DETERMINATION OF THE BUBBLE DIAMETER IN THE PRE-ERUPTION REGION OF THE FLUIDIZED BED

by

Ana MICKOVSKA, Zvonimir KOSTIĆ, Valentino STOJKOVSKI and Aleksandar NOŠPAL

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The experimental determination of the bubble diameter is performed in two-dimensional bed with annular cross section, in the bubble pre-eruption region using video-based method. The results showed that the average bubble diameter in the pre-eruption region is increasing with the increase of the total number of fluidization, and it is not influenced by the composition of the binary mixture. Based on the experimental research , new empirical equation for estimation of the bubble diameter in the pre-eruption region is developed.

Fluidization regime was characterized using fluidization number defined as ratio of superficial gas velocity and minimum fluidization velocity.

The minimum fluidization velocity of the mixture is defined in two ways: as apparent minimum fluidization velocity $u_{mf\text{-}P}$ and total minimum fluidization velocity $u_{mf\text{-}T}$. Respectively, for description of the fluidization regime, apparent and total number of fluidization are defined as a ratio between the superficial gas velocity and minimum fluidization velocity: $u_f/u_{mf\text{-}P}$ and $u_f/u_{mf\text{-}T}$.

Introduction

Bubbling fluidized bed has found widespread application because of its favorable properties, its excellent heat transfer characteristics and the convenience of solids handling (coal combustion, production of chemicals, polymerization processes, heat exchange, granulation, coating, adsorption and biomass processes) (Kuipers and Swaaij, 1997).

Bubbles provide the favorable agitation in the bed but also provide partial short cuts for the gas phase passing through the bed leading to less effective gas-solid contacting. The well-mixed state of solids is not always desired and other overall gas-solids contacting patterns may be beneficial in certain applications. So, the main properties of the bubbling fluidized beds are predominantly influenced by the bubble characteristics. This is the main reason why understanding of the formation, rising and eruption of bubbles is essential for bubbling fluidized bed utilization design.

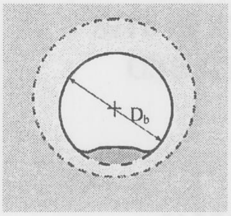

Figure 1. Single bubble in a gas fluidized bed

There are a lot of investigations preformed on bubble properties, its geometry, rising pattern, *etc*. They all agree with the following: the typical bubble shape is nearly spherical with concave bottom. The size of the bubble is defined by the bubble diameter, usually a diameter of a sphere approximating the bubble (Fig. 1). There are many empirical and semi-empirical equations for bubble diameter estimation. Some of them are given in Table 1.

Bubbles appear near the distributor plate, then rise vertically upward and eruptively disintegrate at the bed surface. According to this, the bed can be divided into three regions: (1) region of bubble formation, (2) bubble rising region and (3) region of bubbles disintegration (pre-eruption region) (Fig. 2).

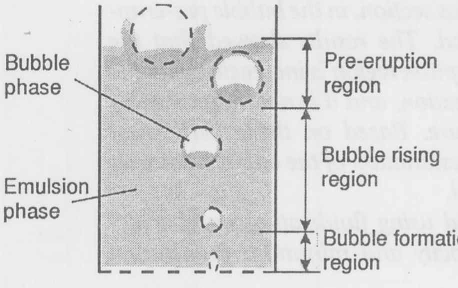

Figure 2. Fluidized bed regions

The eruption of the bubbles at the bed surface leads to ejecting of solids in the freeboard (the space between the bed surface and the gas offtake). Coarse solids and group of solids with terminal velocity larger than the gas superficial velocity can be found above the bed surface traveling up and down. The height in the freeboard region within which the solids loading falls is called transport disengagement height (TDH). This is very important parameter for FB design as entrainment of solids in the exit can be regarded as a nuisance to be endured or advantage even to be essential. It depends of the process of FB utilization (Geldart, 1986). Understanding solids entrainment is getting more complex in case of fluidization of binary mixture.

Investigations had shown that the TDH increases with the bubble size and gas superficial velocity (Zenz, 1983). So, in order to afterward relate the bubble pre-eruption size to TDH, and also the elutriation rate, in case of binary mixtures, an investigation of the bubble diameter in the pre-eruption region of the bed of binary mixture of coarse and fine solids was performed.

The size of the bubble in the pre-eruption region was determined experimentally, and then correlated to the total number of fluidization as a representative of the fluidization regime taking into account the characteristics of the fluidized material. The number of fluidization is defined as a ratio between the superficial gas velocity and the minimum fluidization velocity. The minimum fluidization velocity for the binary mixtures was obtained experimentally. Because of the segregation of binary mixtures the minimum fluidization velocity was defined in two ways: as apparent minimum fluidization velocity $u_{mf\text{-}P}$ and total minimum fluidization velocity $u_{mf\text{-}T}$. Average pre-eruption bubble diameter was presented depending on both apparent and total number of fluidization. Experimental results for the bubble diameter were compared with some published equations. A mathematical approximation of the experimentally obtained bubble diameter depending on total number of fluidization was estimated.

Table 1. Equations for estimation of the bubble diameter D_b, published by different authors (Mori and Wen, 1975; Sung and Burgess, 1987; Kato and Wen, 1969).

Author	Relationship	Notice	Num.
Kobayashi *et al.*, 1965	$D_b = 1.4\rho_p d_p \left(\frac{u_f}{u_{mf}}\right) h$	Kato and Wen (1969)	(1)
Cooke *et al.*, 1968	$D_b = 1.4\rho_p d_p \left(\frac{u_f}{u_{mf}}\right) h + D_0$	Kato and Wen (1969)	(2)
Mori and Wen, 1975	$\frac{D_{bM} - D_b}{D_{bM} - D_{b0}} = \exp\left(-0.265868 \frac{h}{A_s^{0.5}}\right)$ $D_{b0} = 0.347\left[A_s u_{mf}\left(\frac{u_f}{u_{mf}} - 1\right) / n\right]^{0.4}$ $D_{bM} = 0.625\left[A_s u_{mf}\left(\frac{u_f}{u_{mf}} - 1\right)\right]^{0.4}$	$D_s = 30 \div 130$ cm $d_p = 0.006–0.045$ cm $u_{mf} = 0.5–20$ cm/s	(3)
Yasui and Johanson, 1958	$D_b = 1.6\rho_p d_p \left(\frac{u_f}{u_{mf}} - 1\right)^{0.63} h$	glass bead, coal $d_p = 12–240\ \mu$m, $D_s = 15{,}24$ cm	(4)
Park *et al.*, 1969	$D_b = 33.3 d_p^{1.5} \left(\frac{u_f}{u_{mf}} - 1\right)^{0.77} h$	coke $d_p = 86–334$ mm $u_f/u_{mf} = 1.5–10$, $D_s = 10$ cm	(5)
Whitehead *et al.*, 1967	$D_b = 9.76\left(\frac{u_f}{u_{mf}}\right)^{0.006h^{0.54}}$	sand $u_f/u_{mf} = 1–6$ cm/s	(6)
Geldart, 1971	$D_b = D_0 + 0.027 u_{mf}^{0.94} \left(\frac{u_f}{u_{mf}} - 1\right)^{0.94} h$	sand $d_p = 128$ m, $A_s = 68\text{x}1.27$ cm² $D_s = 30$ cm	(7)

Notice: $D_0 = \left(\frac{6G}{\pi}\right)^{0.4} / g^{0.2}$; $G = \frac{u_f - u_{mf}}{n_0}$

n_0 (1/cm²), D_b (cm), h (cm), d_p (cm), A_s (cm²), u_f (cm/s), u_{mf} (cm/s), ρ_p (g/cm³)

Experiments

Experimental installation used for determination of the minimum fluidization velocity and the bubble diameter in pre-eruption region is given on Fig. 3.

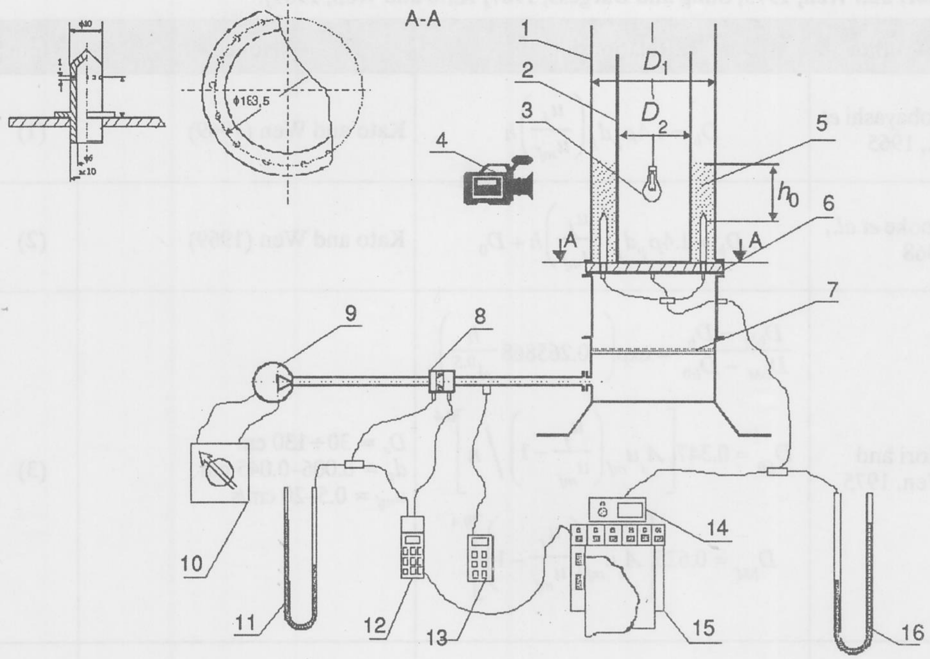

Figure 3. Scheme of the experimental installation used for determination of the fluidization curve and bubble diameter

1 – external tube of the fluidization vessel, 2 – inner tube of the fluidization vessel, 3 – light, 4 – video-camera, 5 – fluidized bed, 6 –distributor plate, 7– settling grid, 8 – measuring diaphragm, 9 – fan, 10 – voltage variator, 11 – U-tube with water, 12 – micro-manometer, 13 – analog/digital temperature converter, 14 – micro-manometer, 15 – chart recorder, 16 – U-tube with water

The experiments for determination of the minimum fluidization velocity and the pre-eruption bubble diameter were made in a cylindrical vessel with annular cross section achieved with two glass tubes. Internal diameter of the external tube is D_1 = 200 mm, and the external diameter of the internal tube is D_2 = 167 mm. The area of the bed is A_s = 9512 mm^2. The air is being introduced in the vessel by distributor plate (3). The distributor plate has 19 nozzles with 12 orifices (1mm) on each nozzle. The pressure above the bed is atmospheric. The height of the fixed bed is h_0 (initial bed height). The construction, design and displacement of the nozzles on the plate are given as a detail A–A on Fig. 3.

There are different methods and techniques for experimental determination of the bubble diameter: direct method (McGrath and Streatfield, 1971), X-rays based methods (Rowe and Matsuno, 1971), laser-based method (Sung and Burgess, 1987), capacitive probe (Cranfield, 1972). Video based method was used in this paper.

Used materials

Two basic fractions of quartz sand were used in this investigation. The material was classified as material of coarse solids and material of fine solids. All magnitudes referring to the coarse solids material are marked with *P*, and to the fine solids material with *F*.

Beside the basic materials, *P* and *F*, percentage mass mixtures from these basic materials were made. Mixtures are marked as: P80/F20, P60/F40 and P40/F60, hence P80/F20 means that 80% of the bed material is coarse particle material and 20% is fine particle material.

Granulation composition of the two basic materials and their percentage mass mixtures were obtained by sieve analysis and are given on Fig. 4 (Stojkovski V., 1995). The values of the estimated equivalent diameters are given in Table 2. Average density of the solids is ρ_p=2660 (kg/m^3).

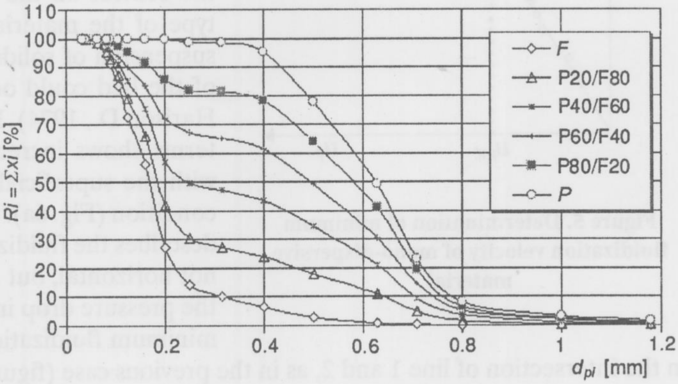

Figure 4. Cumulative remain diagram of the two basic materials, *P* and *F*, and their percentage mass mixtures

Table 2. Equivalent particle diameters, density and voidage of the bed in non-fluidized condition

Material	*F*	P40/F60	P60/F40	P80/F20	*P*
d_{pm} [mm]	0.1487	0.2202	0.2769	0.3788	0.5798
ε_0	0.5188	0.4756	0.4714	0.4699	0.4662
ρ_0 [kg/m^3]	1280	1395	1406	1410	1420

Determination of minimum fluidization velocity

Minimum fluidization velocity was determined from the fluidization curves defined as the dependence of the pressure drop across the bed on the superficial gas velocity, *i. e.*, $\Delta p_{sl} = f(u_f)$. The experimental curves of fluidization were obtained by the method of defluidization, by successive reduction of the superficial gas velocity after the bed has been brought into intensive fluidization. For each material, fluidization curves are determined at three different initial bed heights, h_0.

Fluidized condition of the mono-dispersive fluidizing materials is achieved when the gas superficial velocity reaches certain value necessary to get all solids into floating condition. This velocity is called minimum fluidization velocity and it can be determined graphically from the fluidization curve at the intersection of line 1, which describes the fixed bed, and horizontal straight line 2, which describes the fluidized condition of the bed (Fig. 5).

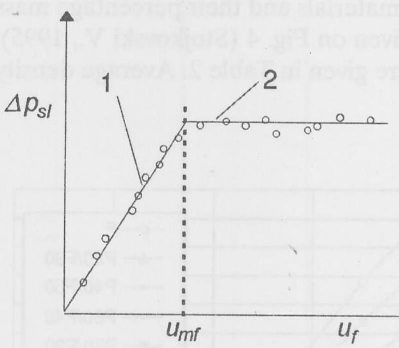

Figure 5. Determination of minimum fluidization velocity of mono-dispersive material

Theoretically, when the bed is fluidized, the pressure drop is equal to the weight of the bed. Thus, the pressure drop is constant when the bed is fluidized (Fig. 5) which is described with the line 2. Because of the influence of the fluidization vessel's shape, displacement of the orifices on the distributor plate and the type of the material that is being fluidized, suspension of solids and uneven fluidization of the bed could occur (Davidson J. F. and Harison D., 1974). Fluidization curve at such terms shows increase of the pressure drop with the superficial gas velocity in fluidized condition (Fig. 6a). In this case, the line 2 that describes the fluidized condition of the bed is not horizontal, but it follows the variation of the pressure drop in fluidized condition. The minimum fluidization velocity is determined in the intersection of line 1 and 2, as in the previous case (figure 6a).

Fluidization curve in case of binary mixtures is characterized by an existence of transition area between the fixed bed regime and fluidized condition of the bed (Fig. 6b). The transition area occurs because not all the solids are brought into fluidized condition at the same superficial gas velocity. In this case apparent and total minimum fluidization velocities could be defined. The value of the apparent minimum fluidization velocity was determined in the intersection of the lines 1 and 2, like in case of mono-dispersive materials. The value of total minimum fluidization velocity is determined at the point of divergence of the fluidization curve from the line 2, when all solids are fluidized. This point is identical to the point of intersection of the fluidization curve obtained by increasing gas velocity and the fluidization curve obtained by reducing gas velocity (Baskakov, 1968).

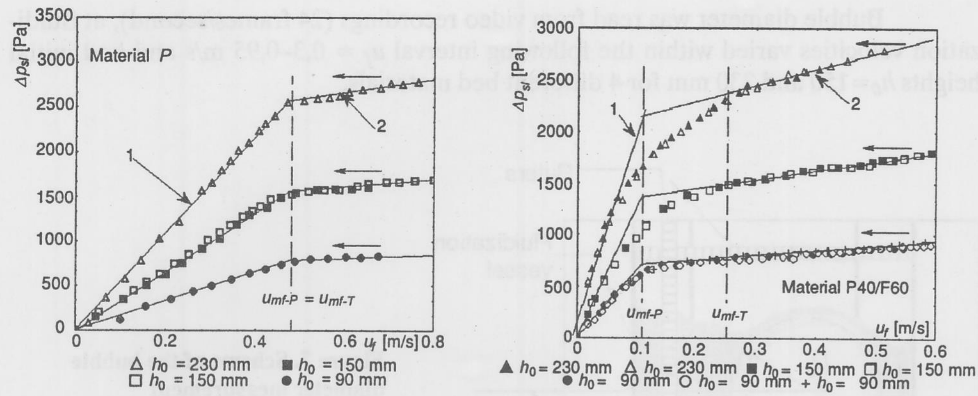

Figure 6. Fluidization curves for materials *P* and P40/F60 at different bed heights.

Based on the fluidization curves, obtained experimentally for all investigated materials, values of the apparent and total minimum fluidization velocities were determined. The last are given in Table 3.

Table 3. Values of the apparent and minimum fluidization velocities for each investigated material

Material	P40/F60	P60/F40	P80/F20	*P*
$u_{mf\text{-}P}$ [m/s]	0.11	0.18	0.34	0.48
$u_{mf\text{-}T}$ [m/s]	0.25	0.37	0.46	0.48

According to the values given in Table 3, apparent and total numbers of fluidization are introduced as a ratio between the superficial gas velocity and the appropriate minimum fluidization velocity, $u_f/u_{mf\text{-}P}$ and $u_f/u_{mf\text{-}T}$.

Measurement of the pre-eruption bubble diameter

The experimental installation is the same as the one for obtaining the fluidization curves. The source of light was placed in the inner tube and the video camera was positioned in front of the fluidization vessel so that it records the pre-eruption region of the bed. The light penetrates through the bubbles and they are observed on the recording as light spots. The recordings were viewed frame by frame. Because of the irregular shape of the bubble, it has been approximated with circle that gives the best approximation to the bubble silhouette (Fig. 7) (Stojkovski *et al.*, 1997).

Bubble diameter was read from video recordings (24 frames/second), at fluidization velocities varied within the following interval u_f = 0,3–0,95 m/s and bed initial heights h_0=150 and 230 mm for 4 different bed materials.

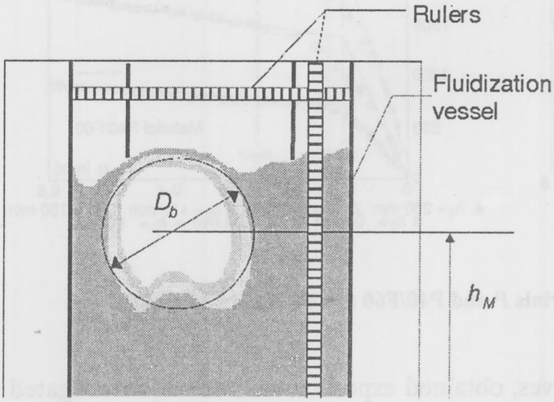

Figure 7. Scheme of the bubble diameter measurement

Experimental results

Bubble diameter obtained by the procedure explained above, was determined for at least five bubbles in the pre-eruption zone at fixed number of fluidization and bed height, h_0. Experimentally obtained data for the bubble diameter at same number of fluidization differs from one another, thus average value of the bubble diameters was taken. Average bubble diameter at given number of fluidization was estimated as arithmetic average from the experimentally obtained values, *i. e.*:

$$D_{bav_i} = \frac{\sum_1^{n_b} D_{b_i}}{n_b}$$

hence, n_b is the number of the measured bubble diameters at fixed number of fluidization. D_{bi} is the measured diameter of a single bubble at unchanged conditions.

The experimentally obtained data for the pre-eruption bubble diameter D_b, compared with the equations given in Table 1, are shown on Fig. 8. Experimentally obtained data for a single measured bubble diameter are presented with light symbols and the average bubble diameter at certain number of fluidization is presented with bold symbols. Equations (1–7), given in Table 1, are inscribed as continuous curves in the diagram, depending on total number of fluidization.

The equations given in Table 1 give the relation between the bubble diameter and the distance from the distributor plate. The distance from the distributor plate at which the bubbles were measured was assumed to be nearly constant and only the number of fluidization and composition of the binary mixture were varied.

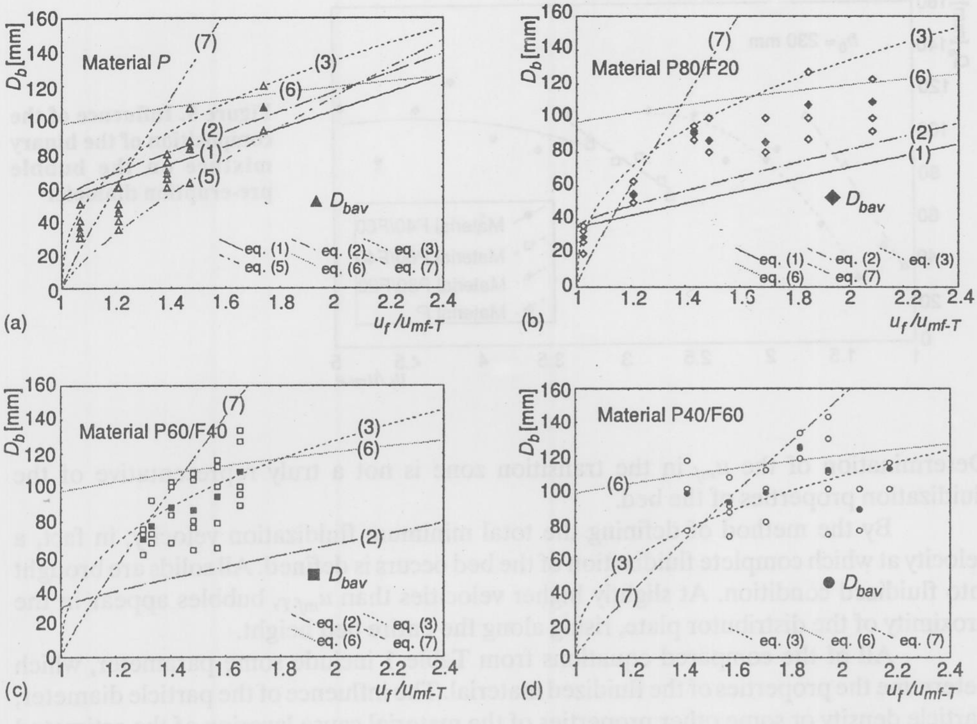

Figure 8 (a–d). Experimentally obtained data for the per-eruption bubble diameter compared with the published equations given in Table 1.

The comparison shows that some of the equations give results close to the experimental data. Experimental bubble diameter for the binary mixtures is larger than the estimated one by the equation (1), (2), (4) and (5). It is assumed that this difference occurs because those equations determine the bubble diameter at certain distance from the distributor plate within the bed but not in the region near the bed surface. The bubble diameter near the bed surface becomes larger due to reduction of the cohesion forces between the solids above the bubble, which is exhibited at small particle materials.

Figures 9 and 10 show the data for the average bubble diameter for all examined materials depending on apparent and total number of fluidization.

The layering of the experimental bubble diameter data when presented depending on apparent number of fluidization arises from the definition of u_{mf-P}. Apparent minimum fluidization velocity, at binary mixtures was defined in the transition area, when the bed is not completely fluidized. The partial fluidization of the bed was noticed visually during the experiments, at fluidization velocity $u_f \approx u_{mf-P}$. At this fluidization velocity, only the upper part of the bed is fluidized, where the material of fine solids is separated. Bubbles are formed in the upper part of the bed, while the rest of the bed remains steady.

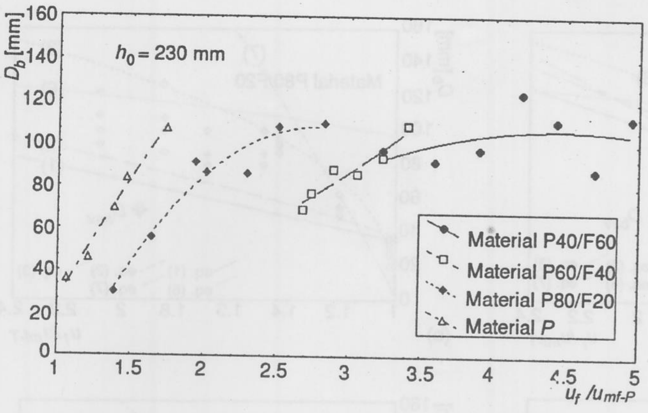

Figure 9. Influence of the composition of the binary mixture on the bubble pre-eruption diameter

Determination of the u_{mf} in the transition zone is not a truly representative of the fluidization properties of the bed.

By the method of defining the total minimum fluidization velocity, in fact, a velocity at which complete fluidization of the bed occurs is defined. All solids are brought into fluidized condition. At slightly higher velocities than u_{mf-T}, bubbles appear in the proximity of the distributor plate, rising along the entire bed height.

All of the compared equations from Table 1 include some parameter, which determine the properties of the fluidized material. The influence of the particle diameter, particle density or some other properties of the material cause layering of the estimated curves. This is not in accordance with the grouping of the bubble diameter data, regardless of the granulation composition of the binary mixture, depending on total number of fluidization (Fig. 10).

A mathematical approximation of the experimental pre-eruption bubble diameter was developed. Fitting of the experimental data for the pre-eruption bubble diameter

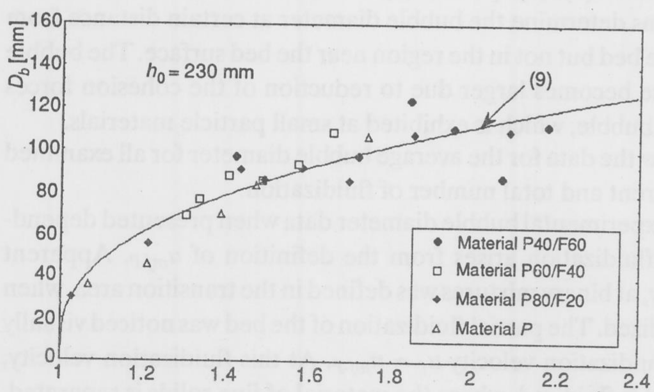

Figure 10. Influence of the composition of the binary mixture on the bubble pre-eruption diameter and the mathematical approximation of the experimental data

is assumed to follow the expression shown bellow:

$$D_b = a + b\left(\frac{u_f}{u_{mfT}} - 1\right)^{0.4} h^c \tag{8}$$

The constant a presents the initial pre-eruption bubble diameter that can be observed in the pre-eruption bed region at total minimum fluidization velocity. The exponent of the number of fluidization is the same as in the equation (3) given in Table 1. As diagrams on Fig. 8 show, and also because of the method at which this exponent was derived (Mori and Wen, 1975), it is assumed the most adequate. Constants $a = 0.48$ cm, $b = 1.49$ cm$^{0.4}$ and the exponent $c = 0.6$, were calculated by least-square method from the experimental data. The final expression is:

$$D_b = 0.48 + 1.49\left(\frac{u_f}{u_{mfT}} - 1\right)^{0.4} h^{0.6} \tag{9}$$

hence D_b cm – bubble diameter, h cm – height of the bed where the bubbles were measured.

The fitting of the expression (9) is presented in the diagram on Fig. 10, for initial bed height h_0 = 230 mm. The correlation between the observed and calculated data is satisfactory (r = 0.91).

The determined equation (9) is compared to the experimental pre-eruption diameter obtained for initial bed height h_0 = 150 mm. The fitting of the estimated correlation is shown on Fig. 11.

The estimated correlation for pre-eruption bubble diameter, equation (9) gives good approximation at h_0 = 150 mm, but for more reliable use of equation (9) it is necessary to compute the constant a, b and c in equation (8) using more variations of the bed height at which bubbles are measured.

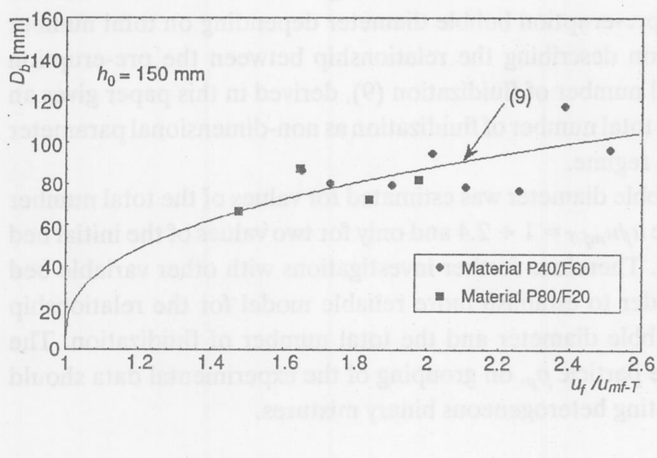

Figure 11. Fitting of the estimated correlation for the pre-eruption bubble diameter at initial bed height h_0 = 150 mm

Conclusions

The analysis of the experimental results leads to the following conclusions.

Experimentally obtained fluidization curves exhibit overall bed pressure drop increase with the superficial gas velocity in the fluidized condition of the bed. This is due to the small thickness of the bed and the position of the nozzles on the distributor plate, causing segregation of solids and uneven fluidization.

Binary mixture was used in the experiment, characterized by transition area from fixed to fluidized condition of the bed. Therefore, apparent and total minimum fluidization velocity were determined. Apparent minimum fluidization velocity is defined in the same way as for mono-dispersive materials and it falls in the region in the transition area, when the bed is segregated and coarse solids are not yet fluidized. Total minimum fluidization velocity is defined when the whole bed is being fluidized. The use of the total minimum fluidization velocity, and herewith the total number of fluidization, is recommended.

The comparison of the experimentally obtained pre-eruption bubble diameter with the published equations determining the relationship of the bubble diameter and total number of fluidization (Table 1), shows that the experimental D_b for the mixtures is larger than the bubble diameter estimated by equations (1), (2), (4) and (5), (Fig. 8). Equation (7) gives close results for smaller numbers of fluidization and equation (6) for larger numbers of fluidization. The equation (3) (Mori and Wen, 1975), gives close results in the whole range of the given total numbers of fluidization for all investigated materials.

Grouping of the pre-eruption bubble diameter experimental data when presented depending on total number of fluidization, regardless of the granulation composition of the used material, is attained. So, it is assumed that all fluidization properties of the fluidized material are introduced by the total minimum fluidization velocity as its truly representative. This leads to the need of establishing a relation that will describe the experimentally obtained pre-eruption bubble diameter depending on total number of fluidization. The correlation describing the relationship between the pre-eruption bubble diameter and the total number of fluidization (9), derived in this paper gives an initiative for further use of the total number of fluidization as non-dimensional parameter for description of fluidization regime.

The pre-eruption bubble diameter was estimated for values of the total number of fluidization in a small range $u_f/u_{mf\text{-}T} = 1 \div 2.4$ and only for two values of the initial bed height $h_0 = 230$ and 150 mm. Therefore further investigations with other variable bed parameters are needed in order to establish more reliable model for the relationship between the pre-eruption bubble diameter and the total number of fluidization. The influence of the density of the particle ρ_p, on grouping of the experimental data should be also examined by investigating heterogeneous binary mixtures.

Nomenclature

A_s – area of the fluidization vessel
D_1, D_2 – inner and outer diameter of the annular fluidization vessel
D_b – bubble diameter
D_{bav} – average bubble diameter
d_p – equivalent particle diameter
d_{pi} – single particle diameter
D_s – diameter of the fluidization vessel
F – small particle material
g – gravitational acceleration
h – distance from the distributor plate
h_0 – initial bed height
h_M – distance of the bubble center from the distributor plate
n_0 – number on holes per unit surface area
n_b – number of observed bubbles
P – large particle material
r – corelation coefficient
u_f – superficial gas velocity
u_{mf} – minimum fluidization velocity
$u_{mf}/u_{mf\text{-}P}$ – apparent number of fluidization
$u_{mf}/u_{mf\text{-}T}$ – total number of fluidization
$u_{mf\text{-}P}$ – apparent minimum fluidization velocity
$u_{mf\text{-}T}$ – total minimum fluidization velocity
p_{sl} – bed pressure drop
ε_0 – voidage of non fluidized bed
ρ_0 – bed density in non-fluidized condition
ρ_p – particle density

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Authors address:
A. Mickovska, Z. Kostić, V. Stojkovski, A. Nošpal
Faculty of Mechanical Engineering, University „Sv. Kiril i Metodij"
P. O. Box 464, 91000 Skopje, Republic of Macedonia

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