Bubble formation is a unique feature of Gas-solid fluidized beds. The behavior of such beds is strongly influenced by the presence and properties of bubbles. The objective of this study was primarily to investigate bubble behavior under different sets of operating conditions, observe various aspects of bubble transformations and develop tools and techniques to predict bubble size and shape and bubble size distribution in gas – solid freely bubbling beds.

3.1 RATIONALE FOR SELECTING 2-DIMENSIONAL FLUIDIZED BED

Although the operation of a freely bubbling fluidized bed is totally influenced by bubbles and the various hydrodynamic phenomena associated with bubbles, but information on bubble properties are not easily decipherable in large scale industrial units, as they are not distinctly observable in such units. Hence, to study the various aspects of bubbling phenomena in fluidized beds, a bench scale unit needs to be rigged up and fluidization of particles needs to be investigated under controlled operating conditions.

There are two choices available to investigators studying the bubbling phenomena in gas – solid fluidized beds; one is to use a circular cross-sectional bed and second is to use a rectangular cross-sectional bed. Bubbling behavior in circular cross-sectional bench scale set-up can be investigated using expensive techniques like γ-ray photography or X-ray photography etc. Although the data obtained in circular cross sectional beds is more representative of actual industrial operations, but the techniques used to obtain the data is limited by the cost and the complexity of the set-up.

On the other hand, most investigators prefer to use transparent rectangular cross sectional columns to study the bubble phenomena in Gas-solid fluidized bed where the bubble behavior is explicitly observable and therefore cheaper instrumental techniques can be used to gather data on bubble parameters. Grace and Baeyens (1986) observed that the two-dimensional Gas-solid fluidized bed has proved to be especially useful for the following applications:
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i. Investigations of bubble properties. Bubbles span the bed thickness and hence are readily viewed with the aid of back lighting.

ii. As an educational tool, in helping those unfamiliar with fluidized beds to visualize and understand basic phenomena.

iii. As a means of qualitatively viewing the fluidization characteristics of given particles.

Other advantage of using rectangular geometry is that cheaper imaging techniques like photography, cine photography, are used to evaluate the bubble properties and bubble behavior without the effect of curvature of experimental column coming in picture. However, rectangular shaped fluidized beds are also susceptible to problem similar to that observed in a circular cross-sectional bed when the bubble sizes are small. The clarity of experimental observations of bubbles in a rectangular fluidized bed becomes high only when the bubbles are near the surface of the wall. In view of this behavior, even the earliest investigators on fluidized bed studies realized that true inference of bubble properties are achieved only when one of the dimension of the bed - the width of the bed is reduced to such an extent that it equals the width of a bubble. Hence, the concept of two-dimensional bed came into existence.

There are important quantitative differences between two- and three-dimensional fluidized beds. It’s a two-dimensional bed that allow investigators to observe the rising bubble with the greatest clarity and image these bubbles using photography and various other techniques as detailed in Chapter-2. However, in literature there is no agreement on what should be the width of the fluidized bed so that it can be termed as ‘two-dimensional fluidized bed’. Lyalls (1969) suggested that if mean column thickness is 30 mean particle diameters for spherical particles, then distinction between 2-D and 3-D beds is negligible. In literature the width of the fluidized beds used by various investigators to study different aspects of fluidization phenomena range from 5mm (Rowe et al., 1964) to 63mm (Hiraki et al., 1966).

In this investigation, two-dimensional beds were used to gather information on bubble size, bubble shape, bubble density, bubble rise velocity, bubble-bubble coalescence, bubble splitting and various other phenomena related to freely bubbling fluidized beds. Efforts
were also made to get a suitable criterion wherein a rectangular fluidized bed can be classified as a truly “two-dimensional bed”.

3.2 EXPERIMENTAL SET-UP

The experimental set-up consisted of two-dimensional fluidized beds and a gas-supplying system. The constructional details of the different segments of the bed, gas distributor and gas supply system are listed below. The various imaging tools and probes used along with the instrumentation specific to the imaging tools and probes are detailed in the respective Chapters discussing those features.

3.2.1 Details of different segments of the bed

The two-dimensional beds were made of Perspex sheets (0.004m thick) and rendered leak proof by adequate sealing. These rectangular assemblies were hydraulic tested to ensure the absence of leaks. Three different units were fabricated. All three units had dimensions of length = 1m and width = 0.1m. The thicknesses of these units were different and had dimension of 0.015m, 0.01m and 0.005m respectively. A fourth unit having same length and width but having thickness of 0.03m was also fabricated and used for a limited number of investigations.

Fig. 3.1 Wind box (all dimensions in mm)
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The lower part of the assembly consisted of the trapezoidal shaped wind box for providing uniform gas distribution across cross section as shown in Fig. 3.1. It was built from 0.004 m thick Perspex sheet. The dimensions were 0.078m high, 0.1m wide and thickness were 0.03m, 0.015m, 0.01m and 0.005m based on the size of the main equipment.

The column was supported on the wind box using flanged assembly. Flanges having flange extension of 0.04 m in all directions were fixed to the column as well as the wind box. The flanges were made of 0.008 m Perspex sheet. Further, to provide additional strength to the main column and to avoid breakages, additional 0.005m Perspex sheets were connected at the outer side of the main body as shown in Fig. 3.2.

![Fig. 3.2 Main body of the equipment (not to scale)](image)

To provide perfect vertical alignment to the whole fluidized bed assembly (i.e. main body of the equipment, distributor and wind box) four holes were drilled on the rectangular flange so that bolted joints could be made to hold the assembly together. To avoid solid and gas leakages, the filter cloth distributor was sandwiched between the column and the wind box with appropriate support and rubber gaskets. The rubber gaskets used had thickness of 0.004 m on bottom side (wind box) and 0.002 m on the topside (column side) as shown in Fig. 3.3. The entire assembly was fixed and supported on a metal frame using 19mm diameter threaded connections as shown in Fig 3.4. The alignment of the set-up was checked with a level gage and a scale was provided at the outer surface of the unit to measure the bed height.
To measure the bed pressure drop, a total of three connections were provided along the fluidized bed column at different heights. The first was below the gas distributor, the second was just above the plate and third one was at 36 cm vertically above the gas distributor. The inner end of each pressure probe was fitted with a screen of 40-mesh size to prevent particles to enter inside the probe. Four similar looking set-ups were prepared for bed thicknesses of 0.03 m, 0.015 m, 0.01 m and 0.005 m respectively.

Fig. 3.3 Plan of the wind box – distributor assembly

Fig. 3.4 Snapshots of experimental setup
3.2.2 Details of distributor

The quality of fluidization is strongly influenced by the type of gas distributor. In literature, a number of distributors have been tried out. The features of these distributors are listed in Table 3.1. For satisfactory operation of a fluidized bed, the gas must be distributed uniformly across the bed area. In addition, the distributor must prevent the solids from entering the air/gas distribution network. Further, the distributor must be able to support the pressure drop associated with the gas flow during operation.

Initial experiments were tried out using steel wire gauze (mesh size 170-BS) as a gas distributor in a larger sized column (0.61 m x 0.285 m x 0.01 m). The gas was supplied through a ‘T’ shaped feeder made of 0.25” G.I. pipe whose ends were plugged. It was located exactly 1 cm below the gas distributor. Total seven holes of 2 mm size were drilled in the feeder pipe, one at the centre of the ‘T’ and three on either side of the centre with a pitch of 0.045 m. The disadvantage of this system was that pressure drop across the bed was quite high hence; uniform gas distribution across the bed was not achieved. Further, the openings in the wire gauze were large hence, particles fell through these holes. Moreover, use of ‘T’ shaped feeder led to local spouting and channeling.

Since, this distribution system was not successful and the fluidization quality attained was unsatisfactory with numerous problems associated with gas distribution, it was thought to device a better distributor. After trying out a number of alternatives such as sieve plates with 1 mm holes, fabric filters etc. without much success, finally woven fabric cloth commonly known as ‘Cambric’ was used as the distributor. This fabric has large number of small holes that lead to uniform distribution of gas throughout the cross section of the bed which resulted in excellent fluidization. Thereafter, Cambric cloth was used a distributor for all experiments carried out in this study.

The characteristic of the fabric was determined using ASTM standard D 123. The thickness of the fiber used was 0.27 mm, warp and weft were 96 and 78 per inch respectively. The air permeability was 10 mm water pressure and pore size was 0.28 square inches. The fluidizing air was supplied by using single stage, two-cylinder, air-cooled compressor. This
### Table 3.1 Features of gas distributors

<table>
<thead>
<tr>
<th>S.No.</th>
<th>Type of gas distributor</th>
<th>Special features/advantages</th>
<th>Limitations</th>
</tr>
</thead>
<tbody>
<tr>
<td>1</td>
<td>Perforated or multi-orifice plate</td>
<td>- Cheap</td>
<td>- Lacks rigidity</td>
</tr>
<tr>
<td></td>
<td>Doganoglu et al. (1978), Exxon model IV FCC reactor</td>
<td>- Easy to fabricate</td>
<td>- Deflection is very high under heavy load</td>
</tr>
<tr>
<td></td>
<td></td>
<td></td>
<td>- Possibility of gas leakage at the bed perimeter during thermal expansion (cannot be used at high temperature)</td>
</tr>
<tr>
<td>2</td>
<td>Tuyeres and Caps</td>
<td>- It can be used at high temperature and at highly reactive environmental conditions</td>
<td>- Particles apt to settle, sinter and stick on the distributor plates</td>
</tr>
<tr>
<td></td>
<td>Perry and Chilton (1973)</td>
<td>- Gas distribution is uniform</td>
<td>- Jetting effect is pronounced which causes considerable particle attrition</td>
</tr>
<tr>
<td></td>
<td></td>
<td></td>
<td>- More expensive than perforated plates</td>
</tr>
<tr>
<td>3</td>
<td>Pipe grids or Spargers, Sohio, acrylonitrile production</td>
<td>- Improved Gas-solid contacting by breaking up growing bubbles and by preventing gulfstreaming or gross circulation of solids</td>
<td></td>
</tr>
<tr>
<td></td>
<td></td>
<td>- Additional refined high-resistance distributors are not needed</td>
<td>- Friction between the particles and surface increases the pressure drop (however, increment is often very negligible)</td>
</tr>
<tr>
<td>4</td>
<td>Multiple nozzle injection system, Werther (1978)</td>
<td>- Gas flow is spread uniformly</td>
<td></td>
</tr>
<tr>
<td>5</td>
<td>Filter cloth</td>
<td>- cheap</td>
<td>- cannot be used for industrial operations due to,</td>
</tr>
<tr>
<td></td>
<td></td>
<td>- easily available</td>
<td>a) low construction strength</td>
</tr>
<tr>
<td></td>
<td></td>
<td>- provides uniform distribution of gas across the bed (thus excellent Gas-solid contacting)</td>
<td>b) low resistivity against thermal stresses</td>
</tr>
</tbody>
</table>
compressor was equipped with a 0.4 m$^3$ horizontal tank, complete with a safety valve, air gauge and drain valve. The maximum compressor capacity was to deliver 0.07 m$^3$/min at 1.5 atm. The compressed air was passed through commercial dehumidifier and filtered to remove oil vapors. The gas flow rate was measured using rotameters that were calibrated prior to use. The isometric view of the experimental layout is shown in Fig. 3.5.

![Fig.3.5 Isometric view of the complete fluidization assembly (not to scale)](image)

### 3.3 PARTICLES AND THEIR CHARACTERISTICS

#### 3.3.1 Rationale of selection of particles

Particle size and shape have a strong influence not only on bubble formation and various bubble parameters but also, particle-gas interactions, particle-particle interactions and particle-wall interactions, which play a key role in deciding the overall performance of fluidized beds. The behavior of fluidized bed largely depends on mean particle size and particle density hence, the bubbling behavior in beds of different sized particles is significantly different and likewise the bubbling behavior in beds of different density particles is also different.

In a bed of Geldart group A particles, bubbles rise more rapidly than the rest of the gas which percolates through the emulsion. These gas bubbles as they rise through the bed coalesce and split frequently, there exists a maximum stable bubble size for group A.
particles. While with group B particles, small bubbles form at the distributor at the minimum fluidization velocity and grow and coalesce as they rise through the bed, bubble size increases almost linearly with distance above the distributor and excess gas velocity \((U_0 - U_{mf})\). Bubble size is nearly independent of mean particle size. Geldart group C particles are difficult to fluidize because of their very small size while Geldart group D particles are the largest sized particles. Beds of group D particles tend to spout further in such particle beds the bubbles coalesce rapidly and grow to larger sizes, often slugs are formed. The bubbles rise slowly in comparison to the gas percolating through the emulsion. In the present work the bubbling behavior of particles belonging to all groups with the exception of group C were investigated. Investigations were carried out with regular (spherical) as well as irregular shaped particles having nearly same size and densities, also with particles having wide density differences. A fairly wide spectrum of particles belonging to Geldart A, B and D groups were chosen in this investigation. Ease of availability was also a criterion for selection of the particles studied.

Regular shaped particles selected in this work were, resin (polystyrene), glass-beads and mustard, all were almost perfectly spherical but had different particle densities \((\rho_p = 1310 \text{ kg/m}^3 \text{ (resin)}; \rho_p = 2480 \text{ kg/m}^3 \text{ (glass-beads)}; \rho_p = 1021 \text{ kg/m}^3 \text{ (mustard)})\). They were available in different sizes. Sand was the fourth particle, it was irregular in size and density was closer to glass beads. FCC catalyst particles were also used in some experiments, these were typically group A particles having mean particle diameter, \(d_{pm} = 150 \mu\text{m}\) and particle density, \(\rho_p = 1440 \text{ kg/m}^3\).

**3.3.2 Physical characterization of particles**

Each of the type of particles used in this work were initially sieved for sufficient time using IS 460 type sieves of different aperture sizes using a sieve shaker. The particles collected between adjacent sieves were collected and microscopically measured to obtain the precise particle size. For e.g. particles passing through a 48 mesh screen but retained on a 65 mesh screen i.e. \((-48 +65)\) were collected, and from this fraction, randomly 40-80 particles were separated. The dimension of each particle was measured using a microscope having magnification of 500 X supplied by BESTO, India. The mean surface/volume diameter of the particles was evaluated using following equation:
### Table 3.2 Summary of particle properties and operating environments

<table>
<thead>
<tr>
<th>S. No.</th>
<th>Particle type</th>
<th>Mean particle diameter, (µm)</th>
<th>Particle density, (kg/m³)</th>
<th>Sphericity (φs)</th>
<th>Min. bed voidage, (εmf)</th>
<th>Min. fluid. velocity (Experimental), Uₘᵣf, (m/s)</th>
<th>Min. fluid. velocity (Ergun's eq.), Uₑᵣ, (m/s)</th>
<th>Geldart's group</th>
</tr>
</thead>
<tbody>
<tr>
<td>1.</td>
<td>Resin</td>
<td>220</td>
<td>1310</td>
<td>1</td>
<td>0.286</td>
<td>0.0083</td>
<td>0.0075</td>
<td>B/A</td>
</tr>
<tr>
<td>2.</td>
<td>Resin</td>
<td>450</td>
<td>1310</td>
<td>1</td>
<td>0.282</td>
<td>0.033</td>
<td>0.0298</td>
<td>B</td>
</tr>
<tr>
<td>3.</td>
<td>Resin</td>
<td>780</td>
<td>1310</td>
<td>1</td>
<td>0.267</td>
<td>0.0667</td>
<td>0.07</td>
<td>B</td>
</tr>
<tr>
<td>4.</td>
<td>Sand</td>
<td>250</td>
<td>2694</td>
<td>0.84</td>
<td>0.405</td>
<td>0.05</td>
<td>0.0475</td>
<td>B</td>
</tr>
<tr>
<td>5.</td>
<td>Sand</td>
<td>350</td>
<td>2694</td>
<td>0.78</td>
<td>0.394</td>
<td>0.0667</td>
<td>0.071</td>
<td>B</td>
</tr>
<tr>
<td>6.</td>
<td>Glass-beads</td>
<td>256</td>
<td>2480</td>
<td>1</td>
<td>0.357</td>
<td>0.05</td>
<td>0.041</td>
<td>B</td>
</tr>
<tr>
<td>7.</td>
<td>Glass-beads</td>
<td>460</td>
<td>2480</td>
<td>1</td>
<td>0.339</td>
<td>0.1</td>
<td>0.106</td>
<td>B</td>
</tr>
<tr>
<td>8.</td>
<td>Glass-beads</td>
<td>640</td>
<td>2480</td>
<td>1</td>
<td>0.338</td>
<td>0.183</td>
<td>0.188</td>
<td>B/D</td>
</tr>
<tr>
<td>9.</td>
<td>Mustard</td>
<td>1002</td>
<td>1021</td>
<td>1</td>
<td>0.364</td>
<td>0.216</td>
<td>0.221</td>
<td>D</td>
</tr>
<tr>
<td>10.</td>
<td>Mustard</td>
<td>2084</td>
<td>1021</td>
<td>1</td>
<td>0.337</td>
<td>0.433</td>
<td>0.443</td>
<td>D</td>
</tr>
<tr>
<td>11.</td>
<td>Catalyst</td>
<td>150</td>
<td>1440</td>
<td>0.89</td>
<td>0.38</td>
<td>0.01</td>
<td>0.0081</td>
<td>A</td>
</tr>
</tbody>
</table>

**Operating Environments:**
- Temperature range: 298°K to 303°K
- Air viscosity: $1.82 \times 10^{-5}$ kg/m-sec
- Air density: 1.2 kg/m³
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\[ d_{\text{sv}} = \frac{N_{p1}d_{p1}^3 + N_{p2}d_{p2}^3 + \ldots}{N_{p1}d_{p1}^2 + N_{p2}d_{p2}^2 + \ldots} \]  

(3.1)

Where, \( N_{pi} \) is the number of particles with size \( d_{pi} \). Knowing the size and number of particles in each fraction the mean particle diameter \( d_{sv} \) was determined.

With the exception of sand the mean particle diameter of all other particles used in this work were determined by the above method. For sand particles due to their coarse shape, arithmetic mean of particle was taken as the mean particle diameter. Scanning Electron Microscopy (JEOL JSM-6380LV and PHILIPS XL30) was also done to estimate particle size and get a visual idea of sphericity. Fig. 3.6 shows the SEM pictures of the different particles used in this study. Densities of the particles were measured by the displacement technique using specific gravity bottle. The measured densities of the particles are reported in Table 3.2 and they closely match the densities of similar materials reported in literature.

### 3.3.3 Minimum fluidization velocity of particles

The minimum fluidization velocity (U\(_{\text{mf}}\)) was determined experimentally by observing the pressure drop across the bed as a function of gas flow rates. Gas was passed through the bed of solids and the pressure drop and bed expansion was noted till the bed was fluidized. Alternatively the bed of particles was initially fluidized vigorously to break down agglomeration between particles, then the gas velocity was decreased in increments and the pressure drop across the bed was recorded. Plotting (-\(\Delta P\)) versus \(G\), the minimum fluidization velocity was determined. No significant hysteresis effect was observed with the chosen particles.

It was found that U\(_{\text{mf}}\) values varied negligibly with an increase in the static bed heights in the range of static bed height selected in this investigation (0.21m, 0.25m and 0.29m). This was in contradiction to the observation of Cranfield and Geldart (1972) who observed an increase in U\(_{\text{mf}}\) with increase in static bed heights.

Ergun’s equation was also used to calculate the minimum fluidization velocity of particles by knowing the particle size, particle density, bed height and pressure drop at different gas
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(f) 640 fim glass-beads
(e) 460 fim glass-beads
(d) 256 um glass-beads

(g) 1002 fim mustard
(h) 2084 fim mustard
(i) 150 fim catalyst

Fig. 3.6 SEM pictures of particles
velocities. The experimentally obtained minimum fluidization velocities for different particles as well as the minimum fluidization velocities calculated using the Ergun’s equation were reasonably in close agreement (Table 3.2). A summary of the particle properties and the operating environments used in this study are also reported in Table 3.2.