho_\mathrm{s}\sqrt{\Theta_\mathrm{s}\pi}}{6\,(3-e_\mathrm{ss})}\left[1+\frac{2}{5}\,(1+e_\mathrm{ss})\,(3e_\mathrm{ss}-1)\,\alpha_\mathrm{s} g_\mathrm{0,ss}\right] \tag{4}$$

$$\mu_\mathrm{s,\,col} = \frac{4}{5}\alpha_\mathrm{s} d_\mathrm{s}\rho_\mathrm{s} g_\mathrm{0,ss}\,(1+e_\mathrm{ss})\left(\frac{\Theta_\mathrm{s}}{\pi}\right)^{1/2} \tag{5}$$

Here d_s is the diameter of the $\mathrm{s^{th}}$ solid phase particles.

Frictional pressure, $p_\mathrm{,fr}$ is given by the Eq. 6, where ϕ, $\mu_\mathrm{,fr}$ and I_2D are the angle of internal friction, frictional viscosity and the second invariant of the deviatoric stress tensor.

$$p_\mathrm{,fr} = \frac{\mu_\mathrm{,fr} * \sqrt[2]{I_\mathrm{2D}}}{\sin\phi} \tag{6}$$

Solids pressure, p_s is presented by the Eq. 7 and the radial distribution function, $g_\mathrm{0,ss}$ is given by the Eq. 8. Where Θ_s and e_ss is the granular temperature and the coefficient of restitution for particle collisions respectively. d_ℓ is the diameter of the ℓ^th solid phase particles and $\alpha_\mathrm{s} = \sum_\mathrm{k=1}^\mathrm{n}\alpha_\mathrm{k}$. α_k and ρ_k are the phasic volume fraction and physical density of each phase if more than one solid phase exists.

$$p_\mathrm{s} = \alpha_\mathrm{s}\rho_\mathrm{s}\Theta_\mathrm{s}\left[\begin{array}{c}(1+4\alpha_\mathrm{s} g_\mathrm{0,ss})+\\ \frac{1}{2}\left[(1+e_\mathrm{ss})\left(1-e_\mathrm{ss}+2\mu_\mathrm{,fr}\right)\right]\end{array}\right] \tag{7}$$

$$g_\mathrm{0,ss} = \frac{1+2.5\alpha_\mathrm{s}+4.59\alpha_\mathrm{s}^2+4.52\alpha_\mathrm{s}^3}{\left(1-\left(\frac{\alpha_\mathrm{s}}{\alpha_\mathrm{s,max}}\right)^3\right)^{0.678}} + \frac{1}{2} d_\ell \sum_\mathrm{k=1}^\mathrm{N}\frac{\alpha_\mathrm{k}}{\rho_\mathrm{k}} \tag{8}$$

3 Computational and Experimental Setup

Four simulations, P1,P2,P3 and P4 are performed with increasing number of particle phases in the bed, such as, the simulation P1 with one particle phase, the simulation P2 with two particle phases and the simulations P3 and P4 with three and four particles phases in each.

Representation of the particle size distribution in the simulations is arranged according to the particle distributions of the particle mixture used in the reference experiment. Each particle phase is represented by the corresponding mean particle diameter. The same total mean particle diameter persists in all four simulations.

All four simulations are performed using the same conditions except the number of particle phases. A wire frame mesh with 0.2 m and 1.5 m as the column width and the height, is used. When the number of solid phases are increasing the compositions of the particle phases are computed as the same mean particle diameter is provided in every mixture. The mean diameter of each particle phase is selected using the particle size distribution of the powders used in the reference experiment.

The reference experiment is performed using a mixture of three type of powders. Each of those powders have their own particle size distributions. Mean particle diameters and the compositions of the particle phases used in the simulations are presented in Tab. 2.

The superficial gas velocity of 0.134 $\mathrm{ms^{-1}}$ is used both in the simulations and the reference experiment. Each simulation represents 30 seconds from the flow time.

4 Results

The simulation P1 is a special case as it didn't give any changes in the VOF and also there were no bubbles in the particle bed. The theoretical $\mathrm{U_{mf}}$ for the corresponding particle size is 0.19535 $\mathrm{ms^{-1}}$ and it is higher than the superficial gas velocity used in

Table 2: Mean particle size and compositions of the particle phases.

Sim: No:	Phase1	Phase2	Phase3	Phase4
Mean diameter (μm)				
P1	487.97			
P2	153	624.79		
P3	153	487.97	960	
P4	153	424.6	577.78	960
Composition (%)				
P1	100			
P2	29	71		
P3	29	50	21	
P4	29	30.5	19.5	21

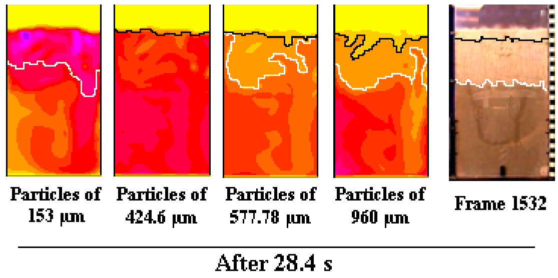

Figure 1: Comparison of particle segregation in the simulation P4 and the reference experiment with respective to all four particle phases in the simulation.

the simulations. That is the reason for not having any bubbles and no variations in the particle VOF in the particle bed.

4.1 Particle Segregation

Particle segregation was clearly noticeable in the reference experiment as well as in all the simulations except simulation P1. Contours from the simulations representing the volume fraction (VOF) distributions of each particle phase were analyzed to check the prediction of the particle size distribution in those simulations. It is easily noticed that the higher the number of particle phases used and better the representation of the particle size distribution the better the prediction of the particle segregation.

Fig. 1 shows the VOF of all the particle phases used in the simulation P4 after 28.4 seconds from the beginning of the simulation. It provided clear evidence about the settling down of the large particles close to the bottom and the small particles close to the top of the particle bed. Eventhough each phase alone doesn't give much meaning, when the prediction in all four phases are considered together, it shows closer prediction to the results of the reference experiment.

In the experiment it is possible to identify two boundaries with regard to particle segregation. The simulation P4 has predicted the second margin also while the simulation P3 doesn't manage to predict the second margin that clearly. The first margin is marked in white and the second margin which is closer to the bed surface is marked with a black line in the simulation frames. In the frame from the experiment the first margin is marked with black and the second margin is marked with a lighter color.

The progress of particle segregation is also studied. Contours of VOF of small particles from the simulation P4 with time is given in the Fig. 2 as a comparison of the simulation P4 with the reference experiment. The reason for selecting only the small particle phase is that the small particles were representing the segregation boundaries as well as the top surface of the particle bed clearly. The comparison proves that the prediction of particle segregation agrees well with the reference experiment.

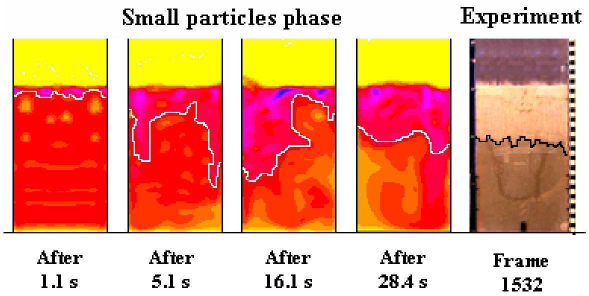

Figure 2: Comparison of particle segregation in the simulation P4 and the reference experiment with respective the small particle's phase in the simulation.

In addition to the contour analysis, plots of VOF values along the height of the bed are also made using the readings from the monitors, which gives a better picture of segregation. VOF are averaged for the last 25 s of the simulation time. Two radial positions are selected and the averaged VOF data are plotted along the height of the particle bed at those positions. One position is close to a wall (0.05 away from a wall) and the other is in the center of the column. It is assumed that the analysis of only one side of the bed cross section is enough even though the behavior of both sides are not exactly the same all the time.

Fig. 3 and Fig. 4 present the plots from the simulation P4. These plots provide evidence about the contribution of different particle types for particle segregation. Gradient of each plot shows how strong the separation at each particle phase. Fig. 3 shows that the VOF of the small particles achieve values that are even smaller than the values at the bottom of the bed. That can be due to bubble formation

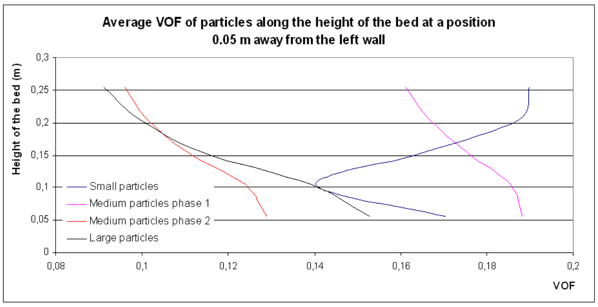

Figure 3: VOF of the particles phases along the height of the bed at a position 0.05 m away from the wall predicted by the simulation P4.

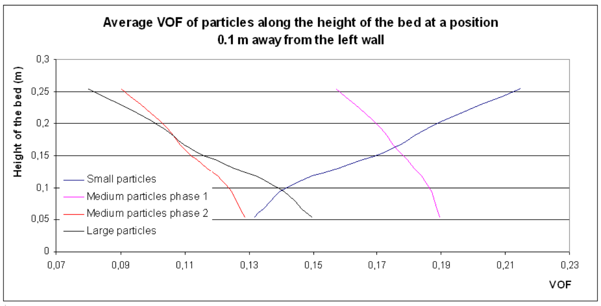

Figure 4: VOF of the particles phases along the height of the bed at a position 0.1 m away from the wall predicted by the simulation P4.

at that area of the bed.

4.2 Bubble Behavior

Bubble distribution in the particle bed of each simulation is compared with the reference experiment using the contours of the VOF of the gas phase. The contours selected for the analysis are well distributed in the whole time domain of the simulations.

Some frames from the movie of the reference experiment are used to present the bubble appearance in the experiment and those are presented in Fig. 5. It provides that the lowest position of bubble appearance in the experiment is 23.2 cm approximately.

Fig. 6 shows the bubble distribution in the particle bed at different time instances of the simulation P2. Eventhough most of the bubbles are appeared close to the walls there are some bubbles appeared in the middle area of the bed also, when the radial positions are considered. In addition, the simulation P2 has predicted bubbles even in lower positions about 7.4 cm in the bed. Bubble distribution predicted by this simulation with two particle phases shows that there is a major effect by introducing particle size distribution to CFD simu-

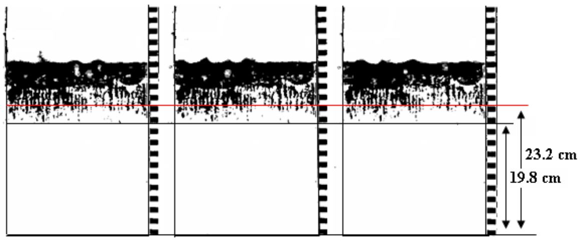

Figure 5: Bubble appearence in the reference experiment.

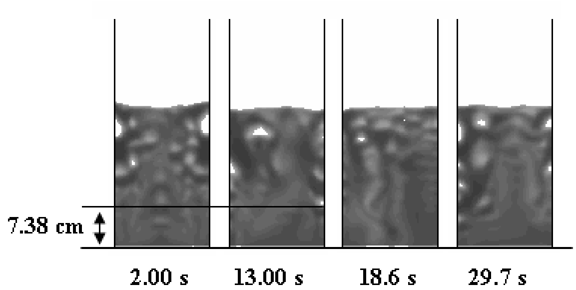

Figure 6: Bubble distribution in the particles bed of the simulation P2.

lations. That is because there is no bubble prediction in the simulation with only one particle phase.

Fig. 7 and Fig. 8 present the bubble distribution in the particle bed of the simulations P3 and P4. Both figures provides that there are not many bubbles at the central area as well as on the walls of the bed when the upper section of the bed is considered. Also the lowest level of bubble appearance is not as low as the previous simulations. Among the simulations P3 and P4, P4 has better prediction of the lowest position predicted is more closer to the reference experiment than any other simulation analyzed.

Bubble velocity (rise velocity) of the simulations and the reference experiment are compared using an averaged velocity value. Rise velocities is useful to study the dynamics in the particle bed and also to compare the prediction of the simulations with the reference experiment to evaluate how close the simulations are to the experiment. Fig. 9 presents the change of the position of a selected bubble with time in the experimental bed. Frame rate of 30 fps have used for filming the experiment, and that rate is used to calculate the bubble velocity. The bubble have a velocity of 0.174 ms^{-1} at the first interval and a velocity of 0.321 ms^{-1} at the second time interval, which gives an average velocity of 0.223 ms^{-1}.

Rise velocities of the bubbles in the simulations are calculated using some of the bubbles raised in

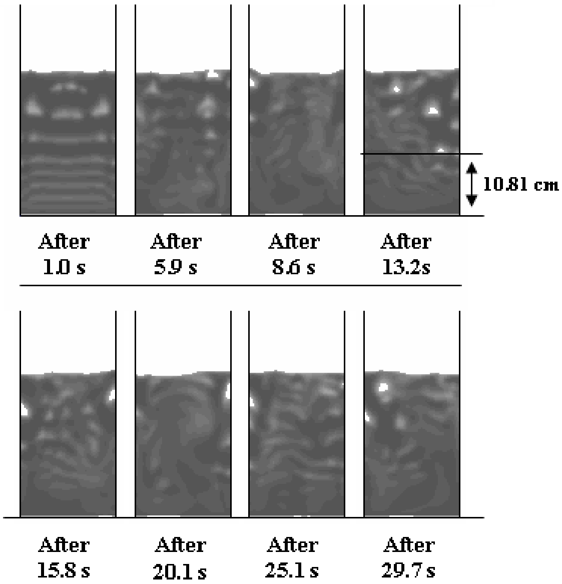

Figure 7: Bubble distribution in the particles bed of the simulation P3.

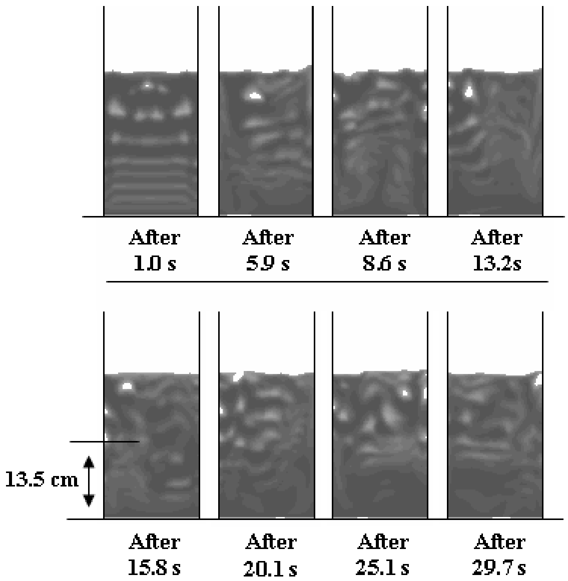

Figure 8: Bubble distribution in the particles bed of the simulation P4.

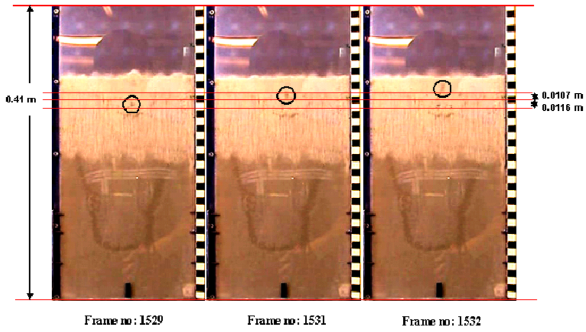

Figure 9: Bubble position with time in the reference experiment.

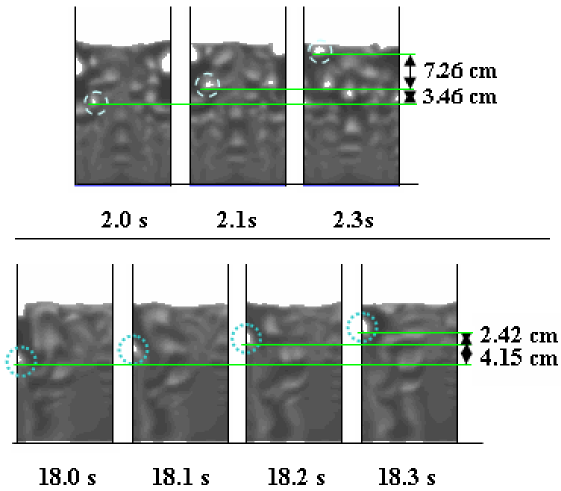

Figure 10: Bubble position with time in the simulation P2.

the particle bed at each simulation. It is performed using contours of VOF of the gas phase. To calculate the rise velocity, one or more bubbles are selected and the change of the position of the bubble with time is measured. Fig. 10 presents the change of the position of two bubbles with time in the simulation P2. Firstly analyzed bubble have an average velocity of 0.357 ms^{-1} and the secondly analyzed bubble have 0.219 ms^{-1}.

All four simulations were analyzed with respective to the rise velocity using the same method. The averaged velocities calculated are presented in Tab. 3.

Above analysis showed that all three simulations which have predicted bubbles, have rise velocities in the same range as the reference experiment. In addition, it was clearly noticeable that the bubbles are growing larger with time and speeds up as the bubbles grow. Also, when the rise velocities are compared with the emulsion gas velocity, it is clear that all of the analyzed bubbles are fast moving

Table 3: Averaged velocities.

Simulation	Averaged rise velocity		
	Bubble 1	Bubble2	Bubble 3
P1	No bubbles		
P2	0.288 m/s	0.219 m/s	
P3	0.393 m/s	0.465 m/s	0.26 m/s
P4	0.254 m/s	0.33 m/s	
Experiment	0.223 m/s		

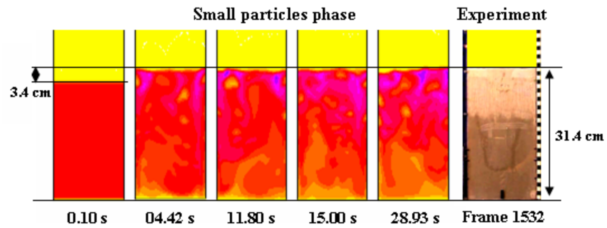

Figure 11: Expansion of the particle bed predicted by the simulation P3.

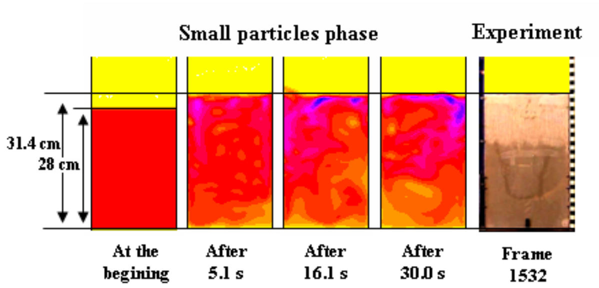

Figure 12: Expansion of the particle bed predicted by the simulation P4.

bubbles.

It is important to realize that the analysis have done only for one bubble from the reference experiment, which is not a precise method. Since the bubbles appeared in the video are not clear enough to be analyzed, a better video will be produced and a similar analysis will be presented in a later publication.

Expansion of the particle bed is also an important factors to check whether a simulation gives reasonable results. If a simulation gives similar bed expansion to that of the reference experiment, the results of the simulation are accepted as a good prediction. To check the reliability of the simulated results of the last four simulations, a bed height analysis is performed. As small particle phase can present the bed height accurately than any other particle phase, only the small particle phase has used for the bed height comparisons.

The previous analysis showed that the simulations P3 and P4 have the closest prediction to the reference experiment. Only those two simulations are used in the bed height analysis. Fig. 11 and Fig. 12 provide the comparison of the predicted bed height by the simulations P3 and P4 with the reference experiment. Analysis of the figures show that both simulation have predicted the bed expansion similar to the experiment, while the simulation P4 has the best prediction.

5 Conclusions

As the simulation with only the single particle phase didn't predict variations in VOF of particles or bubbles in the particle bed it is not used in the analysis. The reason is found as the superficial gas velocity used in the simulations, which is well bellow the theoretical minimum fluidization velocity related to the particle size used in the bed. The comparison of the multiphase simulations with the reference experiment is conducted in terms of the particle segregation, expansion of the particle bed and the bubble characteristics in the particle bed.

Prediction of particle segregation in the simulations is analyzed. The simulations are compared with each other and with the reference experiment. Contours of the particle phases and the VOF data of particle phases along the height of the bed at selected points are used. Comparison showed that the higher the number of particle phases the better the prediction of particle segregation.

Bubble behavior prediction is analyzed in terms of bubble velocity and the lowest position of bubble occurrence in the bed using the contours of the particle phases. The analysis and the comparisons with the reference experiment confirmed that there is an influence on the simulated data from introducing the particle size distributions in the simulations.

The bed expansion in the simulations is presented using the contours of the small particle phase and compared with the reference experiment using a photo frame from the movie of the reference experiment. The comparison showed that the simulation with four particle phases has predicted the bed expansion very close to that of the reference experiment and the prediction is better than all other multiphase simulations performed under this study.

The total comparison of the simulated results with the reference experiment showed that the higher the number of particle phases the better the prediction of particle segregation, bubble behavior and the bed expansion in the simulations.

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