# FLUIDIZATION CHARACTERISTICS OF COARSE PARTICLES

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## ABSTRACT

*Fluidization characteristics of coarse particles was investigated using wheat as sample material. Experiments were conducted in a laboratory fluidized bed 15 cm in diameter and made of plexiglass, using different distributor plates with varying open area ratio and orifice diameter. Material was charged into the bed, and air at room temperature with known velocity was blown into the bed to fluidize the material. A video camera was used to record the particle motion. The motion was then analysed using an automatic editing control unit from which the bubble velocity, frequency as well as size could be calculated. Results showed that bubbles formed at about 1cm above the distributor as lenticular cavities, breaking to smaller bubbles as they rose. The bubble rise velocity was found to be 0.5 to 0.8 of the superficial gas velocity and bubbles tended to move towards the centre of the column.*

## INTRODUCTION

Fluidization behaviour of coarse particles is characterized by slow moving but quickly growing bubbles which soon fills the cross section of the bed resulting into slugging.

The use of fluidized bed driers in agricultural products is very advantageous in many aspects. The main limit however is that the material being dried must be fluidizable. The case of coarse particles in fluidized beds has not been very common until recently when Gilshler[1] tried to develop a fluidized bed drier for wheat. His attempt however resulted in the invention of the spouted bed, which ever since has become increasingly popular for coarse particle drying.

McGaw[2] and Cranfield Geldart[3] studied the bubble behaviour in fluidized beds of large particles, the particle sizes were in the range of 1.83 - 3.89mm and 1 - 2mm respectively. This study therefore undertakes the fluidization behaviour of 4 - 5mm particles.

## EQUIPMENT AND PROCEDURE

### Equipment

The equipment used is as shown in figure 1. The column is cylindrically shaped, 15 cm in diameter, and is made of plexiglass with the bottom closed by a distributor plate. It has a discharge port on its side for discharging the material out of the bed or for taking material samples.

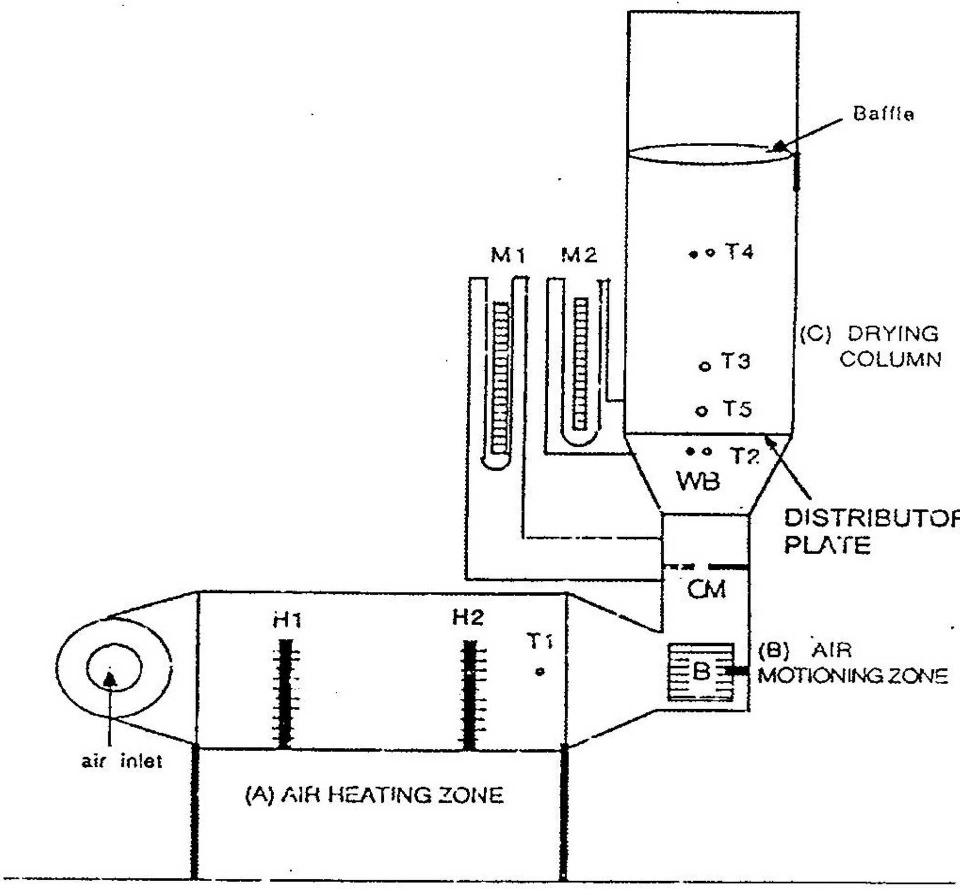


**Fig. 1: Schematic diagram of equipment used**

The gas flow rate was measured using an orifice meter, OM, positioned between the air motioning zone and the drying column. Enough pipe length was provided before the orifice so as to give a fully developed flow before measurements are taken. Readings were obtained from a U-tube manometer, M1; these were then translated to velocities using a Calibration curve. The pressure drop across the distributor and the bed was measured using a U-tube manometer, M2.

Distributor plates with different orifice diameter, open area ratio and pitch were designed and fabricated. The distributors were flat perforated plexiglass plates with orifices arranged on a square pitch. A total of four distributor plates were fabricated with open area ratio ranging from 4.8% to 10.6% and orifice diameter of 1mm and 2mm. Table 1 gives description of distributor plates used.

**Table 1. Description of distributor plates used.**

| Distributor | Open area ratio (%) | Orifice diameter (%) | Pitch (mm) | No. of holes |
|---|---|---|---|---|
| D1 | 6.8 | 1.0 | 2.3 | 1,530 |
| D2 | 6.8 | 2.0 | 5.8 | 384 |
| D3 | 4.8 | 2.0 | 7.8 | 272 |
| D4 | 10.6 | 2.0 | 5.2 | 598 |

## PROCEDURE

The material was charged into the bed, and air at room temperature with a known velocity was blown into the bed to fluidize the material. Distributor plates were changed for each run and wet and dry particles were used for each distributor. A summary of experimental conditions is given in Table 2.

**Table 2: Conditions for studying fluidization characteristics of coarse particles.**

| Run No | Distri-butor | Particle condition | U (m/s) | $\Delta P_T$ (pa) | H (cm) |
|---|---|---|---|---|---|
| 1 | D4 | Dry | 1.2 | 996.0 | 13.0 |
| 2 | D4 | Wet | 1.2 | 1020.9 | 12.0 |
| 3 | D2 | Dry | 1.2 | 1244.6 | 13.0 |
| 4 | D2 | Wet | 1.2 | 1195.2 | 12.0 |
| 5 | D1 | Dry | 1.2 | 1344.6 | 13.0 |
| 6 | D1 | Wet | 1.2 | 1245.0 | 12.0 |
| 7 | D3 | Dry | 0.8 | 1494.0 | 10.0 |
| 8 | D3 | Wet | 0.8 | 1469.1 | 10.0 |

A bed height of 13cm was used for each run of dry particles. Distributor D3, however, had the smallest open area ratio and the bed height of 13 cm could not be fluidized due to the limited capacity of the blower. The bed height was reduced to 10cm, but the fluidization quality was still not satisfactory. Because of this, the study and results reported refer only to distributors D1, D2 and D4.
A video camera was used to record the particle motion, the camera shutter speed was varied from normal speed of 1/250 (i.e 250 frames per minute) to 1/500 and 1/1000. For each shutter speed, the motion was recorded for 1 minute for wet and dry particles. The motion was then analysed with an automatic editing control unit from which the bubble velocity as well as size could be estimated. The editing unit can slow the motion down to 5% of the normal speed and can also count the number of frames and the time taken from a certain required point to another.

To estimate the bubble velocity, the tape was played at 1/20 of the normal speed and bubbles were observed as they formed. To determine the velocity of a particular bubble which may be somewhere higher up in the bed,, the tape was reversed frame by frame until the bubble was at the point of disappearing. Then the frame as well as the time counter were set to zero and the tape was played forward either at 1/20 speed or manually by advancing frame by frame until the bubble reached the reference point marked on the column. The time and number of frames were then recorded. Since the height from the distributor to the reference point was known, the bubble velocity could be calculated.

Estimating the bubble frequency was done by counting the number of bubbles bursting at the top of the bed in a given time. Estimating the bubble frequency from the rising bubbles was not possible since only bubbles on one side of the column (on the camera side) could be seen while those on the other side and at the centre of the column could not. Hence, no attempts were made to relate frequency to bed height. To estimate the bubble size, the tape was put on pause and bubble measurements taken on the video screen. The measurements of the column on the screen (i.e width and height at a selected reference point) were also taken. The bubble measurements taken from the screen were then transformed to actual dimensions by using the ratio of the actual column dimension to those taken from the screen.

## RESULTS AND DISCUSSION

Most bubbles were observed to form at about 1 cm above the distributor plate. The bubble path was not fixed. Most bubbles were visible in the first half of the bed and then disappeared towards the centre of the column, only to be seen again when they burst at the bed surface. A relatively smaller proportion travelled along the column walls until they reached the surface of the bed. There was no fixed pattern of bubble shapes. Most bubbles which formed at the bottom of the bed appeared as horizontal cavities. Some, however, changed to vertical cavities as they rose. Bubbles which appeared a bit higher in the bed tended to be more circular. Most of the bubbles increased in size as they rose (see Table 3), bubbles originated from horizontal cavities which tended to break up into a number of smaller bubbles (in most cases only two).

For a given gas velocity, the bubble size as well as the frequency of formation of cavities decreased with a decrease in grid orifice diameter. Wet particles gave slightly different fluidization behaviour, which was characterized by less cavity formation but more circular bubbles. In general there was less bubbling and more swirling of particles around the bubbles. The observed bubble velocity ranged from 0.5 to 0.8 of the superficial gas velocity depending on the bubble shape and size. Particle mixing was seen to be reasonable, with bubbles bursting from the centre of the column and particles flowing down along the column walls. The bubble vertical dimensions increased more rapidly than the horizontal dimensions as the bubble rose along the column. See Table 3.

Results of this work on fluidization characteristics of coarse particles were compared to those of Cranfield and Geldart(3), Geldart and Cranfield(4) and McGaw(2). The comparison is shown in Table 4.

The comparison reveals a close similarity of the fluidization characteristics of particles used in this work and those used by other workers. However, some differences were also noticed. Whereas Geldart and Cranfield(4) found that bubbles did not appear in the bed until a velocity higher than the minimum fluidization velocity was reached. McGaw(2) found that they appeared at approximately the minimum fluidization velocity. The present work agrees with McGaw(2) in this aspect. Cranfield and Geldart(3) found that a bubble first appeared as a long thin cavity some distance above the distributor; it then moved up the bed before splitting into a number of smaller bubbles. McGaw (2) reported the same formation of bubbles but he observed that the lower part of the bubble tended to remain stationary and only the top part of the bubble rose up the bed. The present work agrees with Cranfield and Geldeart(3) in this aspect.

With respect to physical properties of the particles used, we can see that a range of particle sizes from 1 mm to 5 mm has been covered by these four studies. The observations suggest that as particle size increases, the bubbles form at a lower gas velocity. This can also be substantiated by the work of Broadhurst and Becker(5) and their findings that as particle size increases, the difference between minimum bubbling and minimum fluidization velocities decreases.

The density of particles in this study was almost the same as that used by Cranfield and Geldart(2) and Geldart and Cranfield(4), whereas the density of particles in McGaw's work(2) was almost double that used here. A quick conclusion on bubble measurement would suggest that bubbles formed in beds of less dense particles remain almost stationary. This, however, cannot be absolutely true in this case because the bed used in McGaw's work(2) was a bit too shallow (3.6cm for 1.83 mm particles and 6.3 cm for 3.89 mm particles).

Table 3: Sample bubble measurements. (for wet particles).

| Bubble | Distributor | Frame No. | Bubble width width, (cm) | Bubble height, (cm) |
|---|---|---|---|---|
| A | D4 | 2 | 5.8 | 3.2 |
| | | 4 | 5.2 | 3.2 |
| | | 6 | 6.5 | 5.3 |
| | | 8 | 7.2 | 6.7 |
| | | 10 | 7.2 | 8.8 |
| B | D4 | 4 | 5.3 | 2.2 |
| | | 5 | 6.0 | 4.7 |
| | | 6 | 6.7 | 5.3 |
| | | 7 | 6.7 | 6.1 |
| | | 8 | 7.4 | 7.4 |
| | | 9 | 7.4 | 8.0 |
| | | 10 | 7.4 | 8.7 |
| | | 11 | 7.4 | 9.3 |
| C | D2 | 2 | 5.9 | 4.0 |
| | | 3 | 5.9 | 4.7 |
| | | 5 | 6.5 | 6.7 |
| | | 6 | 5.9 | 7.3 |
| D | D2 | 2 | 5.9 | 3.3 |
| | | 4 | 5.9 | 4.0 |
| | | 6 | 5.2 | 4.7 |
| | | 8 | 4.6 | 5.3 |
| E | D1 | 4 | 4.6 | 4.7 |
| | | 6 | 5.9 | 4.7 |
| | | 8 | 5.9 | 4.7 |
| F | D1 | 3 | 3.3 | 3.3 |
| | | 4 | 4.6 | 4.0 |
| | | 5 | 4.6 | 4.7 |
| | | 6 | 4.6 | 5.3 |
| | | 7 | 4.6 | 6.0 |
| | | 8 | 5.2 | 6.0 |

Deeper beds would probably have given a different observation.

A study on how the distributor plate parameters affect the fluidization quality and hence the drying rate of coarse particles have been undertaken by Katalambulla(6) and results have shown that the distributor plate parameter have a very significant effect on fluidization characteristics and can not be ignored.

## CONCLUSIONS

i Particle mixing was observed to be reasonable, with bubbles bursting from the centre of the column and particles flowing down along the column walls.

ii For a given gas velocity, the bubble size as well as the frequency of formation of cavities decreased with a decrease in grid orifice diameter.

iii There was a slight difference in the fludization behaviour of wet and dry particles. In wet particles there was less cavity formation but more circular bubbles. In general there was less bubbling and more swirling of particles around the bubbles.

iv The fluidization characteristics of the particles investigated (wheat) were similar in many aspects to those reported by other workers. This supports the findings that the difference between the minimum fluidization and minimum bubbling velocities decreases as the particle size increases, as can be seen from the trend of results of Cranfield and Geldard(3), Geldart and Cranfield(4), McGaw(2) and the present work.

## NOMENCLATURE.

H = Bed depth, cm.
U = Superficial gas velocity, m/s.
Umb = Minimum bubbling velocity, m/s
Umf = Minimum fluidization velocity, m/s
$\Delta P_T$ = Total pressure drop across the bed, pa.
$\phi$ = open area ratio
3-D = Three dimensional

**Table 4: Comparison of fluidization characteristics of coarse particles**

| Cranfield & Geldart(3) Geldart & Cranfield(4) | McGaw(2) | Present work |
|---|---|---|
| Equipment | | |
| 3-D 15 cm diameter | 3-D. 29.4 x 5.1cm | 3-D, 15 cm diameter |
| Material: | | |
| Alkalised alumina<br>1-2 mm<br>S.G. 1.38<br>$\phi$= 2.23% | Glass ballotini<br>1.83-3.89 mm<br>S.G. 2.99 | Wheat<br>4-5 mm SG 1.4<br>$\phi$= 4.8-10.6% |
| Bubble formation | | |
| - Form as lenticular cavities breaking to smaller bubble | - same | - same |
| - Form about 5 cm above the grid | - no mention | - Forms at about 1cm or lower. |
| - Form at Umb | - Form at Umf | - Form at Umf |
| - Bubbles rise up the bed | - bubble bottoms remain almost stationary | - Bubbles rise up the bed |
| Bubble rise velocity | | |
| Less than the superficial gas velocity | Same | Same |
| Bubble frequency and concentration | | |
| More bubbles at the bottom than at the top | Same | Same, bubbles move towards the centre of bed due to particles flowing down. |
| Bed expansion | | |
| No expansion up to Umb | Same | Starts below Umf |
| Bubble/bubble interaction | | |
| Cross absorption | No mention | Cross absorption |

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