CHAPTER 3

EXPERIMENTAL INVESTIGATION

3.1 GENERAL

The turbulent bed contactor (TBC) has been recognized as one of the best gas-liquid contacting devices today. This is primarily due to the relative high interfacial mass transfer rates of the TBC when compared to conventional fixed-bed contactors at equivalent gas or liquid flow rates. Because of the presence of moving spheres, the TBC has the ability to handle gases and liquids containing greater amounts of particulate matter as compared to conventional packed bed contactors. In the present work, a three-phase countercurrent fluidized bed system has been fabricated and investigated.

3.2 SELECTION OF TEST SET-UP

3.2.1 GAS-LIQUID-SOLID PHASES

i) Air in gaseous phase and is used as the fluidizing media;

ii) Water in liquid phase which is to be cooled down, is sprayed from the top of the column; and

iii) Low density, large diameter and polystyrene spheres as fluidizing particles.
Good fluidization can be achieved only when each of the three elements, property of fluidizing particles, fluidizing conditions and fluidized bed construction, are properly selected.

3.2.2 COLUMN SIZE

The investigation conducted by Werther [1974] shows that below a bed diameter of 200 mm wall effect exists. The bed diameter has a significant effect on the overall state of fluidization due to the fact that the annular zone gradually moves to the centre with increasing bed height. The bed will exhibit fully developed slugging if its diameter is 100 mm and the transition to slugging will be almost complete if the diameter is 200 mm. According to Han et al. [1990], the gas hold up decreases with increasing column diameters upto 200mm and the liquid hold up shows a reverse trend. Kato et al. [1981] had indicated that the heat transfer coefficient is minimum for a column diameter of 120 mm. Therefore, for the present investigation, a column of 200 mm square section is chosen.

3.2.3 FLUIDIZING PARTICLES

The cooling tower packing, also called fill or contact media, where the heat & mass transfer and the related cooling of water take place, is the heart of the cooling tower. The cooling tower packing plays a dual role - firstly, to increase the effective contact between the air & the water to promote better heat & mass transfer and secondly, to provide least resistance to the movement of air to reduce pressure drop.
The turbulent bed contactor (TBC) is characterized by the use of an essentially non-flooding packing consisting of low density spheres placed over supporting-distributor grids sufficiently far apart to permit turbulent and random motion of the spheres, thus allowing both high velocities of gas & liquid at modest pressure drop. These turbulent conditions result in high capacity and efficiency for a given tower volume. This equipment is essentially non-clogging and can be used when a particle phase is present. The fluidizing particles are presumed to generate turbulence and thus enhance the rates of interfacial heat & mass transfer. Hence the power required, to keep these particles in a fluidized condition, depends to a greater extent on the density of the particles.

Chen & Douglas [1969] had stated that poly-styrene-foam spheres have been found to be satisfactory for many applications. Because of low density of the particles, the particle phase is easily set in motion by upward flow of the gas phase, the ease of the motion being aided by the downfall of the liquid phase. Chang et al. [1986] reported that in three-phase fluidized beds, it is known that the smaller particles \(d_p<2.5\) mm provides bubble coalescence while the large particles \(d_p>2.5\) mm disintegrate bubbles. O'Neill et al. [1972] reported that for fluidization to be possible the fluidizing particles must be light and the bed must be free to expand. The gas-liquid contacting in such units is excellent because of the high interfacial shear stress, and also of the motion of the packing particles themselves.
Strigle [1987] had stated that the pressure drop produced in the actual packed bed must be experimentally determined and the fluidizing particle shape to be developed by is an empirical art. Skold [1981] stated that with the introduction of film type packing, cooling tower is able to handle higher water flow rates, increasing the effectiveness of the cooling tower. Gel'perin et al. [1968] had found that apparatus with fluidized beds of spherical packing has at least 5 or 6 times the output. Geldart [1989] had indicated that spherical smooth particles give better fluidization than rough, angular particles.

Hollow spheres of different diameters & densities made of light polystyrene material, inert in nature, are selected as the fluidizing particles to give better results. The properties of the various particles chosen are given in Table 3.1.

<table>
<thead>
<tr>
<th>Fluidizing particle</th>
<th>Diameter $d_p$ (mm)</th>
<th>Weight/particle $m$ (g)</th>
<th>Density $\rho_p$ (kg/m$^3$)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Particle 1</td>
<td>25</td>
<td>2.602</td>
<td>318.29</td>
</tr>
<tr>
<td>Particle 2</td>
<td>25</td>
<td>1.432</td>
<td>175.67</td>
</tr>
<tr>
<td>Particle 3</td>
<td>23</td>
<td>1.115</td>
<td>175.67</td>
</tr>
<tr>
<td>Particle 4</td>
<td>21</td>
<td>0.849</td>
<td>175.67</td>
</tr>
<tr>
<td>Particle 5</td>
<td>19</td>
<td>0.629</td>
<td>175.67</td>
</tr>
</tbody>
</table>
3.3 EXPERIMENTAL SET-UP

3.3.1 CONSTRUCTION

The experimental set-up is shown in Fig. 3.1. The set-up is fabricated to conduct studies on hydrodynamic behaviour and to determine heat & mass transfer coefficients and tower characteristics. The main column is a 4 mm thick transparent acrylic plastic of 200 mm square cross-section and 1200 mm height, to facilitate visual examination of the bed. The corners of the acrylic sheets are cemented and joined with aluminium angles at the edges. A drift eliminator, made up of steel baffles, is placed on the top of the column, to prevent elutriation of particles and to trap the small drops of water which would otherwise be carried away by the air, hindering the measurement of humidity of the outlet air. The primary function of the supporting-distributor grid is to introduce the fluidizing fluid into the bed evenly and to support the fluidizing particles. An important function of the distributor is to promote uniform fluidization by applying a stabilizing effect on the gas distribution.

The supporting-distributor grid plates are designed as illustrated by Prabir Basu [1984] and fabricated using acrylic plastic sheets, with evenly spaced holes of different diameters as given in Table 3.2. A calming-section has been provided below the supporting-distributor grid to ensure uniform flow of air entering throughout the cross-section of the column for uniform & stable fluidization. The supporting-distributor grid is situated between the main column and
FIG. 3-1 SCHEMATIC DIAGRAM OF EXPERIMENTAL SET-UP
the calming section. The details of design of supporting distributor grid is shown in Appendix 2.

**TABLE 3.2 CHARACTERISTICS OF SUPPORTING-DISTRIBUTOR GRID**

<table>
<thead>
<tr>
<th>Grid</th>
<th>Dia. of Orifice $d_{or}$ mm</th>
<th>No. of holes $n$</th>
<th>Free cross-section $f$</th>
</tr>
</thead>
<tbody>
<tr>
<td>Grid 1</td>
<td>14.50</td>
<td>100</td>
<td>0.41</td>
</tr>
<tr>
<td>Grid 2</td>
<td>18.00</td>
<td>81</td>
<td>0.51</td>
</tr>
<tr>
<td>Grid 3</td>
<td>17.00</td>
<td>100</td>
<td>0.57</td>
</tr>
<tr>
<td>Grid 4</td>
<td>16.15</td>
<td>160</td>
<td>0.82</td>
</tr>
<tr>
<td>Grid 5</td>
<td>15.20</td>
<td>196</td>
<td>0.89</td>
</tr>
</tbody>
</table>

3.3.2 MEASUREMENTS

3.3.2.1 Water Flow Arrangement

Water is uniformly distributed across the cross sectional area of the column by means of a centrifugal type of nozzle distributor. One water reservoir made of thick plastic material, is kept at the bottom of the tower to collect the cooled water from the column. The reservoir tank is provided with immersion water heaters to heat the water to the required temperature level. A variac is used to regulate the temperature of inlet water. A monoblock water pump is used to pump water from the reservoir to the top of the tower column for spraying over the bed of particles. Water flow to the bed can be controlled to any value using the main and bypass valves. The quantity of water can be measured from calibrated flow meter. To minimise the effect of
freeboard and the column wall effect, a provision has been incorporated to adjust the elevation of the spray nozzle mouth, such that the spray is exactly over the particles in the bed and uniform distribution of water is maintained.

3.3.2.2 Air Flow Arrangement

The fluidizing media, air, is supplied to the bed through a centrifugal air blower, of capacity 2.2 kW, duct work and calming-section. The control of required quantity of air is achieved by means of a throttle valve in the air flow line. A bypass line is connected just before the throttle valve so as to avoid any possible harm to the air blower, while operating at very low air flow rates. The quantity of air flow to the bed is measured using pitot tube. Using the calibration curve the quantity of air flow is found out. An electrical air heater is used to control the temperature of inlet air.

3.3.2.3 Temperature and Pressure Measurement

The measurement of temperature distribution along the column height is done using iron-constantan (Fe-K) thermocouples. The thermocouples are evenly embedded at 50 mm intervals on the wall of the test section. In the test section, a column of twenty holes are provided for thermocouples. A digital microvoltmeter is used to measure the induced thermo emf and using calibrated curve the corresponding temperatures are obtained. Similarly, the pressure is measured using manometers connected to the pressure taps, evenly spaced at 75 mm intervals on the wall of the section. Constrictions in
the manometer lines help to damp out the fluctuations in the manometer readings.

The temperature of water inlet to the test section and the water coming out of the column are measured accurately with iron-constantan (Fe-K) thermocouples embedded in position. The inlet water temperature is measured at the spray nozzle, which is just above the fluidized bed. The outlet water temperature is measured just below the calming section. Fe-K thermocouples are provided at air inlet duct (just at the entry to the test section) and at the top of the test section (just above the drift eliminator), as shown in Fig. 3.2. A psychrometer is used to measure the wet-and dry-bulb temperatures of the inlet air and outlet air. A hygrometer is used to find the relative humidity of the air at entry and exit of the column.

3.4 EXPERIMENTAL PROCEDURE

Polystyrene spheres of the desired size are dry-packed to the height required. The water lifting pump is started and the water flow rate is adjusted to the appropriate value. The water heaters are switched on and the temperature of the water at the inlet to the column is adjusted to the appropriate value. The centrifugal air blower is started to send the atmospheric air into the test section, through the air heater, the calming section and the supporting-distributor grid. The temperature of air inlet is adjusted to the predetermined value. This air fluidizes the fluidizing particles. By adjusting the air mass flow rate, the minimum fluidization velocity is found out. The minimum fluidization velocity is also determined by extra-
1. THERMOCOUPLE TO MEASURE WATER INLET TEMPERATURE
2. THERMOCOUPLE TO MEASURE WATER OUTLET TEMPERATURE
3. THERMOCOUPLE TO MEASURE WET- AND DRY-TEMPERATURES OF INLET AIR
4. THERMOCOUPLE TO MEASURE WET- AND DRY-TEMPERATURES OF OUTLET AIR

FIG. 3.2 SCHEMATIC REPRESENTATION OF INSTRUMENTATION OF EXPERIMENTAL SET-UP
polating the plot of 'H/V' against 'U_g' to the value corresponding to the 'H/V'=1. Then the air mass flow rate is increased to the appropriate value to increase the fluidization so that the water and air get mixed to create turbulence. Now, the set-up works at a particular constant air and water flow rates.

After reaching steady state, the pressure drop & temperature along the bed, air/gas mass flow rate, 'G', water/liquid mass flow rate, 'L', fluidized bed height, 'H', inlet and outlet water temperatures, wet- & dry-bulb temperatures of inlet & outlet air are measured.

The experiments are repeated for the variation of the following parameters using the different particles listed in Table 3.1 and for the various supporting-distributor grids as shown in Table 3.2:

1. the air/gas velocity in the free cross-section of the column 'U_g' from 0.60 to 1.75 m/s;

2. the liquid/water velocity in the bed, 'U_i', from 0.0002 to 0.002 m/s (mass flow rate 'L', from 720 to 7200 kg/m^2 h);

3. the packed bed height of the charged particles 'V', from 100 mm to 500 mm;

4. the inlet water temperature to the column 'T_i', from 313 K to 323 K;

5. the inlet air temperature 'T_dbi', from 303 K to 313 K;
3.5 OBSERVATION

The visual observations noticed while conducting the experiments are listed in the following paragraphs.

3.5.1 Effect of Gas Flow

It has been observed that where the gas stream is broken into small bubbles, an increase in the gas flow rate brings about a further expansion of the fluidized bed. The maximum bed height is achieved when a bubble reaches the break-through position, because prior to the break-through the top surface is moving upwards with velocity \((U-U_{mf})\).

The bed is characterized by intensive mixing of the packing, which undergoes a jerky motion over the height of the bed. The fluidized bed height, \('H'\), depends on the gas velocity, \('U_g'\), the liquid mass flow rate, \('L'\), and packed bed height, \('V'\). At incipient fluidization, the bed voidage, \('\varepsilon'\), is equal to the maximum porosity of a fixed bed, and the pressure drop across the bed due to friction is equivalent to the net force exerted by gravity on the fluidizing particles. As the gas velocity is increased beyond the incipient fluidization velocity, bed expansion occurs and causes the pressure drop due to friction to remain essentially constant.

Indeed, introduction of a gas stream into a liquid-particle fluidized bed increases the turbulence and intensity of mixing in the radial and axial directions, and these play an essential role in increasing the heat transfer coefficient.
Uniform bubbling persists to higher gas rates in beds of large (>3mm) particles since large particles break up the bubbles. It is well known that in gas/liquid bubble columns coalescence increases with liquid viscosity, presumably through its effect on the bubble wake.

3.5.2 Effect of Liquid Flow

Typically, it is found that as the liquid flow increases, so the pressure drops and liquid saturation of the bed increases. Visually, it appears that slugs of liquid are passing in rapid succession up the bed but the fluctuations occur too quickly to be followed easily by the eye. With increasing liquid flow, jumping movements in the top 40 mm of the bed were observed, and as the liquid flow is increased still further, the upper zone in which this jumping is occurring extended further down the bed. Although the jumping movements were only a few mm in amplitude, they were sufficient to consolidate the bed appreciably.

In lower ranges of operation, liquid flows quietly down over the particles in a continuous film, gas being the continuous phase. When the gas flow rate is gradually raised with constant liquid flow rate, the fluidizing particles mobilized, but a point is reached at which the tower is slowly filled with liquid. Filling appeared to commence in the lower section of the tower.

3.5.3 Effect of Particles

Since the density of the fluidizing particle is low, the solid phase is easily set in motion by the
upward flow of gas phase, the ease of the motion being aided by the downfall of the liquid phase. A state of vigourous contact between the gas and liquid phases is obtained, as they flow through the bed. When the entire bed becomes fluidized with very vigourous particle movement, the pressure drop across the bed reaches a constant with increasing gas velocity. The bed height progressively increases with gas velocity.

During an otherwise normal fluidization, it is observed that the particles start lining up at the wall. Once a single particle layer is set on the grid around the wall, subsequent stacking of the remaining particles as one layer over the other takes place within a minute. This sort of monolayer stacking of the particles is referred to as "congregation at the wall". This type of congregation is seen to form and break off on its own. Whenever the congregation takes place, it is broken by sudden increase in liquid flow rate.

It has been observed that where the liquid stream is broken up into small bubbles, an increase in the gas flow rate brings about a further expansion of a fluidized bed, suggesting that fluidization is assisted by the flow of gas. The quality of bubbling fluidization is strongly influenced by the type of gas distributor.

3.6 CALCULATION

The performance characteristics, phase hold ups, tower characteristics, mass transfer, heat transfer and the non-dimensional numbers given in the succeeding paragraphs are evaluated using suitable computer programme.
### 3.6.1 Properties of Air

The density of air and the viscosity of air are calculated using the expressions of Wu & Baeyens [1991] for the various trials:

**Density of air,** \( \rho_a = \frac{1.2 \times 273}{T_{db}} \) \hspace{1cm} (3.1)

**Viscosity of air,** \( \mu_a = 1.46 \times 10^{-6} \frac{T_{db}^{1.504}}{(T_{db} + 120)} \) \hspace{1cm} (3.2)

The thermal conductivity of air, \('k_a'\), is calculated using the relation:

\( k_a = \begin{cases} 
25.93 + (T_{db} - 293)0.082 & \text{if } T_{db} \leq 303 \text{ K} \\
26.75 + (T_{db} - 303)0.081 & \text{if } T_{db} > 303 \text{ K} 
\end{cases} \times 10^{-6} \) \hspace{1cm} (3.3)

\( k_a = \begin{cases} 
26.75 + (T_{db} - 303)0.081 & \text{if } T_{db} > 303 \text{ K} 
\end{cases} \times 10^{-6} \) \hspace{1cm} (3.4)

where,

\( T_{db} \) is the dry-bulb temperature of air.

### 3.6.2 Humidity Ratio

When the barometric pressure and the Dew Point (DP) temperature are known, the humidity ratio is readily determined from the following relationship, which is derived from the characteristics gas equation.

**Humidity ratio,** \( w = 0.622 \frac{p_w}{(p-p_w)} \)

where,
The partial pressure of water vapour corresponding to dew point temperature in pascals.

p is barometric pressure in pascals.

### 3.6.3 Phase Hold Up

Evaluation of the volume fraction of the gas, liquid, and particle phases in the three-phase fluidized bed cooling tower is performed by determining one volume fraction and obtaining the other two by solving the following equations simultaneously:

\[
\varepsilon_g + \varepsilon_l + \varepsilon_p = 1 \quad (3.5)
\]

\[
\frac{-dP}{dz} = (f_g\varepsilon_g + f_l\varepsilon_l + f_p\varepsilon_p) + (f_g\varepsilon_g U_g du_g/\varepsilon_l U_l du_l/\varepsilon_p U_p du_p/\varepsilon_p) - F_w 
\]

The above equations are obtained based on the one-dimensional steady-state mass and momentum balances. The second term on the right-hand side of the equation (3.6) is the average convective flux of momentum arising from the axial non-uniformity of phase hold ups. As in a fluidized bed, there is no net flow of solids through the system, the average momentum flux of solids, \( \varepsilon_p U_p (du_p/\varepsilon_p) \), at any axial position is zero. \( F_w \) accounts for the frictional loss due to fluid-wall interactions. Since there is no wall effect, the term \( F_w \) is neglected and the pressure drop may be written as:

\[
\frac{-dP}{dz} = (f_g\varepsilon_g + f_l\varepsilon_l + f_p\varepsilon_p)g 
\]

(3.7)
The particle fraction is calculated from the height occupied by the bed of a given weight and density of the particles.

3.6.4 Diffusivity

Gilliland, E.R., [1934], conducted experiments on air-water system to find out the diffusivity, 'D'. The diffusivity is calculated using the relation:

\[ D = 0.258 + 0.047(T_b - 298.9) - 331000 \] (3.8)

3.6.5 Tower Characteristics

The tower characteristics, 'KaV/L', is the parameter widely used for analysing the performance of the cooling tower. The non-dimensional parameter, 'KaV/L', includes the overall mass transfer coefficient, 'K', the interfacial area available for transport, 'a', the height of packing, 'V', and the liquid flow rate, 'L'. The performance of cooling tower usually referred to as tower characteristics, 'KaV/L', depends on the 'L/G' ratio and the temperature levels of the water cooling and the ambient air wet-bulb temperature. The tower characteristics, 'KaV/L', is given by:

\[ KaV/L = \frac{T_i}{\int_{T_o}^{T_i} dT/(h_s-h_a)} \] (3.9)

The above expression is solved by using a more accurate but still fairly simple approximation of the Tchebycheff Four-Point Quadrature formula inwhich \((h_s-h_a)\) is taken as the mean of its values at 0.1, 0.4, 0.6 and 0.9 of T between 'T_i' and 'T_o'.
The above expression can be rewritten after applying the solution procedure:

\[ \frac{K_a V}{L} = (T_1 - T_0) \left[ \frac{1}{h_1} + \frac{1}{h_2} + \frac{1}{h_3} + \frac{1}{h_4} \right] + 4 \]  \hspace{1cm} (3.10)

where,

- \( h_s \) = the specific enthalpy of air at the surface,
- \( h_a \) = the specific enthalpy of bulk of air locally,
- \( h_1 \) = value of \((h_s - h_a)\) at \((T_0 + 0.1(T_1 - T_0))\),
- \( h_2 \) = value of \((h_s - h_a)\) at \((T_0 + 0.4(T_1 - T_0))\),
- \( h_3 \) = value of \((h_s - h_a)\) at \((T_1 - 0.4(T_1 - T_0))\), and
- \( h_4 \) = value of \((h_s - h_a)\) at \((T_1 - 0.1(T_1 - T_0))\).

The tower characteristics, 'KaV/L', is also found out from the cooling tower characteristics charts, published by Cooling Tower Institute for the range & approach of the experimental data.

### 3.6.6 Transfer Unit

From the tower characteristics, 'KaV/L', the height of transfer unit, 'HTU' and number of transfer unit, 'NTU' are calculated using the following relations:

\[ HTU = \frac{V}{(KaV/L)} \div \frac{L}{G} \] \hspace{1cm} (3.11)

\[ NTU = (KaV/L) \times \frac{L}{G} \] \hspace{1cm} (3.12)

### 3.6.7 Mass Transfer

The mass transfer coefficient, 'K', has been characterized as a function of gas/air & liquid/water
velocities and particle size. Because of the impossibilities of measuring moisture concentrations at the air-water interface, the resulting rates of mass transfer can be expressed only as overall coefficient, rather than as coefficients for the individual fluids. The mass transfer coefficient mainly depends on the particle size since the particle size is known to differentiae the bubble coalescing and disintegrating flow regimes in three-phase fluidized beds.

Knowing height of transfer unit, 'HTU', from equation (3.11), volumetric mass transfer coefficient, 'Ka', is calculated from:

\[ Ka = \frac{G}{HTU} \]  

(3.13)

Interfacial area is the prime factor affecting heat and mass transfer. Interfacial area, 'a', is calculated using the correlation proposed by Dhanuka & Stepanek [1980]. Darton [1986] reported an expression for diameter of bubble, 'db'.

\[ a = \frac{6 \epsilon g}{d_b} \]  

(3.14)

\[ d_b = \left(0.72 \left(\frac{\mu_b}{\rho_b}\right)^{0.22} \epsilon_g^{0.33}\right) 
\quad + \left(1+1.6(\epsilon_p^d_p/1000)^{1.2}\right) \]  

(3.15)

where,

\[ \mu_b = \mu_1 \exp(36.15 \epsilon_p^{2.5}) \]

\[ \rho_b = \rho_g \epsilon_g + \rho_1 \epsilon_1 + \rho_p \epsilon_p \]

Using the above, the interfacial area, 'a', and mass transfer coefficient, 'K', are calculated.
3.6.8 Tower Efficiency and Make-up Requirement

The tower efficiency, $\eta'$, is the ratio of cooling achieved to the cooling potential available. One of the important aspects of a cooling tower is the requirement of make-up water. Make-up water requirement, $\text{m}_\text{up}$, is the ratio of the humidity ratio difference between the inlet and outlet air to Liquid-Gas mass flow rate (L/G). The formulae are:

$$\eta = \left( \frac{T_R}{T_R + T_A} \right) \times 100 \quad (3.16)$$

$$\text{m}_\text{up} = \left( \frac{w_O - w_i}{L/G} \right) \times 100 \quad (3.17)$$

Using the above formulae, the tower efficiency and the make-up water requirement are calculated.

3.6.9 Heat Transfer

The heat flux, '$q$', is calculated using the relation:

$$q = (T_{cpl,i} - T_{cpl,o}) \frac{L}{3600} \quad (3.18)$$

Kito et al. [1981] stated that the heat transfer coefficient, '$h$', for a three-phase fluidized bed system can be calculated using the equation:

$$h = \frac{q}{(T_i - T_b)} \quad (3.19)$$

Using the above equation (3.19), heat transfer coefficient, '$h$', is calculated.
3.6.10 Non-Dimensional Numbers

The following non-dimensional numbers are calculated using the expressions given below:

Modified Reynolds number, \( \text{Re}_m = \frac{U_g d_p \rho \epsilon \mu_g}{6 \mu_g \epsilon_1} \) \hspace{1cm} (3.20)

Modified Nusselt number, \( \text{Nu}_m = \frac{h d_p \epsilon}{6 \epsilon_g \kappa_a} \) \hspace{1cm} (3.21)

Prandtl number, \( \text{Pr} = \frac{c_p \rho \mu_g}{\kappa_a} \) \hspace{1cm} (3.22)

Modified Peclet number, \( \text{Pe}_m = \text{Re}_m \text{Pr} \) \hspace{1cm} (3.23)

Sherwood number, \( \text{Sh} = \frac{K d_p}{3600 D_p} \) \hspace{1cm} (3.24)