Experimental study on solids mixing and bubble behavior in a pseudo-2D, freely bubbling, gas-solid fluidized bed using PIV and DIA

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Abstract
The hydrodynamics of a freely bubbling, gas-solid fluidized bed has been investigated experimentally with non-invasive measuring techniques in a pseudo-2D column filled with glass beads of 400-600 µm fluidized with air. Particle Image Velocimetry (PIV) combined with Digital Image Analysis (DIA) has been used to determine simultaneously the emulsion phase circulation patterns, bubble hold-up and bubble size and velocity distributions. The combination of DIA with PIV allows correcting for the influence of particle raining through the roof of the bubbles on the time-averaged emulsion phase velocity profiles. The time-averaged emulsion phase circulation patterns have been measured as a function of fluidization velocity. Moreover, with DIA the average bubble diameter and bubble velocity as a function of height and fluidization velocity have been determined and found to correspond reasonably well with literature correlations. The experimental data provides a basis for development and validation of CFD models to describe the solids-mixing in gas-solid fluidized beds.

Introduction
Fluidized bed reactors have found widespread application in the chemical and process industries because of their excellent heat and mass transfer characteristics, originating from the strong solids mixing induced by gas bubbles rising through the gas-solid suspension. Although fluidized bed reactors have been industrially operated since the 1920’s, quantitative knowledge about solids mixing is still rather scarce. Most of the experimental research published in open literature on hydrodynamics in fluidized beds, is focused on either the particulate phase or the gas phase, but rarely on both phases simultaneously despite the strong mutual interactions. Therefore, to extend our fundamental knowledge on the interaction between the two phases and the impact on the macro-scale circulation patterns in a fluidized bed, two non-invasive measuring techniques have been applied: Particle Image Velocimetry (PIV) and Digital Image Analysis (DIA).

PIV was first applied by Bokkers et al. 2004 to measure the emulsion phase circulation patterns in freely bubbling gas-solid fluidized beds, in order to validate particle segregation rates predicted by their 3D Discrete Particle Model (DPM). Link et al. 2004 used PIV to establish fluidization regime maps in spouted fluidized beds and found excellent agreement with their DPM simulations. Dijkstra et al. 2007 extended the PIV technique to enable measurement of the granular temperature distribution simultaneously in the fluidized bed. The granular temperature is a very important parameter in the modeling of fluidized beds with Two-Fluid Models (TFM) based on the Kinetic Theory of Granular Flow (KTFG). PIV has been also used to study particle behavior in the freeboard region (a.o. Duursma et al. 2001), and to investigate bubble eruption at the top of the bed (Muller et al. 2007).

Lim et al. 1990 were the first to perform DIA measurements on a pseudo 2D fluidized bed reactor with which they studied the size, velocity distribution and bubble hold-up distribution of bubbles. Goldschmidt et al. 2003 measured the bed expansion and segregation rates of a binary particle mixture using a high speed color camera. Shen et al. 2004 used DIA to derive relations for the bubble growth and bubble rise velocity in a pseudo 2D bubbling fluidized bed filled with Geldart B particles. Finally, Mudde et al. 1994 used DIA to measure the local hold-up, and bubble size, shape and velocity in a bubbling fluidized bed.

To the authors’ knowledge, both techniques have never been used simultaneously. When applying PIV to measure the time-averaged emulsion phase circulation profiles in gas-solid freely bubbling fluidized beds of Geldart B type particles, it is important to correct for the large velocities associated with particles raining through the roofs of the larger bubbles, which can be achieved by combining PIV with DIA. In this paper the time-averaged emulsion phase velocity profiles have been determined with PIV combined with DIA in a pseudo-2D column filled with glass beads of 400-600 µm fluidized with air as a function of the fluidization velocity. Moreover, the DIA results have been used to determine the average bubble diameter and bubble velocity as a function of height and fluidization velocity. First the experimental set-up and measuring techniques are described, followed by a discussion on the results for the emulsion phase circulation patterns. Finally, the determined averaged bubble size and velocity as a function...
of height and fluidization velocity is compared with literature correlations.

**Experimental**

**Setup**

Since PIV and DIA are both optical, non-invasive measuring techniques that require visual accessibility, the experiments were performed in a pseudo-2D fluidized bed with a height of 0.7 m, width of 0.3 m and a depth of 0.015 m, where the front and the back walls of the column were made of glass. The bed was filled with glass beads of 400-600 µm and air was used as fluidization gas, which was introduced via a porous plate distributor (pore size of 19 µm) where the mass flow rate was adjusted with calibrated mass flow controllers. The air was first humidified with steam (60-70% relative humidity) to avoid electrostatic build-up in the bed. Good illumination was achieved with four lamps directly illuminating the front of the fluidized bed (see Figure 1). An overview of the experimental settings can be found in Table 1.

![Figure 1: Top view of the experimental setup](image)

**PIV**

The basic principle of PIV is to divide the recorded images into interrogation areas and cross-correlate two consecutive images to obtain the average displacement $s_p$ of the interrogation areas. Knowing the time $\Delta t$ between the two images, the average velocity $v_p$ of the particles in that particularly interrogation area can be calculated with

$$ v_p(x,t) = \frac{s_p(x,t)}{M \Delta t} $$

where $M$ is the magnification of the image. Careful selection of the time between two consecutive images is required to minimize influence of out-of-plane movement of particles (see e.g. Westerweel 1997 for further details). By combining the velocities of all interrogation areas, the instantaneous velocity profile pattern is obtained. To determine the instantaneous emulsion phase velocity profiles, DIA is applied on the PIV images (as described in the next section) in order to correct for the presence of the bubbles (i.e. filtering the velocities associated with raining of the particles through the roofs of the bubbles). The thus filtered instantaneous velocity profiles were time-averaged.

Images with a resolution of 1024x1280 pixels were recorded with a LaVision ImagerPro HS CCD camera with an internal memory of 2 GB. For the PIV measurements, the camera was located at such a distance from the front of the bed, that a single particle was represented by at least 2-3 pixels in diameter to obtain the desired resolution, which allowed measuring an area of the bed of 11.8 cm x 15 cm. The time-averaged emulsion phase velocity profiles of the entire bed were determined by repositioning the camera 9 times. The bottom 1.5 cm of the fluidized bed just above the distributor could not be measured due to lack of visual accessibility. The frequency with which the PIV image pairs were recorded was 4 Hz. The exposure time was set to 1 ms with an effective time delay of 5.003 ms between the images in a pair. With this scheme, the camera was able to record for 3 minutes.

<table>
<thead>
<tr>
<th>Table 1: Overview of the experimental settings.</th>
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<tr>
<td><strong>Bed</strong></td>
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<td>Depth [m]</td>
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<td>Height [m]</td>
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<td>Initial bed height [m]</td>
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<td>Right and left wall [-]</td>
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<td>Distributor pore size [µm]</td>
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<td><strong>Particles</strong></td>
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<td>Diameter [µm]</td>
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<td>Minimum fluidization velocity [m/s]</td>
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<td><strong>PIV-DIA</strong></td>
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<td>Frequency [Hz] (PIV)</td>
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<td>Time delay [ms] (PIV)</td>
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<td>Time delay [ms] (DIA)</td>
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**DIA**

The basic principle of DIA is to use the pixel intensity to discriminate between bubble or emulsion phase. If the pixel intensity is below a threshold value, the pixel area is assigned to the bubble phase, and otherwise to the emulsion phase. To correct for inhomogeneous lighting, a relative threshold is applied using an averaged pixel intensity (averaged over an area larger than the largest bubbles). For the analyses in this paper, a relative threshold value of 0.90 has been used. This information is subsequently used to determine the bubble size and porosity distribution. All bubbles were labeled and followed during its path through the fluidized bed. The bubble velocities were determined using the displacement of the equivalent spherical bubbles. In addition, the program detects the
walls and freeboard, and eliminates distortions caused by shadows and inhomogeneous lighting. A typical DIA output is shown Figure 2. The bubble size and velocity distributions were determined by performing DIA on images taken with LaVision ImagerPro HS CCD camera of the entire bed in order to avoid problems associated with bubbles that are captured only partially in the image. This allowed measuring for approximately 30 s using a uniform time delay between the images of 10 ms. The DIA program was validated using “synthetic” images, in which the size and position of the bubbles were known exactly. The calculated bubble sizes and velocities were found to correspond excellently to the given sizes and velocities. Moreover, with a relative threshold value of 0.90 to discriminate between the bubble and emulsion phase the best results were obtained.

Figure 2: A typical bubble plot obtained with DIA from a freely bubbling fluidized bed. a) Original picture; b) Detected bubbles; c) Equivalent bubble diameters.

Results and Discussion

In this paragraph, the results from the PIV/DIA measurements are presented and discussed. First it is demonstrated that it is important to correct the PIV measurements for particle raining using DIA to obtain the time-averaged emulsion phase velocity profiles. Then, the time-averaged emulsion phase velocities are presented as function of the fluidization velocity. Finally, the average bubble diameter as function of the axial position and the average bubble rise velocity as function of the average bubble diameter for different fluidization velocities are compared with literature correlations.

Time-averaged emulsion phase velocity profiles

Time-averaging

To determine the number of image pairs required for time-averaging, four PIV measurements were carried out at the same operating conditions (i.e. in total 12 min.). The averaged velocity profile was determined on the basis of first three series, and was subsequently compared to the time-averaged profile determined from only the fourth series. It was found that averaging using approximately 200 image pairs, the maximum deviation in the time-averaged particle phase velocity was less than 1%. In the experiments 712 image pairs were used to determine the time-averaged velocity field.

Combining PIV and DIA

In Figure 3 (A) a typical (instantaneous) PIV result is shown for a test case where a single bubble was injected into a fluidized bed at incipient fluidization conditions. The figure clearly shows the large particle velocities associated with particles raining through the bubble. To obtain the time-averaged emulsion phase velocity profile these velocities need to be filtered out, which was accomplished using DIA. Figure 3 (B) shows the instantaneous velocity profile of the emulsion phase after disregarding particle velocities inside bubbles.

Figure 3: Instantaneous PIV image with velocity vectors, where in the background one of the images is given on which this particular PIV image in based, A. purely PIV image, B. combined PIV and DIA image.

The influence of the filtering of the particle velocities inside bubbles on the time-averaged emulsion phase velocity profiles can be discerned from Figure 4 and Figure 5. Figure 4 shows the time-averaged emulsion phase velocity profiles before (A) and after filtering (B). The figure clearly shows that without filtering using DIA the up-flow of the emulsion phase in the centre of the fluidized bed is strongly underestimated. Since most bubbles move through the centre of the bed, the effect of the filtering is most pronounced at the centre. The extent of down-flow is hardly affected by the filtering. This is also clearly demonstrated in Figure 5, showing the lateral emulsion phase profiles at two different heights in the bed (at 10.5 cm and 24.5 cm above the distributor). The experimental results for the lower bed height (A) clearly show two peaks in the emulsion phase velocity (at approximately 70 mm from both the left and right wall), corresponding to the lateral movement of the bubbles towards the centre of the bed. The PIV result for the emulsion phase profile without filtering (black line) even indicates no up-flow of the emulsion phase, while after filtering (red line) up-flow is correctly measured. The experimental results for the higher bed height (B) shows up-flow in the centre of the bed, where without filtering the maximum axial emulsion phase velocity is underestimated by a factor 2, while the extent of down-flow is hardly influenced by the filtering.
Figure 4: Time-averaged emulsion phase velocity profiles for 2.5 $u/u_{mf}$. A) before filtering using DIA, B) after filtering using DIA.

Figure 5: Time-averaged lateral emulsion phase velocity profile at 10.5 cm (A) and 24.5 cm (B) above the distributor. Fluidization velocity was 2.5 $u/u_{mf}$. The black and red lines indicate the profiles without and with filtering using DIA.

Influence of fluidization velocity

In Figure 6 the (filtered) time-averaged emulsion phase velocity profiles are given for different fluidization velocities. To keep the figure orderly not all measured velocity vectors are plotted (approximately one out of three measured vectors is shown). It can be seen that for higher fluidization velocities, the emulsion phase velocity patterns become much more pronounced, showing two symmetric vortices with their centers located at the top of the bed. At higher fluidization velocities, the centers of the vortices move towards each other and the up and down-flow become much more distinct. At a fluidization velocity of 2.5 $u/u_{mf}$ the down-flow extends down to the distributor. Finally, it is noted that the emulsion phase velocity above the distributor in the middle of the fluidized bed is directed downwards. The lateral movement of the emulsion phase just above the distributor could not be measured in our set-up because of lack of visual accessibility.
Figure 6: Time-averaged emulsion phase velocity profiles. A) fluidization velocity $1.5 \frac{u}{u_{mf}}$, B) fluidization velocity $2.0 \frac{u}{u_{mf}}$, C) fluidization velocity $2.5 \frac{u}{u_{mf}}$. 
Average bubble diameter and velocity

Average bubble diameter

First the averaged bubble diameter as a function of the height in the bed was investigated for different fluidization velocities (see Figure 7). Figure 7 shows that, as expected, the laterally averaged bubble diameter increases less than proportional with the axial position in the bed and that much larger bubble diameters are found for higher fluidization velocities.

The experimental results were compared with the correlation suggested by Shen et al. 2004, who fitted a Darton-type bubble size equation with their DIA experiments in a much larger pseudo-2D freely bubbling fluidized bed. They performed their measurements in a 0.68 x 0.07 m pseudo 2D fluidized bed using several different particle diameters in the Geldart B group and correlated the average bubble diameter with

\[
d_b = 0.89 \left[ \left( u_s - u_{mf} \right) \left( h + 3.0 \frac{A_0}{t} \right) \right]^2 \frac{g}{t^{0.3}}
\]  

(2)

where \( h \) is the height, \( A_0 \) is the catchment area, \( t \) is the depth of the column and \( g \) is the gravitational acceleration. Their correlation overpredicts our results, especially at higher superficial gas velocities. The difference between the measured averaged bubble diameter in this work and the predictions using the equation by Shen et al. can be explained by the fact that they used a much larger fluidized bed. Not only can the bubbles grow to a larger maximum bubble diameter in their set-up, also the bubbles with a diameter smaller than the bed depth (0.07 m) could not be well detected.

Average bubble velocity

The measured average bubble velocity as a function of the bubble diameter for different fluidization velocities is plotted in Figure 8 and compared with literature findings.

The bubble rise velocity \( u_b \) in a freely bubbling fluidized bed is usually correlated to the bubble diameter and fluidization velocity via

\[
u_b = \left( u_s - u_{mf} \right) + C \sqrt{g d_b}
\]

(3)

where different values between 0.5 and 1 has been proposed for the constant \( C \). Using \( C = 0.5 \) as suggested by a.o. Mudde et al. 1994, the measured bubble velocities agree reasonably well with this correlation. However, the bubble rise velocity is strongly overestimated for very small bubbles, where experimental errors (e.g. the selected relative threshold value) and wall effects play an important role.

Figure 7: Bubble diameter as a function of the axial position in the bed for different fluidization velocities. The solid lines represent the bubble diameter calculated with equation 2, suggested by Shen et al. 2004
Figure 8: Average bubble velocity vs. average bubble diameter at different fluidization velocities. The solid lines represents the prediction using equation 3, using $C = 0.5$.

Conclusions

Two non-invasive measuring techniques were used to measure hydrodynamic properties of a freely bubbling gas-solid fluidized bed. Glass beads with a diameter of 400-600 µm were fluidized with air in a pseudo 2D bed. Particle Image Velocimetry (PIV) combined with Digital Image Analysis (DIA) has been used to determine simultaneously the emulsion phase circulation patterns, gas hold-up and bubble size and velocity distributions. The combination of DIA with PIV allows correcting for the influence of particle raining through the roof of the bubbles on the time-averaged emulsion phase velocity profiles. The significant influence of particle raining on the time-averaged emulsion phase velocity profiles was demonstrated. The time-averaged emulsion phase circulation patterns have been measured as a function of the fluidization velocity. The time-averaged emulsion phase velocity profiles show two symmetric vortices with their centers located at the top of the bed, where at higher fluidization velocities the centers move closer together and the up and down-flow become much more pronounced.

Moreover, with DIA the average bubble diameter and average bubble velocity as a function of height and fluidization velocity have been determined and found to correspond reasonably with literature correlations. The difference between the predicted and measured bubble diameter and bubble velocity could be ascribed to differences in the experimental set-ups used. In the present work a much smaller bed depth was used, which allows detecting smaller bubbles.

The experimental data provide a basis for further development and validation of CFD models to describe the solids-mixing in gas-solid fluidized beds. Future work will focus on the influence of the column width, bed aspect ratio and particle diameter and material on the emulsion phase circulation patterns. Moreover, the experimental technique will be used to elucidate and quantify the interaction forces between the bubbles and their influence on the emulsion phase circulation patterns (see e.g. Laverman et al. 2006).

Nomenclature

- $A_0$ Catchment area
- $C$ Constant
- $d$ diameter
- $D$ Dimensional
- $g$ Gravitational acceleration [m.s$^{-2}$]
- $h$ Height [m]
- $M$ Magnification
- $s$ Displacement [m]
- $t$ Time [s]
- $v$ Velocity [m.s$^{-1}$]
- $u$ gas velocity [m.s$^{-1}$]

Subscripts

- $b$ Bubble
- $0$ superficial
- $p$ Particle
- $mf$ Minimum fluidization
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References


