Mass Transfer in a Turbulent Contact Absorber

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Dedicated to

*My loving late mother*

*who departed during my Ph.D. studies*

*and whose prayers succeeded me in life*
Declaration

I declare that all material in this thesis which is not my own work has been identified and that no material has previously been submitted and approved for the award of a degree by this or any other university.

Signature: ______________________
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It is certified that the work in this thesis is carried out and completed under my supervision.

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Abstract

Vast amount of discrepancies are found in literature reporting the hydrodynamic and mass transfer characteristics of the turbulent contact absorbers. Most of the literature available in this regard is for small diameter columns having wall effects. Therefore hydrodynamic and mass transfer characteristics of turbulent contact absorbers have been studied in a relatively large scale, 44.7 cm diameter Perspex column so that there are no wall effects. The data reported can now be used in the designing of these absorbers. Efforts have also been made to point out the reasons of discrepancies in the reported literature.

Dimensionless correlations have been developed from the data obtained in this study based on the criteria that the ratio of the column diameter to packing diameter should be greater than 10, the column diameter should be greater than 15 cm, grid free area should be greater than 70%, the ratio of static bed height to column diameter should be less than 1 and the contributions of the gas and liquid distributors should be accounted for.

The correlations developed were used to simulate the data of the earlier workers, which shows that the data of those workers who have fulfilled the above criteria, their data can also be represented with in ±30% error band, using the correlations developed in this study.

To develop these correlations, necessary detailed pressure drop, liquid holdup, expanded bed height and the volumetric gas film mass transfer coefficients data was obtained in this study. To do this the variation of packing diameter (25mm, 38mm and 45mm), apparent packing density (160, 270, 354, 442, 547 kg/m³) and static bed height (15cm, 25cm and 35 cm) were studied. Gas and liquid velocities in the range of 1.8 to 3.6 m/s and 0 to 0.012 m/s were used respectively. For the hydrodynamic study water and air were used as working fluids. For the mass transfer coefficients the absorber was used in an adiabatic humidification mode.

Violations of one or more of the above mentioned criteria are the main reasons for the vast amount of discrepancies in the literature data. These discrepancies fall within the experimental errors, only in the reported literature which has explicitly taken into consideration of the effect of gas and liquid distributors.

It is also shown that all the reported data can be represented in a dimensionless form.
Table of Contents

List of Figures ............................................................................................................... ix
List of Tables .............................................................................................................. xiv
List of Publications ................................................................................................... xx
Chapter 1 ........................................................................................................................ 1
  1.1 Introduction ............................................................................................................ 1
  1.2 Historical background .......................................................................................... 2
  1.3 Hydrodynamics of TCA ...................................................................................... 3
  1.4 Mass Transfer in TCA ....................................................................................... 3
  1.5 Operating regimes of TCA ................................................................................. 3
  1.6 Objectives of the present study .......................................................................... 5
  1.7 Outline of the thesis ........................................................................................... 5

Chapter 2 ...................................................................................................................... 7
  2.1 Review of Hydrodynamics in TCA ..................................................................... 7
  2.2 Pressure Drop ..................................................................................................... 7
  2.3 Liquid Holdup ....................................................................................................... 13
  2.4 Expanded Bed Height ....................................................................................... 16

Chapter 3 ...................................................................................................................... 19
  3.1 Review of mass transfer in TCA ......................................................................... 19
  3.2 Interfacial area, volumetric liquid film and overall mass transfer coefficients ..... 19
  3.3 Volumetric gas film transfer coefficients ........................................................... 27

Chapter 4 ...................................................................................................................... 35
  4.1 Experimental Setup and Procedures ................................................................... 35
  4.2 Pressure drop measurements .............................................................................. 36
  4.3 Liquid holdup measurements ............................................................................. 37
  4.4 Volumetric gas film transfer coefficient .............................................................. 38

Chapter 5 ...................................................................................................................... 42
  5.1 Mathematical Modeling ....................................................................................... 42
  5.2 Humidification and dehumidification in TCA ..................................................... 43
  5.3 Mathematical modeling for adiabatic humidification ......................................... 47

Chapter 6 ...................................................................................................................... 50
  6.1 Hydrodynamics Results and Discussion ............................................................. 50
  6.2 Pressure Drop ..................................................................................................... 50
  6.3 Expanded Bed Height ....................................................................................... 60
  6.4 Liquid holdup ...................................................................................................... 77

Chapter 7 ...................................................................................................................... 96
  7.1 Mass Transfer Results and Discussions ............................................................. 96
  7.2 Presentation of results ....................................................................................... 96
  7.3 Mass transfer in the plenum .............................................................................. 98
  7.4 Volumetric Mass transfer coefficients in the TCA bed based on the static bed volume... 99
<table>
<thead>
<tr>
<th>Chapter</th>
<th>Section</th>
<th>Title</th>
<th>Page</th>
</tr>
</thead>
<tbody>
<tr>
<td>7.5</td>
<td>Operational efficiencies of Bed, $\eta_{op,b}$</td>
<td>104</td>
<td></td>
</tr>
<tr>
<td>7.6</td>
<td>Comparison with literature</td>
<td>107</td>
<td></td>
</tr>
<tr>
<td>7.7</td>
<td>Graphs for mass transfer in TCA</td>
<td>109</td>
<td></td>
</tr>
<tr>
<td>8</td>
<td>Conclusions and Future Recommendations</td>
<td>127</td>
<td></td>
</tr>
<tr>
<td>8.1</td>
<td>Conclusion</td>
<td>127</td>
<td></td>
</tr>
<tr>
<td>8.2</td>
<td>Future Recommendations</td>
<td>128</td>
<td></td>
</tr>
<tr>
<td></td>
<td>References</td>
<td>129</td>
<td></td>
</tr>
<tr>
<td>9</td>
<td>Figures of mass transfer in the whole column including plenum</td>
<td>134</td>
<td></td>
</tr>
<tr>
<td>10</td>
<td>Sample calculations for mass transfer experiments</td>
<td>146</td>
<td></td>
</tr>
</tbody>
</table>
List of Figures

Figure 1-1 Type 1 and type 2 fluidization operation from Vunjak-Novakovic et al. (1987a) ........................................................................................................................................4

Figure 2-1 Comparison of pressure drop calculated from literature correlations for Type 1 fluidization ........................................................................................................................................12

Figure 2-2 Comparison of pressure drop calculated from literature correlations for Type 2 fluidization ........................................................................................................................................12

Figure 2-3 Comparison of liquid holdup calculated from literature correlations for Type 1 fluidization ........................................................................................................................................15

Figure 2-4 Comparison of liquid holdup calculated from literature correlations for Type 2 fluidization ........................................................................................................................................15

Figure 2-5 Comparison of expanded bed height correlations ........................................................................................................................................18

Figure 4-1: Experimental setup ..........................................................................................................................39

Figure 4-2: Gas distributor ....................................................................................................................................39

Figure 4-3: Mist eliminator ....................................................................................................................................40

Figure 4-4: Pressure drop measurement arrangement ..........................................................................................40

Figure 4-5: Photo of experimental set up ...........................................................................................................41

Figure 4-6: Photo of experimental setup with insulation .....................................................................................41

Figure 5-1: Liquid and gas films for mass transfer ............................................................................................42

Figure 5-2: Humidification of air in TCA ...........................................................................................................43

Figure 5-3: Adiabatic humidification of air in TCA ............................................................................................47

Figure 6-1 Effect of gas and liquid velocities on pressure drop for type 1 ..........................................................63

Figure 6-2 Effect of gas and liquid velocities on pressure drop for type 2 ..........................................................63

Figure 6-3: Effect of liquid and gas velocities on pressure drop for 25 mm dia packing ........................................64

Figure 6-4: Effect of liquid and gas velocities on pressure drop for 45 mm dia packing ........................................64

Figure 6-5 Effect of density of packing on pressure drop ....................................................................................65

Figure 6-6 Effect of diameter of packing on pressure drop ..................................................................................65

Figure 6-7 Effect of height of packing on pressure drop for type 1 TCA operation ..............................................66

Figure 6-8 Effect of height of packing on pressure drop for type 2 TCA operation ..............................................66
Figure 6-9: Effect of pressure drop per unit static bed height of packing on pressure drop .............................................................................................................................. 67
Figure 6-10 Effect of pressure drop across grid.............................................................................................................................. 67
Figure 6-11 Comparison of pressure drop by direct measurement with that calculated from liquid holdup ................................................................................................................. 68
Figure 6-12 Comparison with literature correlations for type 1 TCA operation at constant gas velocity .............................................................................................................................. 68
Figure 6-13 Comparison with literature correlations for type 1 TCA operation at constant liquid velocity .............................................................................................................................. 69
Figure 6-14 Comparison with literature correlations for type 2 TCA operation at constant gas velocity .............................................................................................................................. 69
Figure 6-15 Comparison with literature correlations for type 2 TCA operation at constant liquid velocity .............................................................................................................................. 70
Figure 6-16 Pressure drop from literature vs. present study correlation for type 1 fluidization .............................................................................................................................. 70
Figure 6-17 Pressure drop from literature vs. present study correlation for type 2 fluidization .............................................................................................................................. 71
Figure 6-18 Pressure drop across plenum without incorporating velocity head in air inlet pipe .............................................................................................................................. 72
Figure 6-19 Pressure drop across plenum by adding velocity head in air inlet pipe into measured pressure drop .............................................................................................................................. 72
Figure 6-20 Effect of liquid velocities on expanded bed height for type 1 TCA operation .............................................................................................................................. 73
Figure 6-21 Effect of liquid velocities on bed expansion for type 2 TCA operation .............................................................................................................................. 73
Figure 6-22 Effect of gas velocities on expanded bed height for type 1 TCA operation .............................................................................................................................. 74
Figure 6-23 Effect of gas velocities on bed expansion for type 2 TCA operation .............................................................................................................................. 74
Figure 6-24 Effect of diameter of packing on bed expansion .............................................................................................................................. 75
Figure 6-25 Effect of density of packing on bed expansion .............................................................................................................................. 75
Figure 6-26 Effect of gas velocity on bed expansion for different static bed heights for type 1 .............................................................................................................................. 76
Figure 6-27 Effect of gas velocity on bed expansion for different static bed heights for type 2 .............................................................................................................................. 76
Figure 6-28 Comparison of present correlation with literature correlations .......... 77
Figure 6-29: Effect of liquid and gas velocities on liquid holdup for type 1 fluidization ........................................................................................................................................ 85
Figure 6-30: Effect of liquid and gas velocities on liquid holdup for type 2 ............. 85
Figure 6-31 Effect of liquid and gas velocities on liquid holdup with plenum for type 1 ................................................................................................................................................. 86
Figure 6-32 Effect of liquid and gas velocities on liquid holdup with plenum for type 2 ................................................................................................................................................. 86
Figure 6-33 Effect of density of packing on liquid holdup ........................................ 87
Figure 6-34 Effect of diameter of packing on liquid holdup ........................................ 87
Figure 6-35 Effect of static bed height on liquid holdup for type 1 ......................... 88
Figure 6-36 Effect of static bed height on liquid holdup for type 2 ............................. 88
Figure 6-37 Effect of static bed height on liquid holdup for type 1 with plenum ...... 89
Figure 6-38 Effect of static bed height on liquid holdup with plenum for type 2 ....... 89
Figure 6-39: Liquid holdup in empty column ............................................................. 90
Figure 6-40 Comparison of effect of liquid velocity on liquid holdup for type 1 ...... 90
Figure 6-41 Comparison of effect of liquid velocity on liquid holdup for type 2 ...... 91
Figure 6-42 Comparison of effect of gas velocity on liquid holdup for type 1 ............ 91
Figure 6-43 Comparison of effect of gas velocity on liquid holdup for type 2 ............ 92
Figure 6-44 Comparison of effect of diameter of packing on liquid holdup type 2 .... 92
Figure 6-45 Comparison of effect of static bed height on liquid holdup for type 1 .... 93
Figure 6-46 Comparison of effect of static bed height on liquid holdup for type 2 .... 93
Figure 6-47: Liquid holdup from literature vs. present study correlation for type 1 fluidization ................................................................................................................................. 94
Figure 6-48: Liquid holdup from literature vs. present study correlation for type 2 fluidization ................................................................................................................................. 94
Figure 7-1: Effect of liquid and gas velocities on \( (k_ga)_p \) in plenum ...................... 109
Figure 7-2: Effect of liquid and gas velocities on \( (k_ga)_{b,s} \) for density 180 kg/m$^3$ 109
Figure 7-3: Effect of liquid and gas velocities on \( (k_ga)_{b,s} \) for density 270 kg/m$^3$ 109
Figure 7-4: Effect of liquid and gas velocities on \( (k_ga)_{b,s} \) for density 354 kg/m$^3$, dia 38 mm ............................................................................................................................................. 110
Figure 7-5: Effect of liquid and gas velocities on \( (k_ga)_{b,s} \) for density 354 kg/m$^3$, dia 25 mm ............................................................................................................................................. 111
Figure 7-6: Effect of packing size on \((k_ga)_{b,s}\) for density 354 kg/m\(^3\) at different liquid velocities ................................................................. 111
Figure 7-7: Effect of packing size on \((k_ga)_{b,s}\) for density 354 kg/m\(^3\) at different gas velocities .................................................................................................................... 112
Figure 7-8: Effect of static bed height on \((k_ga)_{b,s}\) at different gas velocities for density 180 kg/m\(^3\) ................................................................................................................... 112
Figure 7-9: Effect of static bed height on \((k_ga)_{b,s}\) at different liquid velocities for density 180 kg/m\(^3\) ................................................................. 113
Figure 7-10: Effect of static bed height on \((k_ga)_{b,s}\) for density 354 kg/m\(^3\) for different gas velocities .............................................................................................................. 113
Figure 7-11: Effect of static bed height on \((k_ga)_{b,s}\) for density 354 kg/m\(^3\) at different liquid velocities ................................................................. 114
Figure 7-12: Effect of packing density on \((k_ga)_{b,s}\) at different liquid velocities .... 114
Figure 7-13: Effect of packing density on \((k_ga)_{b,s}\) at different gas velocities .... 115
Figure 7-14: Effect of liquid and gas velocities on \(\eta_{op,p}\) for plenum ............ 115
Figure 7-15: Effect of liquid and gas velocities on \(\eta_{op,b}\) for density 180 kg/m\(^3\) ...... 116
Figure 7-16: Effect of liquid and gas velocities on \(\eta_{op,b}\) for density 270 kg/m\(^3\) ...... 116
Figure 7-17: Effect of liquid and gas velocities on \(\eta_{op,b}\) for density 354 kg/m\(^3\), dia 38 mm ................................................................. 117
Figure 7-18: Effect of liquid and gas velocities on \(\eta_{op,b}\) for density 354 kg/m\(^3\), dia 25 mm ................................................................................................................................. 117
Figure 7-19: Effect of packing size on \(\eta_{op,b}\) for density 354 kg/m\(^3\) at different gas velocities .................................................................................................................... 118
Figure 7-20: Effect of packing size on \(\eta_{op,b}\) for density 354 kg/m\(^3\) at different liquid velocities .................................................................................................................... 118
Figure 7-21: Effect of static bed height on \(\eta_{op,b}\) at different gas velocities for density 180 kg/m\(^3\) ................................................................................................................... 119
Figure 7-22: Effect of static bed height on \(\eta_{op,b}\) at different liquid velocities for density 180 kg/m\(^3\) ................................................................................................................... 119
Figure 7-23: Effect of static bed height on \(\eta_{op,b}\) for density 354 kg/m\(^3\) for different gas velocities .................................................................................................................... 120
Figure 7-24: Effect of static bed height on \(\eta_{op,b}\) for density 354 kg/m\(^3\) at different liquid velocities .................................................................................................................... 120
Figure 7-25: Effect of packing density on \( \eta_{op,b} \) at different gas velocities .......... 121
Figure 7-26: Effect of packing density on \( \eta_{op,b} \) at different liquid velocities .......... 121
Figure 7-27: Literature comparison with effect of gas velocity on \( k_{ga} \) ...................... 123
Figure 7-28: Literature comparison with effect of liquid velocity on \( k_{ga} \) .................. 124
Figure 7-29: \((kga)_{b,s}\) from literature vs. present study correlation for type1 fluidization ........................................................................................................................................... 125
Figure 7-30: \((kga)_{b,s}\) from literature vs. present study correlation for type2 fluidization ........................................................................................................................................... 125
Figure 9-1: Effect of liquid and gas velocities on \((kga)_{c,s}\) for density 180 kg/m\(^3\) ...... 134
Figure 9-2: Effect of liquid and gas velocities on \((kga)_{c,s}\) for density 270 kg/m\(^3\) ...... 134
Figure 9-3: Effect of liquid and gas velocities on \((kga)_{c,s}\) for density 354 kg/m\(^3\), dia 38 mm ........................................................................................................................................... 135
Figure 9-4: Effect of liquid and gas velocities on \((kga)_{c,s}\) for density 354, kg/m\(^3\), dia 25 mm ........................................................................................................................................... 135
Figure 9-5: Effect of packing size on \((kga)_{c,s}\) for density 354 kg/m\(^3\) at different gas velocities ........................................................................................................................................... 136
Figure 9-6: Effect of packing size on \((kga)_{c,s}\) for density 354 kg/m\(^3\) at different liquid velocities ........................................................................................................................................... 136
Figure 9-7: Effect of static bed height on \((kga)_{c,s}\) at different gas velocities for density 180 kg/m\(^3\) ........................................................................................................................................... 137
Figure 9-8: Effect of static bed height on \((kga)_{c,s}\) at different liquid velocities for density 180 kg/m\(^3\) ........................................................................................................................................... 137
Figure 9-9: Effect of static bed height on \((kga)_{c,s}\) for density 354 kg/m\(^3\) for different gas velocities ........................................................................................................................................... 138
Figure 9-10: Effect of static bed height on \((kga)_{c,s}\) for density 354 kg/m\(^3\) at different liquid velocities ........................................................................................................................................... 138
Figure 9-11: Effect of packing density on \((kga)_{c,s}\) at different gas velocities .......... 139
Figure 9-12: Effect of packing density on \((kga)_{c,s}\) at different liquid velocities .......... 139
Figure 9-13: Effect of liquid and gas velocities on \( \eta_{op,c} \) for density 180 kg/m\(^3\) ........... 140
Figure 9-14: Effect of liquid and gas velocities on \( \eta_{op,c} \) for density 270 kg/m\(^3\) ........... 140
Figure 9-15: Effect of liquid and gas velocities on \( \eta_{op,c} \) for density 354 kg/m\(^3\), dia 38 mm ........................................................................................................................................... 141
Figure 9-16: Effect of liquid and gas velocities on $\eta_{op,c}$ for density 354 kg/m$^3$, dia 25 mm .......................................................... 141
Figure 9-17: Effect of packing size on $\eta_{op,c}$ for density 354 kg/m$^3$ at different gas velocities ..................................................... 142
Figure 9-18: Effect of packing size on $\eta_{op,c}$ for density 354 kg/m$^3$ at different liquid velocities ..................................................... 142
Figure 9-19: Effect of static bed height on $\eta_{op,c}$ at different gas velocities for density 180 kg/m$^3$ .................................................. 143
Figure 9-20: Effect of static bed height on $\eta_{op,c}$ at different liquid velocities for density 180 kg/m$^3$ .................................................. 143
Figure 9-21: Effect of static bed height on $\eta_{op,c}$ for density 354 kg/m$^3$ for different gas velocities ..................................................... 144
Figure 9-22: Effect of static bed height on $\eta_{op,c}$ for density 354 kg/m$^3$ at different liquid velocities ..................................................... 144
Figure 9-23: Effect of packing density on $\eta_{op,c}$ at different gas velocities ............. 145
Figure 9-24: Effect of packing density on $\eta_{op,c}$ at different liquid velocities ............. 145

List of Tables

Table 2-1: Variables used in literature for hydrodynamic studies ......................... 9
Table 2-2: Pressure drop correlations in literature ................................................. 11
Table 2-3: Liquid holdup correlations in literature ................................................. 14
Table 2-4: Bed expansion correlations available in literature ............................... 17
Table 3-1: Comparison between TCA and packed columns ............................... 26
Table 3-2: Variables used in literature for mass transfer studies .......................... 33
Table 4-1: Range of Variables studied for liquid holdup ..................................... 37
Table 4-2: Specifications of the packings used ...................................................... 38
Table 6-1: Parameters used in graphs for the comparison of pressure drop with present study correlation .............................................. 71
Table 6-2: Parameters used in graphs for the comparison of liquid holdup with present study correlation .............................................. 95
Table 7-1: Equations used in volumetric gas film transfer coefficient and operational efficiency .............................................................. 97
Table 7-2: No. of balls per unit height of expanded bed height at various liquid and gas velocities

Table 7-3: \((k_g a)_{b, s}\) and mass transferred at different static bed height

Table 7-4: Mass transfer contribution of 15 cm deep bed

Table 7-5: Comparison of \(k_g a\) with literature

Table 7-6: Conditions for readings of Figure 7-27

Table 7-7: Conditions for readings of Figure 7-28

Table 7-8: Parameters used in graphs for the comparison of mass transfer coefficient with present study correlation
Nomenclature

\( A_c \) = Area of column, m\(^2\)
\( a, a \) = Interfacial area, m\(^2\)/m\(^3\)
\( a_s, a_s \) = Interfacial area based on static bed volume
\( a_e, a_e \) = Interfacial area based on expanded bed volume, m\(^2\)/m\(^3\)
\( a_c, a_c \) = Interfacial area based on column cross section, m\(^2\)/m\(^2\)
\( c_s \) = Humid heat of the air, J/(kg-K)
\( c_i \) = Specific heat of liquid, J/(kg-K)
\( C = \text{Concentration of dissolved gas A, kmoles/m}^3 \)
\( C_A^* = \text{Concentration of dissolved gas A at the surface, kmoles/m}^3 \)
\( C_0 = \text{Concentration of dissolved gas in the bulk of the liquid, kmoles/m}^3 \)
\( C_{B_0} = \text{Concentration of Component B in the bulk liquid, kmoles/m}^3 \)
\( d \) = Equivalent diameter of hole of support grid, m
\( D_c \) = Diameter of column, m
\( D = \text{Equivalent diameter for free open area of the support grid (} D_c f^{0.5} \text{), m} \)
\( D_A = \text{Diffusivity of component A in liquid, m}^2/s \)
\( d_p \) = Diameter of packing, m
\( d_o \) = Diameter of hole in perforated packing, m
\( f \) = Fractional free area of support grid
\( G' = \text{Superficial mass velocity of air, kg/(m}^2\text{-s)} \)
\( G_M = \text{Superficial molar mass velocity of air, kmoles/(m}^2\text{-s)} \)
\( Ga_t = \text{Grashof number,} \frac{d_p^3 g \rho_p^2}{\mu_i^2} \)
\( Fr_i = \text{Froud number,} \frac{u_i}{\sqrt{gd_p}} \)
\( h_l \) = Liquid held, m
\( H \) = Height of packing, m
$H_0 =$ Static bed height, m

$H_e =$ Expanded bed height, m

$H'$ = Distribution coefficient (Also known as Henry coefficient) the ratio of O$_2$ concentration in gas to sulfite solutions (kmoles/m$^3$)/(kmoles/m$^3$).  

$H_e =$ Henry’s constant, Pa/(kmoles/m$^3$)

$\mathcal{H} =$ Humidity of the air, kg/kg

$\mathcal{H}_s =$ Humidity of saturated air, kg/kg

$h =$ Heat transfer coefficient between gas and liquid surface, W/(m$^2$-K)

$h_l a_c =$ Liquid side heat transfer coefficient, W/(m$^3$-K)

$h_0 =$Determining linear dimension, m

$i_o =$ Enthalpy of unit mass of dry air along with vapors it contains, J/kg

$i_{o,sat} =$ enthalpy of saturated air, kJ/kg

$k_i =$ Rate constant for first order reaction, s$^{-1}$

$k' =$ Mass transfer coefficient kg/(m$^2$-s)

$k'a_e =$ Volumetric gas film mass transfer coefficient, kg/(m$^3$-s)

$K'a_e =$ Volumetric overall mass transfer coefficient, kg/(m$^3$-s)

$k_l =$ Liquid film transfer coefficient, m/s

$k_{la} =$ Volumetric liquid film mass transfer coefficient, s$^{-1}$

$k_{ga} =$ Volumetric gas film mass transfer coefficient, kmoles/(m$^3$-s-atm)

$L =$ Liquid superficial mass velocity, kg/(m$^2$-s)

$m_s =$ Number of liquid transfer units.

$M_o =$ Molecular weight of air, kg/kmole

$N_o =$ No. of gas phase transfer units

$P_o =$ Air pressure, Pa

$\Delta p =$ Pressure drop across bed, Pa

$\Delta p_f =$ The sum of all pressure drops except due to weight of liquid and packing, Pa

$\Delta p_0 =$ Pressure drop across grid, Pa

$R =$ Rate of absorption per unit interfacial area, kmoles/(m$^2$-s)
Re_\text{e} = \text{Reynolds number}, \quad \frac{d_p u_t \rho_t}{\mu_t}

\hat{Re}_e = \text{Reynolds number}, \quad \frac{D u_t \rho_t}{\mu_t}

\hat{Re}_g = \text{Reynolds number}, \quad \frac{D u_g \rho_g}{\mu_g}

Re_L = \text{Reynolds number used for mass transfer correlation}, \quad \frac{h_{fg} u_t \rho_t}{\mu_t}

r = \text{Rate of reaction of component A per unit volume, kmoles/m}^3\cdot\text{s} \quad T = \text{Temperature C}^0

T_0 = \text{Reference temperature, } 0\text{C}^0

T_{GI} = \text{Air inlet temperature, } C^0

T_{GO} = \text{Air outlet temperature, } C^0

T_s = \text{Adiabatic saturation temperature, } C^0

u_g = \text{Superficial gas velocity in column, m/s}

u_t = \text{Superficial liquid velocity in column, m/s}

u_{mf} = \text{Minimum fluidization velocity, m/s}

V = \text{Volume of liquid held in column, m}^3

We = \text{Weber number}, \quad \frac{d_p u_t^2 \rho_t}{\sigma_t}

x = \text{Distance from surface of the film, m}

y = \text{Mole fraction of water vapor in air}

\textbf{Greek letters}

\epsilon_0 = \text{Void fraction in the bed.}

\epsilon_{l,\text{st}} = \text{Liquid holdup in the bed based on static bed volume, m}^3/\text{m}^3

\epsilon_{d,\text{st}} = \text{Dynamic liquid holdup based on static bed height, m}^3/\text{m}^3

\epsilon_{c,\text{st}} = \text{Liquid holdup in the column based on static bed volume, m}^3/\text{m}^3

\epsilon_{g,\text{st}} = \text{Gas fraction in the bed.}

\epsilon_{p,\text{st}} = \text{Packing fraction in the bed.}

\epsilon_l = \text{Liquid holdup determined by shut off valve technique
\( \varepsilon^b \) = Liquid holdup determined by tracer technique
\( \varepsilon^c \) = Liquid holdup determined by pressured drop measurements
\( \rho_g \) = Density of the gas, kg/m\(^3\)
\( \rho_p \) = Apparent density of packing, kg/m\(^3\)
\( \rho_l \) = Density of liquid, kg/m\(^3\)
\( \mu_l \) = Viscosity of liquid, kg/(m-s)
\( \sigma_l \) = Surface tension of liquid, N/m
\( \nu_g \) = Kinematic viscosity of the gas, m\(^2\)/s
\( \eta_{OP} \) = Operational efficiency (Kmoles/N-m)
\( \lambda \) = Latent heat of vaporization kJ/kg
\( \lambda_0 \) = Latent heat of vaporization at temperature \( T_0 \)
\( \phi \) = Fractional free area in perforated packing
\( \delta \) = Film thickness, m
\( \xi \) = Sphericity

**Subscripts**
1 = Inlet
2 = Outlet
c = Column
e = Expanded bed
\( G, g \) = Gas, (i.e. air)
i = Interface
\( L, l \) = Liquid
p = Plenum
s = Static bed
List of Publications


3- Ijtaba Husain, M. Zaman, Mansoor H.Inayat, Azharul Haq” Mass Transfer in a Trayed Bubble Column Reactor for Air Sodium Sulfite System”Proceedings of 3rd Symposium on Engineering Sciences Lahore, Pakistan March 10, 2010
Chapter 1

1 Introduction

1.1 Introduction

In order to achieve high mass transfer between a liquid and a gaseous phase, one needs as high contact as possible between the two phases. Humans have endeavored to construct such devices which could generate as high interfacial area as possible. Various ideas have been used successfully in the chemical industry such as tray type devices (i.e. sieve trays, bubble cap trays or valve trays etc), packed towers (with various types of structured and random packing), bubble columns, venturi type devices and fluidized beds. Fluidized beds offer very high interfacial area per unit volume of the active tower. One such device which uses relatively large size solid inert packing as the fluidizing media is turbulent contact absorber. In industry, turbulent contact absorber (TCA) has been described as turbulent contacter, floating bed scrubber, fluidized bed scrubber, turbulent bed cooling tower and fluidized packing contactor. It is a mass transfer equipment in which light packing is fluidized by counter current flow of gas (flowing upwards) and liquid (flowing downwards). Gas flows as continuous and liquid as dispersed phase in TCA. Mostly hollow spherical shaped packings are used in TCA. However, irregular shaped packing and packing with perforations have also been used in TCA. Vigorous motion of the packing in the bed causes turbulence and hence results in enhanced mass transfer as compared to the packed beds. Other advantages are non-choking, self-cleaning due to tumbling motion of packing and low capital cost (due to high gas and liquid flow rates, as flooding is not a problem for TCA at high throughputs unlike packed and tray columns). Furthermore, TCA does not require fine liquid distribution which results in little pressure drop across liquid distributor. The packing in TCA is easily handled. It can be removed from equipment by air transport means. Manufacturing of TCA packing does not need special expertise. It can easily be made from corrosion and chemical resistant plastics depending upon the chemicals involved. There has been a problem of mass breakage of spheres due to the accelerated number of collisions.
caused by severe tumbling action and by the inherent problem of uneven (thinned) wall thickness unavoidable when using blow moulded 'stretched' spherical balls. However, now-a-days these are mostly made by joining two threaded hemispheres of uniform thickness, thus significantly improving the wall thinning phenomenon. Mass transfer efficiency in TCA can be controlled by changing density, shape and height of the packing.

TCA is generally used in particulate removal, air cooling, humidification, dehumidification, absorption, desorption, flue gas desulfurization and alcohol fermentation. There are also some disadvantages of TCA i.e. back mixing in liquid phase, breakage of packing and bed pulsations particularly in deep and heavy beds.

Important parameters to be considered in the design of TCA are minimum fluidization velocity, liquid holdup, pressure drop, expanded bed height, liquid and gas phase transfer coefficients.

1.2 Historical background

TCA was first introduced by Aluminum Company of Canada (Kielback 1959) to overcome fouling problems in the scrubbers in use by the company. Aluminum was produced by electrolysis of Alumina in cells made of steel shells. Electrolytes in the cells consisted of fluoride salts. During production, hydrogen fluoride and fumes containing fluorides were released. Exhaust from the cells contained particulate matter such as alumina and carbon. A pilot scale TCA was connected to the exhaust of one cell. The TCA removed 99% of total solids and 95% of fluorides. Kielback (1959) mentioned that fourteen large scale TCA units were under construction to scrub more than 950 m³/sec gas. Silicon tetra fluoride and silicon tetra chloride were produced in aluminum fluoride converter and magnesium chlorinator. When these were contacted with water, silica produced was deposited in packed beds. This caused a choking problem, which was successfully removed by Aluminum Company by using TCA’s to handle these gases. Various gases have been treated in TCA’s such as ammonia, chlorine, fluorine, SO₂, hydrogen sulfide, hydrogen fluoride, hydrogen chloride, acetic acid, acetic anhydride and VOC’s.
1.3 Hydrodynamics of TCA

Hydrodynamics of TCA comprises of liquid holdup, pressure drop, expanded bed height and minimum fluidization velocity. Various empirical correlations for pressure drop, liquid hold up, minimum fluidization velocity and expanded bed height are available in literature but most of these are for small diameter columns i.e. up to 15 cm diameter and those correlations are specific to the systems studied. There is a very large variation in the hydrodynamic characteristics reported in the literature. This large variation seems to be the reason why a single correlation is not available to predict the available experimental data. This issue raises two questions, either there is some information missing in the reported data, or there is some fundamental error in the understanding of the phenomenon. A detailed experimental investigation is done during this study in an attempt to sort out the cause of this large variation.

1.4 Mass Transfer in TCA

Height of TCA bed for a desired physical absorption can only be determined if volumetric liquid and gas film transfer coefficients are known. Various chemical systems have been used to measure the interfacial areas, individual film transfer coefficients and volumetric film transfer coefficients e.g. absorption of CO₂ in caustic solution or amines and air oxidation of sodium sulfite etc.

The reported literature has all the same very large variations in the mass transfer characteristics reported. Hence, a detailed experimental investigation is required to find the cause of such large variations.

1.5 Operating regimes of TCA

O'Neil et al (1972) classified the regime of operation of TCA into type 1 and type 2. In type 1, fluidization starts before flooding in the column whereas in type 2, fluidization starts after flooding in the column. Vunjak-Novakovic et al (1987a) developed a chart for the demarcation of type 1 and type 2 modes of operation of TCA as shown in Figure 1-1. Density of packing imparts major contribution in deciding the mode of operation. Normally, packing with density greater than 300 kg/m³ exhibits fluidization which is termed as type 2 fluidization while, packing with
density less than 300 kg/m$^3$ causes type 1 fluidization. Increase in liquid flow rate and decrease in packing diameter also shifts the regime from type 1 into type 2. Therefore unlike packed towers, TCAs’ can be operated at flooding conditions in type 2 fluidization regime. Above classification of type 1 and type 2 regimes of TCA operation is very broad. Bruce et al (2002) have developed flow regime maps for type 1 and type 2 TCA operations in the bed with and without downcomer. They have divided flow regime maps into static bed regime, partially fluidized and partially packed regime, complete fluidization regime, flooding regime and annular regime. All these maps were developed by pressure drop method.

Figure 1-1 Type 1 and type 2 fluidization operation from Vunjak-Novakovic et al.(1987a)
1.6 Objectives of the present study

Over the years many people have postulated theories about fluidized beds for absorption and desorption. While reviewing the literature one is struck by the enormous diversity of equations and correlations available. This is extremely confusing to the practicing chemical engineer and has undoubtedly contributed to the view that fluidized beds were little understood novel devices, often designed using rules of thumb. Hence the primary objective of the present study is to systematically study the hydrodynamics and mass transfer behavior of TCA in pilot scale equipment and try to resolve the enormous amount of diversity of data produced in literature over the years. In order to achieve these objectives, the following steps were undertaken.

i. To design, fabricate and install a pilot scale Turbulent Contact Absorber.
ii. To determine liquid holdup.
iii. To determine pressure drop over a large range of liquid and gas velocities for different packing diameter, density and height.
iv. To determine expanded bed height so that free board above packing can be specified.
v. To determine over all volumetric gas phase transfer coefficients by adiabatic humidification method.
vi. All the steps ii to v are to be repeated for the tower without packing in order to exclude the hydrodynamic and mass transfer contribution of the gas distribution section (plenum). This may lead to the reason of the enormous diversity in the data reported on TCAs’.
vii. To compare results of the present study with those in literature and find out the reasons of discrepancies in results.
viii. To develop correlations for hydrodynamics and mass transfer for design of practical TCA columns.

1.7 Outline of the thesis

Chapter 1: This chapter is about introduction to the turbulent contact absorbers. It includes historical background, advantages, disadvantages and objectives of the present study.
Chapter 2: Hydrodynamics of TCA is reviewed in this chapter. Review includes pressure drop, liquid holdup, minimum fluidization velocity and expanded bed height.

Chapter 3: It includes review of gas phase, liquid phase transfer coefficients (both individual and volumetric) and interfacial area in turbulent contact absorbers.

Chapter 4: Experimental setup and measurement methods are mentioned in this chapter.

Chapter 5: Mass transfer theory along with mathematical model used to determine the gas phase transfer coefficients is included in this chapter.

Chapter 6: Results of liquid holdup, pressure drop and expanded bed height and their comparison with correlations and data given in literature are given in this chapter.

Chapter 7: Results of gas phase mass transfer coefficients and their comparison with correlations and data available in literature are given in this chapter.

Chapter 8: Conclusions and future recommendations are mentioned in this chapter.
Chapter 2

2 Review of Hydrodynamics in TCA

2.1 Introduction

A number of investigators have experimentally determined the hydrodynamic characteristics of the TCAs'. The hydrodynamic characteristics include pressure drop, liquid holdup and expanded bed height. Different experimental methods were used for the determination of these hydrodynamic characteristics. Hence the results reported by different investigators differ in values as well as trends (especially for liquid holdup). Correlations available for pressure drop, liquid holdup and expanded bed height determination are given here and the values determined from these correlations for specific case are compared.

2.2 Pressure Drop

Pressure drop in any mass transfer device determines its operating cost, liquid holdup and hence interfacial area, all of which are key parameters for design.

In literature, dependence of pressure drop on liquid and gas velocities, diameter and density of packing, static bed height and support grid free area have been studied. There is consensus in literature that pressure drop in TCA increases with increasing liquid flow rate at a fixed gas flow rate (Blyakher et al. 1967; Chen and Douglas 1968; Gelperin et al. 1968; Balabekov et al. 1969; Tichy and Douglas 1972; Tichy and Douglas 1973; Wozniak 1977; Miconnet et al. 1982; Rama et al. 1983; Gimenes 1992; Hekmat-Nazemi 1992).

There is also consensus that the TCA pressure drop increases with increasing static bed height (Kito 1976; Wozniak 1977).

Increase in TCA pressure drop with decrease in packing diameter is concluded by Vunjak-Novakovic et.al(1987b) while Tichy et.al (1972; 1973) observed no effect of packing size on pressure drop.

Some investigators observed an increase in pressure drop with increasing gas velocity (Blyakher et al. 1967; Levsh et al. 1968a; Balabekov et al. 1971; Gelperin et al. 1976; Miconnet et al. 1982; Inayat 1995; Soundarajan and Krishnaiah 1999; Lyashuk 2001) while
Others (Barile and Meyer 1971; Handle 1976; Uchida et al. 1977; Rama et al. 1983; El-Dessouky 1993; Zahedi et al. 2006) did not observe any effect of gas velocity on pressure drop.

Earlier authors worked on columns with grid free area less than 60% while later with more than 60%. Therefore, the effect of gas velocity on pressure drop can be divided into two categories i.e. for low grid free area and for high grid free area. The increase in pressure drop with gas velocity for low grid free area is due to the enhanced liquid retained on such grids causing more pressure drop.

Theoretical model have also been developed to predict pressure drop in TCA. The first model was developed by Wozniak (1977). His model is based on a steady state macroscopic force balance on TCA bed. Such a force balance is given as;

\[ \Delta p_b = (\rho_p \varepsilon_{p,at} + \rho_l \varepsilon_{l,at} + \rho_g \varepsilon_{g,at})gH_0 + (-\Delta p_f) \]  

(2.1)

Where

\( \Delta p_b \) = pressure drop across bed

\( \rho_p, \rho_l, \rho_g \) = Particles, liquid and gas densities respectively.

\( \varepsilon_{p,at}, \varepsilon_{l,at}, \varepsilon_{g,at} \) = Particles, liquid and gas fractions based on static bed height of packing

\( H_0 \) = Static bed height

\( \Delta p_f \) = The sum of all pressure drops due to forces of gas on dry grid, walls of the column, mechanical mixing of three phase and the force for overcoming surface tension in the development of interfacial area in the bed.

Neglecting weight of the gas in the first term and all other forces in second term except due to grid, the above equation (2.1) becomes;

\[ -\Delta p_b = (\rho_p \varepsilon_{p,at} + \rho_l \varepsilon_{l,at})gH_0 + (-\Delta p_0) \]  

(2.2)

Where, \( \Delta p_0 \) is pressure difference across grid.

Bruce et al (2006) also developed a model for pressure drop, liquid holdup, gas holdup and bed expansion in Turbulent Contact Absorber. He based his model on the model developed for pressure drop by Chern et al. (1983) for a three phase counter current packed bed. Grid free area is not incorporated in the model, inspite of that there is an increase in pressure drop with increase in gas velocity. Variables and their ranges by different investigators for TCA hydrodynamic study are mentioned in Table 2-1.
Table 2-1: Variables used in literature for hydrodynamic studies

<table>
<thead>
<tr>
<th>Investigators</th>
<th>H₀ (McMichael et al.)</th>
<th>uᵣ (m/s)</th>
<th>uᵣ (m/s)</th>
<th>Dc (cm)</th>
<th>dp (mm)</th>
<th>ρₚ (kg/m³)</th>
<th>Measured parameters</th>
</tr>
</thead>
<tbody>
<tr>
<td>Zahedi et al. (2006)</td>
<td>0.70</td>
<td>30</td>
<td>0.0094-0.014</td>
<td>0-3</td>
<td>15</td>
<td>15</td>
<td>250</td>
</tr>
<tr>
<td>Bruce et al. (2004)</td>
<td>0.7-0.82</td>
<td>7-22.6</td>
<td>0.0014-0.0096</td>
<td>1-5</td>
<td>11.3</td>
<td>12-25</td>
<td>110-835</td>
</tr>
<tr>
<td>Lyashuk (2001)</td>
<td>0.21-0.40</td>
<td>10-22</td>
<td>0.008-0.024</td>
<td>0.4-4.5</td>
<td>20</td>
<td>20</td>
<td>600</td>
</tr>
<tr>
<td>Shackley (2000)</td>
<td>0.41-0.96</td>
<td>3.8-21</td>
<td>0-0.0167</td>
<td>0.9-3.7</td>
<td>46</td>
<td>25-45 &amp; ellipsoids</td>
<td>168-640</td>
</tr>
<tr>
<td>Soundarajan and Krishnaiah (1994; 1998b;</td>
<td>0.155-0.728</td>
<td>11.3-51</td>
<td>0-0.032</td>
<td>0-3</td>
<td>11.3</td>
<td>12-25</td>
<td>110-995</td>
</tr>
<tr>
<td>1998a; 1999)</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Inayat (1995)*</td>
<td>0.72</td>
<td>10.5</td>
<td>0-0.011</td>
<td>0-3.6</td>
<td>22</td>
<td>25, 38, 45, 38x50 oblate spheroids</td>
<td>150-327</td>
</tr>
<tr>
<td>El-Dessouky (1993)</td>
<td>0.708</td>
<td>30-50</td>
<td>0.0013-0.008</td>
<td>0.34-0.65</td>
<td>20</td>
<td>12.7</td>
<td>375</td>
</tr>
<tr>
<td>Hekmat-Nazemi (1992)</td>
<td>0.72</td>
<td>10.5-16.5</td>
<td>0-0.0088</td>
<td>0-4</td>
<td>22</td>
<td>25, 38, 38x50 oblate spheroids</td>
<td>161-327</td>
</tr>
<tr>
<td>Gimenes (1992)</td>
<td>0.84</td>
<td>13.5-38</td>
<td>0.0015-0.0169</td>
<td>0-4</td>
<td>26.4</td>
<td>38,38x5 oblate spheroids</td>
<td>159-162</td>
</tr>
<tr>
<td>Paterson and Clift (1987)*</td>
<td>0.405-0.615</td>
<td>14.6-58.6</td>
<td>0.0055-0.0222</td>
<td>1-4</td>
<td>45.7</td>
<td>38</td>
<td>432</td>
</tr>
<tr>
<td>Vunjak-Novakovic et al. (1987a; 1987b)*</td>
<td>0.36-0.78</td>
<td>10-30</td>
<td>0-0.034</td>
<td>0-4</td>
<td>14-29</td>
<td>10-38</td>
<td>182-980</td>
</tr>
<tr>
<td>Rama et al (1983)*</td>
<td>0.70</td>
<td>12-48</td>
<td>0.011-0.044</td>
<td>0-5</td>
<td>15</td>
<td>23.7-38</td>
<td>53-183</td>
</tr>
<tr>
<td>Miconnet et al. (1981; 1982)</td>
<td>0.82</td>
<td>10-30</td>
<td>0-0.0075</td>
<td>0-6</td>
<td>30</td>
<td>20-38</td>
<td>86.6-806</td>
</tr>
<tr>
<td>Kuroda and Tabei (1981)</td>
<td>0.705-0.84</td>
<td>10-20</td>
<td>0-0.03</td>
<td>0-umᵣₘᵠf</td>
<td>10</td>
<td>10-28.5</td>
<td>170-760</td>
</tr>
<tr>
<td>Uchida et al. (1980)</td>
<td>0.72</td>
<td>10.5</td>
<td>0-0.03</td>
<td>0.5-5.7</td>
<td>16</td>
<td>30-38</td>
<td>83-173</td>
</tr>
<tr>
<td>Uysal (1978)</td>
<td>0.87</td>
<td>29-58</td>
<td>0.0047-0.033</td>
<td>0.5-5.5</td>
<td>29</td>
<td>19-38</td>
<td>157</td>
</tr>
<tr>
<td>Investigators</td>
<td>f</td>
<td>(H_0) (McMichael et al.)</td>
<td>(u_l) (m/s)</td>
<td>(u_g) (m/s)</td>
<td>(D_c) (cm)</td>
<td>(d_p) (mm)</td>
<td>(\rho_p) (kg/m³)</td>
</tr>
<tr>
<td>---------------------------------------</td>
<td>--------</td>
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<td>---------------</td>
<td>--------------</td>
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<td>-------------------</td>
</tr>
<tr>
<td>Kito et al. (1978)*</td>
<td>0.712-0.84</td>
<td>10-30</td>
<td>0.003-0.035</td>
<td>(\varepsilon_l^a) 3.5</td>
<td>10</td>
<td>9.7-28.5</td>
<td>170-1250</td>
</tr>
<tr>
<td>Wozniak (1977)</td>
<td>0.6</td>
<td>20-41</td>
<td>0.005-0.027</td>
<td>1.7-3.1</td>
<td>20</td>
<td>19</td>
<td>266</td>
</tr>
<tr>
<td>Uchida et al. (1977)</td>
<td>n/m</td>
<td>12.7-19</td>
<td>0.0086-0.025</td>
<td>0-4</td>
<td>19.5</td>
<td>12.7-19</td>
<td>170</td>
</tr>
<tr>
<td>Kito et al. (1976a; 1976b)</td>
<td>0.705-0.84</td>
<td>5-20</td>
<td>0-0.035</td>
<td>0-3.7</td>
<td>10</td>
<td>9.7-28.5</td>
<td>1250</td>
</tr>
<tr>
<td>Gelperin et al. (1976)</td>
<td>0.3-0.5</td>
<td>10,15</td>
<td>20,30</td>
<td>200,550</td>
<td>153-458</td>
<td>1250</td>
<td>(\Delta P, \varepsilon_l^a), (\varepsilon_l^b), (\varepsilon_l^c)</td>
</tr>
<tr>
<td>Handle (1976)</td>
<td>0.36-0.78</td>
<td>29</td>
<td>20,38</td>
<td>200,400</td>
<td>153-458</td>
<td>1250</td>
<td>(\Delta P, \varepsilon_l^a), (\varepsilon_l^b), (\varepsilon_l^c)</td>
</tr>
<tr>
<td>Tarat (1974)</td>
<td>0.40</td>
<td>40,100</td>
<td>36</td>
<td>22.6</td>
<td>(\Delta P, \varepsilon_l^a), (\varepsilon_l^b), (\varepsilon_l^c)</td>
<td></td>
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<tr>
<td>Strumillo et al. (1974)</td>
<td>0.65</td>
<td>5-30</td>
<td>0.0056-0.0333</td>
<td>1-6</td>
<td>20</td>
<td>15-30</td>
<td>440-1050</td>
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<tr>
<td>Tichy and Douglas (1972; 1973),</td>
<td>0.78</td>
<td>14-28</td>
<td>0-0.032</td>
<td>0-5.4</td>
<td>14-29.2</td>
<td>12.3-19.1</td>
<td>153-458</td>
</tr>
<tr>
<td>Tichy et al. (1972)</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Balabevkov et al. (1971)</td>
<td>0.30-0.60</td>
<td>6-12</td>
<td>0.0-0.035</td>
<td>0.1-8.0</td>
<td>17.5</td>
<td>6-22</td>
<td>356-1036</td>
</tr>
<tr>
<td>Khanna (1971)</td>
<td>0.7</td>
<td>14</td>
<td>0.0054-0.0325</td>
<td>0-4</td>
<td>14</td>
<td>12.7-38.1</td>
<td>150</td>
</tr>
<tr>
<td>Barile and Meyer (1971)</td>
<td>0.82</td>
<td>15-53</td>
<td>0.013-0.038</td>
<td>0.5-5.1</td>
<td>28.6</td>
<td>19.38</td>
<td>109,160</td>
</tr>
<tr>
<td>(Aksel’rod and Yakovenko (1969)</td>
<td>0.30-0.65</td>
<td>4.5-22.5</td>
<td>0.0055-0.0417</td>
<td>1-7.5</td>
<td>15</td>
<td>16-29</td>
<td>170-470</td>
</tr>
<tr>
<td>Mayak and Matrozov (1969)</td>
<td>0.393-0.61</td>
<td>1.8-10</td>
<td>0.0056-0.0253</td>
<td>2.23-4.48</td>
<td>12</td>
<td>8-12</td>
<td>800-900</td>
</tr>
<tr>
<td>Gelperin et al. (1968)</td>
<td>0.25-0.7</td>
<td>9-26</td>
<td>0.0056-0.0667</td>
<td>0.5-8</td>
<td>6-37</td>
<td>6-35.5</td>
<td>160-910</td>
</tr>
<tr>
<td>Levsh (1968a)</td>
<td>0.29-0.50</td>
<td>7-21</td>
<td>0.0028-0.017</td>
<td>0.5-4.5</td>
<td>17.8</td>
<td>Rings18(\times9\times2)</td>
<td>1083</td>
</tr>
<tr>
<td>Chen and Douglas (1968)*</td>
<td>n/m</td>
<td>30.5</td>
<td>0.0025-0.015</td>
<td>0-3.2</td>
<td>30.5</td>
<td>12.5-37</td>
<td>156-170</td>
</tr>
<tr>
<td>Blyakher et al. (1967)</td>
<td>0.19-0.90</td>
<td>15-70</td>
<td>0-0.0144</td>
<td>0.4-5</td>
<td>20-35</td>
<td>38</td>
<td>90-180</td>
</tr>
</tbody>
</table>

\(\varepsilon_l^a\) = Liquid holdup determined by shut off valve technique  
\(\varepsilon_l^b\) = Liquid holdup determined by tracer technique  
\(\varepsilon_l^c\) = Liquid holdup determined by pressured drop measurements  
n/m = Not mentioned  
* = Authors who either have mentioned or from their experimental setup, it can be concluded that they have excluded liquid held in plenum.
Table 2-2 shows a summary of the pressure drop correlations developed by various researchers for the experimental data obtained by them in TCAs’.

**Table 2-2: Pressure drop correlations in literature**

<table>
<thead>
<tr>
<th>Investigators</th>
<th>Correlations</th>
</tr>
</thead>
<tbody>
<tr>
<td>Lyashuk (2001)</td>
<td>( \Delta p = 163u_l^{0.34} \rho_l^{0.34} u_g^{0.29} H_0^{0.38} f^{-1.35} )</td>
</tr>
<tr>
<td>Ramaet al. (1983)</td>
<td>( \Delta p = \rho_p (1 - \varepsilon_0) gH_0 + 12.8 \left( \frac{H_0}{d_p} \right)^{0.4} \left( \frac{d_f}{D} \right)^{-0.58} \frac{Fr_l^{1.66} Ga_l^{0.09}}{\rho_l} )</td>
</tr>
<tr>
<td>Wozniak (1977)</td>
<td>( \Delta p = \rho_p (1 - \varepsilon_0) gH_0 + 467200 \left( \frac{H_0}{D_c} \right)^{0.4515} \left( \frac{Re_l}{\rho_l} \right)^{1.798} \frac{\rho_g u_g^2}{\rho_l} )</td>
</tr>
<tr>
<td>Uchida et al. (1977)</td>
<td>( \Delta p = \rho_p (1 - \varepsilon_0) gH_0 + 9.38 \times 10^8 \frac{\mu_l^2 f^{-0.42}}{D_c^{0.84} d_p^{0.84} \rho_p^{0.18} H_0 u_l} )</td>
</tr>
<tr>
<td>Gelperin et al. (1976)</td>
<td>( \Delta p = \rho_p (1 - \varepsilon_0) gH_0 + 0.536u_l^{0.14} \rho_l^{0.14} u_g^{0.24} \rho_p^{-0.1} H_0^{0.92} )</td>
</tr>
<tr>
<td>Handle (1976)</td>
<td>( \Delta p = \rho_p (1 - \varepsilon_0) gH_0 + 26.8 \varepsilon_0 Re_l^{-0.189} Fr_l^{0.948} gH_0 \rho_l )</td>
</tr>
<tr>
<td>Tarat (1974)</td>
<td>( \Delta p = \rho_p (1 - \varepsilon_0) gH_0 + 0.824u_l^{0.96} \rho_l^{1.04} u_g^{1.04} \rho_g^{1.04} )</td>
</tr>
<tr>
<td>Tichy et al. (1972)</td>
<td>( \Delta p = \frac{H_0 \rho_g u_g^2}{2D_c^2} \left( 1 + Cu_l \rho_l \right) \left( \frac{4^{0.03 - 2.44u_l \rho_l + 0.84u_g \rho_g - 0.127u_l^{2} \rho_l^{2}}}{k} \right) )</td>
</tr>
<tr>
<td>C = 0.1083, 0.0816 and 0.078 for ( H_0/D_c = 1, 1.5 ) and 2 respectively</td>
<td></td>
</tr>
<tr>
<td>Balabekov et al (1971)</td>
<td>( \Delta p = \rho_p (1 - \varepsilon_0) gH_0 [1 + C \left( \frac{P_l u_l^2}{\rho_g u_g^2} \right)^{k+1}] )</td>
</tr>
<tr>
<td>Where: ( C = 3.4733 \times 10^{-7} \frac{\rho_l^{2.74} u_l^2}{f^3 d_p \rho_p^{0.74}} ), ( k = -2.6838 f^{0.16} d_p^{-0.1} )</td>
<td></td>
</tr>
<tr>
<td>Note: Dry grid hydraulic resistance given in original paper is neglected here.</td>
<td></td>
</tr>
<tr>
<td>Barile and Meyer (1971)</td>
<td>( \Delta p = \rho_p (1 - \varepsilon_0) gH_0 + \rho_g gH_0 \left( 1160 Fr_l^{1.56} Re_l^{-0.51} \left( \frac{H_0}{d_p} \right)^{-0.36} \right) )</td>
</tr>
<tr>
<td>Aksel'rod and Yakovenko (1969)</td>
<td>( \Delta p = (1 + 39u_l) \rho_p gH_0 ) (neglecting pressure drop due to grid as it was not mentioned clearly in correlation)</td>
</tr>
</tbody>
</table>

Figure 2-1 and Figure 2-2 represent these correlations in graphical form to point out the huge diversity in the experimental data reported. Most of the work done in literature is on small diameter columns (Table 2-1) and the corresponding correlations are specific to the systems studied. This large variation in pressure drop necessitates further investigation in the pressure drop characteristics of TCA. Hence, in the present study, pressure drop in a large scale column of diameter 44.7 cm was determined so as to exclude the effect of column
diameter if any. This reported experimental data on a 44.7 cm column will be more worthy for use in designing industrial columns as compared to the data reported on small diameter columns.

Figure 2-1 Comparison of pressure drop calculated from literature correlations for Type 1 fluidization

Figure 2-2 Comparison of pressure drop calculated from literature correlations for Type 2 fluidization
### 2.3 Liquid Holdup

In literature, liquid holdup in TCA is taken as total volume of liquid held in the bed (Paterson and Clift 1987), volume of liquid held in the bed per unit expanded bed volume (Chen and Douglas 1968; Lyashuk 2001) and volume of liquid retained in the bed per unit static bed volume (Kito et al. 1978; Rama et al. 1983; Vunjak-Novakovic et al. 1987a; Gimenes 1992; Inayat 1995; Soundarajan and Krishnaiah 1998a; Bruce et al. 2004). Thus the literature supports the definition of liquid holdup as follows

\[ \varepsilon_{l,\text{st}} = \frac{V_f}{A_c H_o} \]  

(2.3)

Where \( V_f \) is the total volume of liquid retained in the bed and \( A_c \) is the area of the column.

Liquid holdup based on expanded bed height (\( H \)) can be calculated as follows

\[ \varepsilon_{l,e} = \frac{H_o \varepsilon_{l,\text{st}}}{H} \]  

(2.4)

Liquid holdup determines interfacial area, pressure drop and mass transfer efficiency in any mass transfer equipment.

Three methods have been reported in the literature for the determination of liquid holdup in TCA for both types of fluidization regimes. These are

i. Shut off valve
ii. Tracer Technique
iii. From pressure drop measurements

There are significant differences among the results reported by different investigators. Inayat (1995), Vunjak-Novakovic et al. (1987a), Gimenes (1992) and Kito et al. (1978) excluded the liquid holdup in the plenum section. However, Vunjak-Novakovic (1987a) and Kito et al. (1978) did not mention whether they excluded the liquid volume held in liquiddistributor or not. Gel'perin (1968), Hekmat-Nazemi (1992), Soundarajan and Krishnaiah (1998a), Lyashuk (2001), and Bruce et al. (2004) measured liquid holdup by shut off valve technique but did not mention about the liquid volume held in plenum or liquid distributor along with its pipe. The determination of liquid holdup by shut off valve technique is more reliable than the tracer method or pressure drop method (Bruce et al. 2004). Correlations available in literature for liquid holdup are presented in Table 2-3.
### Table 2-3: Liquid holdup correlations in literature

<table>
<thead>
<tr>
<th>Investigators</th>
<th>Correlations</th>
</tr>
</thead>
<tbody>
<tr>
<td>Bruce et al (2004)</td>
<td>( \varepsilon_{i,\text{st}} = 2.01 Fr_t^{1.06} Re_t^{-0.655} \left( \frac{H_0}{d_p} \right)^{-0.853} \left( \frac{f d}{D} \right)^{-0.203} )</td>
</tr>
<tr>
<td></td>
<td>( \varepsilon_{i,\text{st}} = 1.36 Fr_t^{1.44} Re_t^{-0.927} \left( \frac{H_0}{d_p} \right)^{-0.593} \left( \frac{f d}{D} \right)^{-0.213} )</td>
</tr>
<tr>
<td>Lyashuk (2001)</td>
<td>( \varepsilon_{i,\text{st}} = 0.001648 \rho_i^{0.16} u_i^{0.05} F_l^{0.95} H_0^{-1.09} f^{-2.02} )</td>
</tr>
<tr>
<td>Shackley (2000)</td>
<td>( \varepsilon_{i,\text{st}} = 0.00035 L^{0.7} u_g^{0.75} H_0^{-0.92} f^{-2.5} )</td>
</tr>
<tr>
<td>Soundarajan and Krishnaiah (1998a)</td>
<td>Type1: ( \varepsilon_{i,\text{st}} = 9.5 Ga_t^{0.09} Fr_t^{1.66} Re_t^{-0.34} We_t^{-0.34} \left( \frac{H_0}{d_p} \right)^{-0.4} \left( \frac{f d}{D} \right)^{-0.58} u_g^{0.57} ) Type2: ( \varepsilon_{i,\text{st}} = 7.705 Ga_t^{0.09} Fr_t^{1.66} Re_t^{-0.34} We_t^{-0.34} \left( \frac{H_0}{d_p} \right)^{-0.4} \left( \frac{f d}{D} \right)^{-0.58} u_g^{0.57} )</td>
</tr>
<tr>
<td>Gimenes (1992)</td>
<td>For plain spheres and oblate spheroids: ( \varepsilon_{i,\text{st}} = 0.001884 \exp \left[ 0.237 u_g \left( \rho_i u_i \right)^{0.616} H_0^{-0.357} \left( \frac{d_p}{\xi} \right)^{-0.411} \right] ) For perforated spheres: ( \varepsilon_{i,\text{st}} = 0.00123 \exp \left[ 0.185 u_g \left( \rho_i u_i \right)^{0.622} H_0^{-0.197} \left( \frac{\phi d_p}{d_o} \right)^{-0.676} \right] )</td>
</tr>
<tr>
<td>Vunjak-Novakovic et al. (1987a)</td>
<td>Type1: ( \varepsilon_{i,\text{st}} = 6.49 Fr_t^{0.855} Re_t^{-0.139} \left( \frac{H_0}{D_e} \right)^{0.567} ) Type2: ( \varepsilon_{i,\text{st}} = 7.33 Fr_t^{0.87} Re_t^{-0.059} \left( \frac{H_0}{D_e} \right)^{-0.433} \left( \frac{\rho_i}{\rho_p} \right)^{0.09} )</td>
</tr>
<tr>
<td>Paterson and Clift (1987)</td>
<td>( \varepsilon_{i,\text{st}} = 2.29 u_g^{-0.07} u_i^{0.71} H_0^{-0.52} f^{-0.874} ) (for 38mm dia balls, 40% voids)</td>
</tr>
<tr>
<td>Rama et al (1983)</td>
<td>( \varepsilon_{i,\text{st}} = 11 Ga_t^{0.09} Fr_t^{1.66} Re_t^{-0.34} We_t^{-0.34} \left( \frac{H_0}{d_p} \right)^{-0.4} \left( \frac{f d}{D} \right)^{-0.58} + 0.086 )</td>
</tr>
<tr>
<td>Kito et al (1978)</td>
<td>( \varepsilon_{i,\text{st}} = 12.8 Ga_t^{0.09} Fr_t^{1.66} Re_t^{-0.34} We_t^{-0.34} \left( \frac{H_0}{d_p} \right)^{-0.4} \left( \frac{f d}{D} \right)^{-0.58} )</td>
</tr>
<tr>
<td>Chen and Douglas (1968)</td>
<td>( \varepsilon_{i,\text{st}} = 0.00237 u_i^{0.6} \rho_i^{0.6} d_p^{-0.5} )</td>
</tr>
<tr>
<td>Gel'perin et al. (1968)</td>
<td>( \varepsilon_{i,\text{st}} = 0.138 u_i^{0.5} \rho_i^{0.5} d_p^{-0.5} )</td>
</tr>
</tbody>
</table>

In all these correlations, the operational liquid holdup is based on static bed height. For the total liquid holdup, a factor of 0.02 is added in operational liquid holdup to incorporate the liquid stuck to the packing (Vunjak-Novakovic et al. 1987a). Liquid holdup calculated by literature correlations are compared in Figure 2.3 for type 1 mode of operation and in Figure 2.4 for type 2 mode of TCA operation.
Figure 2-3 Comparison of liquid holdup calculated from literature correlations for Type 1 fluidization

Figure 2-4 Comparison of liquid holdup calculated from literature correlations for Type 2 fluidization
Literature data (this includes those researchers who have reported their experimental data either in tabular or graphical form but have not put forward any correlation) show that liquid holdup increases with increasing liquid velocity (Chen and Douglas 1968; Gel'perin et al. 1968; Kito et al. 1978; Rama et al. 1983; Paterson and Clift 1987; Vunjak-Novakovic et al. 1987a; Gimenes 1992; Inayat 1995; Soundarajan and Krishnaiah 1998a; Lyashuk 2001; Bruce et al. 2004).

Liquid holdup also increases with increasing density of packing (Miconnet et al. 1981; Miconnet et al. 1982; Vunjak-Novakovic et al. 1987a; Soundarajan and Krishnaiah 1998a; Bruce et al. 2004).

Liquid holdup decreases with increasing static bed height (Uchida et al. 1977; Wozniak 1977; Vunjak-Novakovic et al. 1987a; Gimenes 1992; Soundarajan and Krishnaiah 1998a; Lyashuk 2001; Bruce et al. 2004), however, the total liquid held increases with increasing static bed height.

Effect of grid free area on liquid holdup studied in literature shows an increase in liquid holdup with decreasing grid free area (Aksel'rod and Yakovenko 1969; Paterson and Clift 1987; Vunjak-Novakovic et al. 1987a; Lyashuk 2001; Bruce et al. 2004).

There is a contradiction in literature as far as effect of gas velocity on liquid holdup is concerned. Chen and Douglas (1968), Kito et al. (1978), Vunjak-Novakovic et al (1987a) and Bruce et al (2004) reported no significant effect of gas velocity on liquid holdup while Rama et al (1983), Hekmat-Nazemi (1992), Inayat (1995), Gimenes and Handley (1998), Soundarajan and Krishnaiah (1998a) and Lyashuk (2001) reported increase in liquid holdup with gas velocity. However, there is consensus in literature that for type-2 fluidization, there is an increase in liquid holdup before the bed is fully fluidized.

2.4 Expanded Bed Height

For designing Turbulent Contact Absorber, expected expanded bed height is required for deciding the height of the column or spacing between the two grids in case of multi-stage columns. Smaller distance between the support grids may cause flooding and high pressure drop since the packing sticks to the upper grid at high gas velocities in such cases. Height of the column in excess of the required height increases the capital cost. Variables affecting bed expansion are: gas flow rate, liquid flow rate, free area of the grid, density and diameter of the packing and static bed height. In literature, expanded bed height was determined by visual
observations and in case of bed pulsation, maximum and minimum heights of the bed were noted and average height was reported (Chen and Douglas 1968; Tichy and Douglas 1972; Vunjak-Novakovic et al. 1987b; Soundarajan and Krishnaiah 1998a). Correlations for bed expanded height are given in Table 2-4. Expanded bed heights are calculated from correlations developed in literature and compared in Figure 2.5

<table>
<thead>
<tr>
<th>Investigators</th>
<th>Correlations</th>
</tr>
</thead>
<tbody>
<tr>
<td>Lyashuk (2001)</td>
<td>[ \frac{H}{H_0} = 0.16u_g^{0.44}L^{0.27}H_0^{-0.71}f^{-1.54} ]</td>
</tr>
<tr>
<td>Shackley (2000)</td>
<td>[ \frac{H}{H_0} = 0.0833(-14.9f^2 + 15.7f - 2.1)H_0^{-0.34}L^{0.26}\rho_p^{-0.43}u_g^{0.78}d_p^{-0.85} ]</td>
</tr>
<tr>
<td>Gimenes and Handley (1998)</td>
<td>[ \frac{H}{H_0} = 1.862d_p^{0.306}(1 - \frac{\phi}{\varepsilon})^{3.37}L^{0.108}(1000\rho_p\nu)^{-0.107}\exp(0.356u_g) ]</td>
</tr>
<tr>
<td>Vunjak-Novakovic et al (1987b)</td>
<td></td>
</tr>
<tr>
<td><strong>For type 1:</strong></td>
<td>[ \frac{H}{H_0} = \frac{(1 - \varepsilon_0) + 0.00248(H_0 \frac{H_0}{D_p})^{-0.567}d_p^{-0.568}L^{0.719} + 0.02}{(1 - 0.628u_g^{0.237})} ]</td>
</tr>
<tr>
<td><strong>For type 2:</strong></td>
<td>[ \frac{H}{H_0} = \frac{(1 - \varepsilon_0) + 0.00443(H_0 \frac{H_0}{D_p})^{-0.437}d_p^{-0.494}L^{0.812}(\rho_p^{\rho})^{0.090} + 0.02}{(1 - 0.628u_g^{0.237})} ]</td>
</tr>
<tr>
<td>Uysal (1978)</td>
<td>[ \frac{H}{H_0} = 1 + \frac{0.147(\rho_p u_g - \rho_g u_{mf})}{H_0} \text{ where } u_{mf} = \frac{10.86d_p^{0.488}10^{-0.01985u_g\rho_g}}{\rho_g} ]</td>
</tr>
<tr>
<td>O'Neill et al (1973)</td>
<td>[ \frac{H}{H_0} = \frac{1 - \varepsilon_0}{1 - \varepsilon} \text{ where } \varepsilon = (K\frac{(27 + 4K)}{108})^{1/2} + (K\frac{(27 + 4K)}{108})^{1/3} - (K\frac{(27 + 4K)}{108})^{1/2} - K^{1/3} \text{ and } u_g^{1/2} + u_f^{1/2}(\frac{\rho_f}{\rho_g})^{1/4} + 0.775\frac{gd_p\rho_f}{6\rho_g} \text{ for } u_g &lt; 6 \text{ m/s} ]</td>
</tr>
<tr>
<td>Khanna (1971)</td>
<td>[ \frac{H}{H_0} = 1 + 0.414(\rho_g u_g - \rho_g u_{mf})(\rho_g u_{mf})^{0.2} \text{ where } u_{mf} = \frac{526.47d_p^{1.5}10^{-0.0317u_g\rho_g}}{\rho_g} ]</td>
</tr>
<tr>
<td>Aksel'rod and Yakovenko (1969)</td>
<td>[ \frac{H}{H_0} = 0.933u_t^{0.3}H_0^{-0.4}(\frac{u_g}{f})^{0.93} \text{ for } \frac{u_g}{f} \geq 6 \text{ m/s} ]</td>
</tr>
<tr>
<td></td>
<td>[ \frac{H}{H_0} = 2.16u_t^{0.3}H_0^{-0.4}(\frac{u_g}{f})^{0.43} \text{ for } \frac{u_g}{f} &lt; 6 \text{ m/s} ]</td>
</tr>
<tr>
<td>Levsh et al. (1968a)</td>
<td>[ H = 4.4u_t^{0.43}u_g^2 \text{ for } u_t &lt; 0.0078m/s, u_g &lt; 2.5 \text{ m/s} ]</td>
</tr>
<tr>
<td></td>
<td>[ H = 2.28u_t^{0.35}u_g^2 \text{ for } u_t &gt; 0.0078 \text{ m/s, } u_g &gt; 2.5 \text{ m/s} ]</td>
</tr>
</tbody>
</table>
Figure 2-5 Comparison of expanded bed height correlations

- $d_e = 0.038 \text{ m}$
- $H_o = 0.25 \text{ m}$
- $\rho_g = 180 \text{ kg/m}^3$
- $u_i = 0.008 \text{ m/s}$

1. Vunjak et al. 1987
2. Uysal 1978
3. Gimenes and Handley 1998
4. Lyashuk 2001
5. O'Neill et al. 1973
6. Aksef'rod 1969
7. Soundararajan and Krishnaiah 1998
Chapter 3

3 Review of mass transfer in TCA

3.1 Introduction

In this chapter, various mass transfer characteristics of TCA as studied by various investigators are discussed. This chapter also includes the chemical or physical methods used to experimentally determine these mass transfer characteristics. These mass transfer characteristics include the interfacial area and the individual and volumetric mass transfer coefficients both on the liquid and gas sides. The available empirical correlations developed so far have also been reviewed.

3.2 Interfacial area, volumetric liquid film and overall mass transfer coefficients

To experimentally determine the liquid phase mass transfer coefficients, various methods have been reported in literature. Liquid phase transfer coefficients are determined by absorption or desorption of poorly soluble gases in liquids. Desorption of O₂ from water, absorption of CO₂ in NaOH and absorption of O₂ in Na₂SO₃ solution have been used for liquid film transfer coefficients in Turbulent Contact Absorbers (Elenkov and Kosev 1970; Hekmat-Nazemi 1992; Inayat 1995).

Levsh et al (1968b) studied desorption of O₂ from water in a transparent column. Rings made of polymeric materials were used as fluidized packing. Volumetric mass transfer coefficients evaluated per unit operating volume were 2-20 times larger than the ordinary absorbers. The values of volumetric liquid film mass transfer coefficients decreased with increasing static bed height and gas velocity. This decrease with superficial gas velocity was larger for higher liquid velocities because of rapid increase in bed height at higher liquid velocities. \( k_a \) increased with increase in liquid velocity for constant \( H_0 \) and \( u_g \). The oxygen recovery coefficient increased with increase in superficial liquid and gas velocities. For gas velocities 2-5 m/s and liquid velocities 0.0097-0.017 m/s, the oxygen recovery coefficient was in the range of
0.78-0.95. For the same superficial gas velocity range and superficial liquid velocities 0.0028-0.017 m/s, $k_a$ ranged 0.0103-0.125 s$^{-1}$.

Elenkov and Kosev (1970) studied the effect of the support grid free area, packing density, packing diameter, static bed height, liquid and gas flow rates on $k_a$ by desorbing $O_2$ from water. They analyzed water for $O_2$ at inlet to the column and directly below the grid so that the mass transfer in the bed could be determined. They concluded that volumetric liquid film mass transfer coefficients in columns with mobile packing are 1.5 to 2 times greater than corresponding values for plate columns. They observed that $k_{lac}$ is a linear function of superficial liquid velocity up to a certain maximum value above which, it decreased. This decrease was due to increased bed height at higher liquid flow rates causing bed pulsations. The maxima of $k_{lac}$ shifted towards higher liquid flow rates as $u_g$ and $H_0$ increased and $f$ decreased. Above certain values of $u_g$ and $H_0$, no maxima of $k_{lac}$ was observed with increase in liquid flow rates. For $H_0=0.17$ m, $f=0.79$, there was a maxima of $k_{jac}$ at $u_g=4.82$ m/s and $u_l=0.0194$ m/s while at $u_g=5.2$ m/s and above, no maximum value of $k_{lac}$ was observed at any liquid flow rate. Similarly for $u_g=4.04$ m/s, $f=0.417$, no maxima of $k_{lac}$ was observed for static bed heights at and above 17 cm. The values of $k_{lac}$ observed were 0.014-0.21 m/s. Elenkov and Kosev (1970) developed the following correlation from their data such that 65% of the results obeyed the correlation within ±15%, 85% within ±20% and 95% within ±25%.

$$m_x = 0.024 \frac{D}{h_0 u_l} Re_L^{0.1} Re_G^{1.5} Sc^{0.5}$$

(3.1)

where

$m_x =$Number of liquid phase transfer units.

$h_0 = \frac{\Delta p - \Delta p_a}{\rho_l}$ =Determining linear dimension, m

$Re_L = \frac{h_0 u_l \rho_l}{\mu_l}$ = Reynolds number for liquid.

$Re_G = \frac{h_0 u_g \rho_g}{\mu_g}$ = Reynolds number for gas.

$Sc = \frac{\mu_l}{\rho_l D_l}$ = Schmidt number.
\( D_l \) = Diffusion coefficient of liquid, \( m^2/s \)  
\( \Delta p \) = Pressure drop across bed, \( N/m^2 \)  
\( \Delta p_0 \) = Pressure drop across grid, \( N/m^2 \)

Gel’perin et al. (1972) studied the effect of superficial liquid velocity, superficial gas velocity, grid free area and static bed height on interfacial area by absorbing \( CO_2 \) in NaOH/MEA (methyl ethanol amine) solutions in water. They neglected the gas phase resistance which contributed up to a maximum of 10% of the total resistance in mass transfer. They developed the following correlations for interfacial area in TCA.

\[
a_c = A_b \left( \frac{1.685 \times 10^5 f^{1.1}}{A_b^{1.55} u_g^{0.29}} \right)^H_0
\]

(3.2)

Where,

\[
A_b = 58.7 U_j^{0.44} \left( \frac{u_g}{f} \right)^{0.92}
\]

Interfacial area “\( a \)” per unit volume based on static bed height can be related with interfacial areas based on column cross-section, \( a_c \), based on expanded bed height \( a_e \) and based on liquid holdup \( a_l \) as follows.

\[
a_c = a_e = a_{e,H} = a_c \varepsilon_{l,ef} H_0
\]

Gel’perin et al. (1972) evaluated total contact area of the column and area of the separation section. The later was subtracted from the former to determine phase contact area in the bed. Interfacial area per unit volume of separation space was correlated as follows.

\[
a_s = 130 u_l^{0.35} u_g^{3.31}
\]

(3.3)

Separation space/section is the free area above expanded bed to mist eliminator.

Their data showed that phase contact surface in separation section \( A_s \) constituted from 5 to 30% of the total contact surface in the fluidized bed column. The experimental results of Gel’perin et al. (1972) indicated that there is significant increase in \( a_c \) with increase in static bed height. The values of the phase contact surface \( a_c \) were 40-140 \( m^2/m^2 \).

Wozniak and Ostergaard (1973) absorbed \( CO_2 \) from \( CO_2\)-air mixture in NaOH solution for determination of \( a \), \( k_g \alpha \) and \( K_g \alpha \). They used the following equation by
adding the resistances of mass transfer on both sides of the interface to evaluate overall mass transfer coefficient

\[
\frac{1}{K_g a} = \frac{H e}{a(D_{CO_2}k_COH^-)^{1/2}} + \frac{1}{k_g a}
\]

(3.4)

where

- \( H e \) = Henry’s constant, Pa/(kmole/m³)
- \( D_{CO_2} \) = Diffusivity of CO₂ in liquid phase, m²/s
- \( k_r \) = Rate of chemical reaction, m³/(kmoles-s)
- \( C_{OH^-} \) = Bulk concentration of NaOH, mean value over a run, kmoles/m³
- \( K_{ga} \) = Overall mass transfer coefficient, kmoles/(N-m-s)
- \( k_{ga} \) = Volumetric gas film mass transfer coefficient, kmoles/(N-m-s)

\( K_{ga} \) in above equation (3.4) was calculated by the following equation

\[
\bar{Ra} = K_{ga} p_{CO_2}
\]

(3.5)

Where,

- \( \bar{Ra} \) = Rate of absorption of CO₂, kmoles/(m³-s)
- \( p_{CO_2} \) = Partial pressure of CO₂, N/m²

Rate of absorption of CO₂ was determined at various concentrations of NaOH. A plot of \( 1/K_{ga} \) vs. \( He/(D_{CO_2}k_COH^-) \) gave interfacial area “a” and \( k_{ga} \) from the slope and y-intercept of the plot.

CO₂ was 4.5 % in inlet gas while molarity of NaOH varied from 0.15-1.5 kmole/m³. The system could be considered as liquid film controlled as 90 % of the mass transfer resistance lied in the liquid film and only 10 % in the gas film. Volumetric overall mass transfer coefficient \( K_{ga} \) increased with increasing liquid flow rate and NaOH concentration. Interfacial area “a” increased while the average gas film transfer coefficient “\( k_{ga} \)” decreased with increasing liquid flow rate.

The interfacial areas studied based on static bed volume were 600-1500 m²/m³ and based on whole system volume (which was not clearly mentioned in paper) were 120-300 m²/m³. Volumetric over all mass transfer coefficients \( K_{ga} \) were 0.028-0.083 kmoles/(m³-s-atm). Interfacial area “a” was found to be of the same order as in bubble columns and somewhat lower than in paddle and propeller agitators.
Wozniak (1977) determined interfacial area using the same method as by Wozniak and Ostergaard (1973) i.e. by equation (3.4). However, he determined $K_g a$ using the following equation instead of equation (3.5).

\[
K_g a = \frac{V_l (C_{l_f} - C_{l_p})}{2V_{sb} \left( B + \frac{P_1 + P_3}{2} \right) \left( \frac{C_{g_f} - C_{g_p}}{2} \right) t} 
\]

(3.6)

Where:

- $V_l$ = Volume of liquid used for absorption, m$^3$
- $V_{sb}$ = Volume of static bed, m$^3$
- $B$ = Atmospheric pressure, Pa
- $P_1, P_3$ = Pressure under the lower and above the upper plate of the column, Pa
- $C_l$ = NaOH concentration, kmole/m$^3$
- $C_g$ = Volumetric mole fraction of CO$_2$ in the gas phase.
- $t$ = Time of absorption, sec

Wozniak (1977) developed the following relationship for interfacial area in TCA.

\[
A = 6.189 \times 10^{-7} \left[ \frac{\varepsilon_g}{1 - \varepsilon_g} \right]^{-0.8022} \left[ \frac{H_c \Delta p}{u_g \mu_g} \right]^{-0.9337} 
\]

(3.7)

Where,

- $\varepsilon_g$ = Gas phase porosity of three phase bed.
- $\mu_g$ = Viscosity of gas N-s/m$^2$

$A = \frac{a}{a_b} = $ Dimensionless interfacial area.

$a =$ Interfacial area per unit static bed volume, m$^2$/m$^3$.

$a_b =$ Geometrical surface area of packing per unit static bed volume.

$H_c =$ Expanded bed height, m

The interfacial area $a$ determined in his experiments was 250-800 m$^2$/m$^3$

Strumillo and Kudra (1977) absorbed CO$_2$ from air in NaOH solution and determined interfacial area using Danckwerts model (Danckwerts 1970). The NaOH and CO$_2$ concentrations in their work were 2-3 N and 3% by volume respectively. Their experimental data show that increasing the superficial gas velocity and the static bed height up to a certain value causes an increase in interfacial area. At higher
superficial gas velocities and static bed heights, the interfacial area decreases, i.e. there is a maxima of interfacial area. They explained that it was because of pressure oscillations and non-homogeneity of the floating bed. They found that the interfacial area per unit column cross-section $a_c$ ranged from 15-100 m$^2$/m$^2$.

They developed the following empirical equation for determination of interfacial area per unit cross-section of the column.

$$a_c = 16.27u_g^{0.92}u_f^{0.34}H_0^{0.83}d_p^{-0.04}$$

Palaty (1989) absorbed CO$_2$ from air in aqueous solution of NaOH/Na$_2$CO$_3$ to determine interfacial area. Concentration of inlet CO$_2$ was maintained at 3-4 volume percent while 3N NaOH solution was made to circulate in the system. They conducted experiments in two columns A & B. Column A with diameter 5.8 cm contained support plate having 65 % free area while column B with diameter 10.9 cm contained support plate having 41 % free area. Packing diameter and density for column A was 6 mm and 1000 kg/m$^3$ while diameter and density of packing for column B was 10 mm and 420 kg/m$^3$ respectively. The specific interfacial area $a_e$ for absorber A varied from about 82-145 m$^{-1}$ while for absorber B, it varied from about 85-130 m$^{-1}$. Palaty (1989) developed the following correlation from experimental data of column A.

$$a_c = 234u_g^{0.35}u_f^{0.11}H_0^{0.12}$$

Palaty (1989) desorbed CO$_2$ from water into air and measured volumetric liquid film transfer coefficient $k_l$. Interfacial area “$a$” was already known from CO$_2$ absorption in aqueous NaOH/Na$_2$CO$_3$. If it is assumed that the interfacial area in air water system is almost the same as in CO$_2$-NaOH/Na$_2$CO$_3$ system then the $k_l$ values can be determined by division. The $k_l$ values thus determined were 2.2×10$^{-4}$ m/s to 4.4×10$^{-4}$m/s.

Hekmat-Nazemi (1992) absorbed 2% by volume CO$_2$ in NaOH solution and established interfacial area $a$ and liquid film transfer coefficient “$k_l$” employing the pseudo first order reaction model for the rate of absorption. He used slotted packing and oblate spheroid packing in addition to plain spherical packing. The interfacial area per unit static bed volume “$a$” and per unit expanded bed volume “$a_e$” was higher for slotted packing than for oblate spheroids and plain spherical packing. However, the “$a$” and “$a_e$” were higher for plain spherical packing than for oblate spheroids. The
interfacial area per unit volume of liquid holdup \(a_l\) for plain spherical packing was higher than for slotted spheres and oblate spheroids packing. Slotted packings offered high volumetric mass transfer coefficients than for plain spheres and oblate spheroids.

Hekmat-Nazemi (1992) introduced a new parameter termed as operational mass transfer efficiency which was defined as the ratio of the volumetric liquid film mass transfer coefficient to the fluid energy consumption in the bed, i.e.

\[
\eta_{op} = \frac{k_{la} H_e}{\Delta p u_g}
\]  

(3.10)

Hekmat-Nazemi (1992) found that the operational mass transfer efficiency for the oblate spheroids packing was higher than for both 25 mm plain and 25 mm slotted spherical packings, particularly at the highest superficial gas velocities. This efficiency was also higher for slotted packing than for the plain packing. The operational mass transfer efficiency for all the packing increased with increase in both gas and liquid velocities.

Hekmat-Nazemi (1992) also determined overall volumetric mass transfer coefficient \(K_{ga}\) by measuring rate of absorption and average CO\(_2\) concentration across the bed.

The parameters \(K_{ga}, k_{la}, k_b, \) and “a” measured by Hekmat-Nazemi (1992) were 0.01-0.044 kmoles/(m\(^3\)-s-atm), 0.11-0.48 s\(^{-1}\), 6×10\(^{-4}\)-14×10\(^{-4}\) m/s and 2900-6900 m\(^{-1}\) respectively.

Inayat (1995) used the same column as was used by Hekmat-Nazemi (1992) and determined volumetric liquid film transfer coefficients \(k_{la}\) by air oxidation of Na\(_2\)SO\(_3\). In his experiments, CoSO\(_4\) was used as a catalyst in such a quantity that the reaction was fast enough in the bulk of the liquid for the oxygen concentration to be effectively zero whilst the reaction was slow enough for the oxygen not to react significantly in the film. To achieve such conditions, concentration of CoSO\(_4\) used was 2×10\(^{-7}\) M and that of Na\(_2\)SO\(_3\) was 0.8M. Concentration of Na\(_2\)SO\(_3\) was not allowed to fall below 0.4M because below certain concentrations (called critical concentration) the rate of absorption becomes dependent on concentration of Na\(_2\)SO\(_3\) (Linek 1981). Inayat (1995) determined mass transfer in the plenum, the lower part of the column and subtracted it from the total mass transfer in the column to evaluate the mass transfer coefficient \(k_{la}\) in the bed only. An increase in \(k_{la}\) was observed with both gas and liquid flow rates for all types of packing and in most cases
it passes through a maximum at some combination of gas and liquid flow rates. It was found that at low superficial gas velocity, 25 mm plain spheres offered much higher values of $k_a$ than for other packings used in his study for lower liquid flow rates. However, for higher liquid flow rates, oblate ellipsoids gave higher values because of more random motion than the spherical packing. But for higher gas flow rates, this increased random motion resulted in slugging in the bed which lowered the mass transfer efficiency of oblate ellipsoids. This slugging problem at higher gas velocities was overcome by using slotted oblate ellipsoids although slotted oblate ellipsoids did not perform well as compared to oblate ellipsoids and 25 mm plain spheres for $u_g<2.7$ m/s and $u_l>0.00657$ m/s.

Little effect of superficial liquid velocity was observed on volumetric liquid film transfer coefficient in the plenum ($k_{la})_p$ based on plenum height. For all the superficial liquid velocities, ($k_{la})_p$ increases slowly from $u_g=1.8-2.1$ m/s and then becomes almost constant up to $u_g=2.7$ m/s. After $u_g=2.7$ m/s, ($k_{la})_p$ increases sharply.

As much as 80% mass transfer was observed in the plenum section at higher liquid and gas flow rates. The values of mass transfer coefficients in the bed and plenum studied by Inayat (1995) were 0.025-0.33 s$^{-1}$ and 0.05-0.33 s$^{-1}$. Static bed height for all the experiments was 10.5 cm and height of the plenum section was 65 cm.

Shabani et al. (2010) investigated interfacial area “a” and overall volumetric mass transfer coefficient by absorption of CO$_2$ from air into Na$_2$CO$_3$ solution in water." They applied the same technique for determining interfacial area as was applied by Wozniak and Ostergaard (1973). They compared their data of TCA with the data of packed column from Ph.D thesis of Shahraki (1998) and produced Table 3-1. They concluded that TCA is five times more efficient than packed column. However, they did not mention about the effect of mass transfer in gas distribution section of TCA. The interfacial area in their study is 250-3200 m$^2$/m$^3$.

Table 3-1: Comparison between TCA and packed columns

<table>
<thead>
<tr>
<th>Turbulent Contact Absorber</th>
<th>Packed Column</th>
</tr>
</thead>
<tbody>
<tr>
<td>NaOH (Normality)</td>
<td>a (m$^{-1}$)</td>
</tr>
<tr>
<td>0.1</td>
<td>1306.27</td>
</tr>
<tr>
<td>0.5</td>
<td>1157.38</td>
</tr>
</tbody>
</table>
### 3.3 Volumetric gas film transfer coefficients

Different chemical methods have been adopted in literature for measuring the gas film transfer coefficients in various mass transfer equipments. For the determination of volumetric gas phase transfer coefficients, gas absorption is accompanied with instantaneous chemical reaction, so that the liquid side mass transfer resistance is negligible. In gas phase controlled mass transfer processes, concentration of the reagent in the gas phase is as low as possible and that in the liquid phase is higher. So processes such as absorption of SO$_2$ in NaOH, absorption of NH$_3$ in aqueous solution of strong acids and evaporation of water have been used for determination of the gas phase transfer coefficients in Turbulent Contact Absorbers (Douglas et al. 1963; Douglas 1964; El-Dessouky 1993). Gas side controlled mass transfer coefficient in packed towers has also been measured using sublimation of naphthalene from packings covered with it (H.L.Shulman et al. 1955), evaporation of liquid in a gas stream (R.G.Taecker and O.A.Hougen 1949) and absorption of water from humid air by calcium chloride solution (H.D.Doan and M.E.Fayed 2000).

Douglas et al. (1963; 1964) were the first to determine mass transfer coefficients in Turbulent Contact Absorbers. They observed an increase in $K_{g,a}$ with increase in liquid velocity and decrease in $K_{g,a}$ with increase in gas velocity for ammonia absorption. They also observed a decrease in $K_{g,a}$ with increasing static bed height $H_0$. The effect of gas velocity on $K_{g,a}$ for dehumidification and cooling was the same as for absorption of ammonia. The overall mass transfer coefficient $K_{g,a}$ decreased with increasing liquid velocity for the case of dehumidification and cooling while in ammonia absorption case the reverse was true. The overall volumetric mass transfer coefficient $K_{g,a}$ and the heights of overall gas phase transfer unit
Hog determined by Douglas (1964) by absorption of NH₃ from air in boric acid solution in water were 0.21-0.56 kmoles/(m³-s-atm) and 0.11-0.41 m respectively. By dehumidification and cooling of hot air by water, Douglas (1964) obtained the values of Hog in the range 0.06-0.17 m.

Kosev and Eelenkov (1973) used evaporation of water to determine gas phase transfer coefficient. The gas phase inlet temperature was varied from 6 to 20 °C. It was maintained constant for a single experiment. They studied the effect of liquid and gas flow rate, packing size, packing density, static bed height and grid free area on kgs. They measured the water vapor concentration in the gas phase psychrometrically and used the following equation

\[
k_{gs} = u_g \left[ \frac{Y_{ex} - Y_{in}}{\Delta Y_m} \right] 
\]  

(3.11)

Where,

\( k_{gs} \) = mass transfer coefficient in the gas phase referred to unit of column section, m/s.

\( Y_{ex} \) = water concentration in air at exit, kg/kg

\( Y_{in} \) = water concentration in air at inlet, kg/kg

\( \Delta Y_m \) = mean water concentration in gas phase in column, kg/kg.

They found an increase in kgs with increase in gas velocity, liquid velocity, static bed height and density of packing. For heavier packing, the dependence of kgs on gas velocity was parabolic while for lighter packing the dependence was linear. A decrease in kgs was observed with increasing grid free area. They showed that for heavier packing, kgs increased with increasing gas velocity up to about 4m/s then a decreasing trend of kgs was observed with increasing gas velocity. Kosev and Eelenkov (1973) proposed the following correlation for determination of kgs with maximum error of ±50% and mean error of ±8%.

\[
Nt_g = 0.11 \text{Re}_g (\text{Pr}_g)^{0.5} (\text{Ga}_g)^{0.1} 
\]  

(3.12)

Where,

\[
Nt_g = \frac{k_{gs} h_0}{D_g} \text{= Nusselt diffusion number of the gas}
\]
Re_g = \frac{u \cdot h_0}{v_g} = \text{Reynolds number for the gas.}

Pr_g = \frac{v_g}{D_g} = \text{Prandtl diffusion number for the gas}

Ga_g = \frac{gh_0^3}{v_g^2} = \text{Galilei coefficient for the gas}

h_0 = \frac{\Delta p - \Delta p_0}{\rho_l} = \text{determining linear dimension, meters}

\nu_g = \text{kinematic viscosity of the gas, m}^2/\text{s}

D_g = \text{molecular diffusion coefficient of water vapor in air, m}^2/\text{s}

Wozniak and Ostergaard (1973) measured gas film mass transfer coefficient as mentioned in section 3.2. Gas film mass transfer coefficient \( k_g \) was \( 1 \times 10^{-4} \) to \( 3.5 \times 10^{-4} \) kmoles/(m\(^2\) -s-atm). It decreased with increase in liquid flow rate.

Barile et al. (1975) studied the cooling performance of a Turbulent Contact Absorber. They used the following equation to determine cooling performance of the column.

\[ \int_{t_1}^{t_2} dt_j \left( i_j - i_g \right) = \frac{KaH_0}{L} \]  
(3.13)

Where

\( t_1 \) and \( t_2 \) = Inlet and outlet water temperatures, respectively.

\( i_j \) and \( i_g \) = Enthalpy of saturated air at the bulk water temperature and the enthalpy of air at the bulk water temperature, kJ/kg dry air

\( Ka \) = Overall mass transfer coefficient, kg/(m\(^3\) s)

\( H_0 \) = Packing depth, m

\( L \) = Liquid flux, kg/(m\(^2\) s)

The term \( \frac{KaH_0}{L} \) is called tower characteristic and is used to specify tower performance. Tower characteristic reflects tower performance over a wide variety of possible conditions. Tower characteristic is calculated by numerical integration of the left side of eq.(3.13).
Barile et al. (1975) developed the following correlation for their data.

\[
\frac{KaH_o}{L} = 0.0819 \text{Re}^{0.025} Fr_l^{-0.338} \left( \frac{L}{G} \right)^{-0.248} \left( \frac{H_o}{d_p} \right)^{0.309}
\]  

(3.14)

Where, \( Fr_{r,l} = \frac{L}{\sqrt{d_p g \rho_l^2}} \)

The data fit had an average error of 9.8% in prediction of tower characteristic.

Barile et al. (1975) observed an increase in Ka with increasing gas and liquid velocities and decrease in Ka with increase in static bed height.

El-Dessouky (1993) also solved eq. (3.13) numerically to evaluate the tower characteristic at different operating conditions of his experimental data. The following two equations were used in the numerical integrations:

\[
i_l = 21089.12 - 143.055T + 0.2435T^2
\]  

(3.15)

\[
i_g = i_o + \left( \frac{LC_{p_v}}{G} \right)(T - T_{w_o})
\]  

(3.16)

Where

\[
i_o = (T_o - 273.15)(C_a + WC_{p_v}) + W \lambda
\]

\[i_o\] = Enthalpy of air-water vapor mixture at water outlet temperature.

\[T_o\] = Cold water outlet temperature, K.

\[C_a\] = Air-water vapor mixture specific heat at constant pressure, kJ/(kg K)

\[C_{p_v}\] = Water vapor specific heat at constant pressure, kJ/(kg K)

\[W\] = Humidity of air, kg water vapor/kg dry air

\[\lambda\] = Latent heat of vaporization, kJ/kg

The results of experimental data by El-Dessouky (1993) show that for all packing heights, the tower characteristic decreases with the increase in L/G. By increasing L/G, its effect on the tower characteristic becomes less pronounced. Increasing the packing height increases the tower characteristic. However, at large packing heights and high air velocities, the rate of increase in tower characteristic with packing height is relatively low. Tower characteristic is strongly dependent on water inlet temperature. Higher the temperature, greater is the tower characteristic. El-Dessouky (1993) believes that the increase in tower characteristic with increase in inlet water
temperature is due to decrease in surface tension and viscosity of liquid. This decrease in surface tension and viscosity will cause a decrease in bubble size which will increase the contact area between the water and air per unit mass of air. He developed the following relation from his data.

\[
\frac{K_a V}{L} = -2.297 - 0.152 \frac{L}{G} + 1.47Z + 0.0086T_i
\]

(3.17)

Where, \(T_i\) is hot water inlet temperature.

The range of tower characteristic studied by El-Dessouky (1993) is from 0.6 to 1.2.

Inayat (1995) determined gas film mass transfer coefficient by adiabatic humidification. This allowed the temperature measurements to be used directly to determine the volumetric gas film mass transfer coefficient. Inayat (1995) determined gas film mass transfer coefficient \((k_{ga})_{c,e}\) in the column based on the expanded bed height plus height of the plenum and \((k_{ga})_{i,n}\) in the bed based on static bed height of the packing. He observed an increase in \((k_{ga})_{c,e}\) with increasing \(u_g\) for all the packings over the entire range of \(u_l\) investigated, however for slotted oblate ellipsoids, \((k_{ga})_{c,e}\) seemed to pass through a maximum for \(u_l \geq 0.00657\) m/s. This maximum occurred at \(u_g = 3.0\) m/s. With increasing \(u_l\), \((k_{ga})_{c,e}\) increased slightly for 25 and 45 mm plain spheres for \(u_g \leq 2.4\) m/s. For \(u_g \geq 2.4\) m/s, \((k_{ga})_{c,e}\) for these two packing virtually remained constant. \((k_{ga})_s\) increased with both increasing gas and liquid flow rates. For \(u_g = 1.8\) to 2.1 m/s, \((k_{ga})_s\) remained almost constant for all the packings. For slotted oblate ellipsoids this effect was only for \(u_l < 0.00657\) m/s and for 25 mm plain spheres this effect went up to \(u_g = 2.7\) m/s. For slotted oblate ellipsoids and 45 mm spheres, it seemed that \((k_{ga})_s\) passed through a maximum close to \(u_g = 3\) m/s for almost the entire range of liquid flow rates studied. There was no significant effect of packing shape for same size of packing but a significant effect of packing size on \((k_{ga})_s\) was observed. Inayat (1995) determined the \((k_{ga})_{c,e}\) values in the range 0.32-1.15 kmole/(m\(^3\).s.atm) and \((k_{ga})_s\) values in the range 0.13-2.20 kmole/(m\(^3\).s.atm).

Wen and Chang (1978) scrubbed SO\(_2\) with lime and developed following correlation for gas side mass transfer coefficient.

For spray section of TCA
\[ k_g a = 0.08455G^{0.8}L^{0.4} \]  \hspace{1cm} (3.18)

For packed section of TCA

\[ k_g a = 0.03959G^{0.9}L^{1.4} \]  \hspace{1cm} (3.19)

Where, L and G are molar flow rates of gas and liquid in k-moles/(m\(^2\)-s) and \(k_g a\) is
volumetric gas film mass transfer coefficient in k-moles/(m\(^3\)-s-atm)

The parameters for mass transfer study by different investigators are mentioned in Table 3-2. The authors who excluded the mass transferred in the plenum are labeled with * in this table.
<table>
<thead>
<tr>
<th>Investigators</th>
<th>f</th>
<th>(H_0) ((\text{cm}))</th>
<th>(u_l) ((\text{m/s}))</th>
<th>(u_g) ((\text{m/s}))</th>
<th>(D_c) ((\text{cm}))</th>
<th>(d_p) ((\text{mm}))</th>
<th>(\rho_p) ((\text{kg/m}^3))</th>
<th>Systems used</th>
<th>Measured parameters</th>
</tr>
</thead>
<tbody>
<tr>
<td>Shabani et al. (2010)</td>
<td>N.A</td>
<td>20</td>
<td>0.002-0.005</td>
<td>0-3</td>
<td>10</td>
<td>15</td>
<td>335</td>
<td>CO(_2)-air-NaOH</td>
<td>a,K(_g)a</td>
</tr>
<tr>
<td>Zahedi et al. (2006)</td>
<td>0.70</td>
<td>30</td>
<td>0.0094-0.014</td>
<td>0-3</td>
<td>15</td>
<td>15</td>
<td>250</td>
<td>CO(_2)-K(_2)CO(_3)</td>
<td>CO(_2) removal</td>
</tr>
<tr>
<td>Shackley (2000)*</td>
<td>0.41-0.96</td>
<td>3.8-21</td>
<td>0-0.0167</td>
<td>0.9-3.7</td>
<td>46</td>
<td>25-45 &amp; ellipsoids</td>
<td>168-640</td>
<td>O(_2)-air-water</td>
<td>k(_a)</td>
</tr>
<tr>
<td>Guerriere et al. (1995)*</td>
<td>0.79</td>
<td>7.2-36.2</td>
<td>0.0017-0.011</td>
<td>1-5</td>
<td>29</td>
<td>20</td>
<td>315</td>
<td>Air-water-CaCl(_2)</td>
<td>K(_g)a</td>
</tr>
<tr>
<td>Inayat (1995)*</td>
<td>0.72</td>
<td>10.5</td>
<td>0.00232-0.00877</td>
<td>1.8-3.0</td>
<td>22</td>
<td>25, 38, 45, 38×50 oblate spheroids</td>
<td>128-327</td>
<td>Air-water, O(_2)-Na(_2)SO(_3)</td>
<td>k(_g)a, k(_a)</td>
</tr>
<tr>
<td>El-Dessouky (1993)</td>
<td>0.708</td>
<td>30-50</td>
<td>0.0013-0.008</td>
<td>0.34-2.65</td>
<td>20</td>
<td>12.7</td>
<td>375</td>
<td>Air-water</td>
<td>K(_a)V(_c)/L.</td>
</tr>
<tr>
<td>Hekmat-Nazemi (1992)</td>
<td>0.72</td>
<td>10.5-16.5</td>
<td>0-0.0088</td>
<td>0-4</td>
<td>22</td>
<td>25, 38, 38×50 oblate spheroids</td>
<td>161-327</td>
<td>CO(_2)-air-NaOH</td>
<td>K(_g)a, K(_a), a</td>
</tr>
<tr>
<td>Palaty (1989)</td>
<td>0.41-0.655</td>
<td>21-47</td>
<td>0.00536-0.0125</td>
<td>1-3</td>
<td>5.8, 10.9</td>
<td>6,10</td>
<td>420-1000</td>
<td>CO(_2)-air-NaOH-Na(_2)CO(_3)</td>
<td>a, k(_l)</td>
</tr>
<tr>
<td>Tabei K. et al. (1987)</td>
<td>0.817</td>
<td>5-20</td>
<td>0.001-0.025</td>
<td>1.4-2.6</td>
<td>10.9</td>
<td>19.5</td>
<td></td>
<td>H(_2)Cl-air-water</td>
<td>Efficiency</td>
</tr>
<tr>
<td>Miconnet et al. (1981)*</td>
<td>0.82</td>
<td>10-30</td>
<td>0.001-0.0075</td>
<td>0-6</td>
<td>30</td>
<td>20-38</td>
<td>86.6-806</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Paterson (1979)*</td>
<td>0.405-0.615</td>
<td>15-58</td>
<td>0.0056-0.0222</td>
<td>1-4</td>
<td>45.7</td>
<td>24.7-37</td>
<td>Not given</td>
<td>CO(_2)-air-water, O(_2)-CoSO(_4)</td>
<td>K(_g)a, K(_g)a or k(_a), K(_a)</td>
</tr>
<tr>
<td>Wozniak (1977)</td>
<td>0.6</td>
<td>20-40</td>
<td>0.005-0.027</td>
<td>1.7-3.1</td>
<td>20</td>
<td>19,38</td>
<td>266</td>
<td>CO(_2)-air-NaOH</td>
<td>a</td>
</tr>
<tr>
<td>Zhukov (1977)</td>
<td>Not reported</td>
<td>0-25</td>
<td>0.0039-0.0139</td>
<td>3.1</td>
<td>Not reported</td>
<td>Not reported</td>
<td></td>
<td></td>
<td>K(_g)a</td>
</tr>
<tr>
<td>Strumillo and Kudra (1977)</td>
<td>0.65</td>
<td>2-16</td>
<td>0.009-0.0306</td>
<td>0.5-3.5</td>
<td>8.5</td>
<td>5-10</td>
<td>1050</td>
<td>CO(_2)-air-NaOH</td>
<td>a</td>
</tr>
<tr>
<td>Barile et al. (1975)</td>
<td>0.82</td>
<td>0-46</td>
<td>0.00247-0.0117</td>
<td>1.5-4.5</td>
<td>28.6</td>
<td>19-38</td>
<td>160,109</td>
<td>air-water</td>
<td>K(_a)V(_c)/L.</td>
</tr>
<tr>
<td>Investigators</td>
<td>f</td>
<td>H₀ (cm)</td>
<td>uᵢ (m/s)</td>
<td>uₚ (m/s)</td>
<td>Dₑ (cm)</td>
<td>dᵢ (mm)</td>
<td>ρₚ (kg/m³)</td>
<td>Systems used</td>
<td>Measured parameters</td>
</tr>
<tr>
<td>------------------------------------</td>
<td>------------</td>
<td>---------</td>
<td>-----------</td>
<td>-----------</td>
<td>---------</td>
<td>---------</td>
<td>------------</td>
<td>-----------------------------------</td>
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</tr>
<tr>
<td>Wozniak and Ostergaard(1973)</td>
<td>Not reported</td>
<td>22</td>
<td>0.02-0.07</td>
<td>0.7</td>
<td>10</td>
<td>9.7</td>
<td>388</td>
<td>CO₂-air-NaOH</td>
<td>a, kₔ, Kₔa</td>
</tr>
<tr>
<td>Kosev and Elenkov(1973)</td>
<td>0.417-0.79</td>
<td>8-17</td>
<td>0.00715-0.0236</td>
<td>1.2-6.2</td>
<td>19</td>
<td>17.18</td>
<td>167-1090</td>
<td>air-water</td>
<td>kₔa</td>
</tr>
<tr>
<td>Gel’perin et al.(1972)*</td>
<td>0.19-0.55</td>
<td>4.5-18</td>
<td>0.0056-0.0167</td>
<td>1.5-4</td>
<td>14.5-20</td>
<td>15.5</td>
<td>470</td>
<td>CO₂-air-NaOH/MEA</td>
<td>aₑ</td>
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<tr>
<td>Khanna (1971)</td>
<td>0.7</td>
<td>14</td>
<td>0.0054-0.0325</td>
<td>0-4</td>
<td>14</td>
<td>12.7-38.1</td>
<td>150</td>
<td>Hₜₔmf, εₐ</td>
<td></td>
</tr>
<tr>
<td>Barile and Meyer (1971)</td>
<td>0.82</td>
<td>15-53</td>
<td>0.013-0.038</td>
<td>0.5-5.1</td>
<td>28.6</td>
<td>19.38</td>
<td>109,160</td>
<td>air-water</td>
<td>Hₜₒ</td>
</tr>
<tr>
<td>Elenkov and Kosev(1970)*</td>
<td>0.417-0.79</td>
<td>8-20</td>
<td>0.0056-0.024</td>
<td>2.85-5.54</td>
<td>19</td>
<td>18</td>
<td>167-1090</td>
<td>O₂-air-water</td>
<td>kₐa</td>
</tr>
<tr>
<td>Levsh et al.(1968b)</td>
<td>0.288</td>
<td>7-21</td>
<td>0.0028-0.017</td>
<td>2-5</td>
<td>17.8</td>
<td>Rings 18×9×2, 8×4×0.3</td>
<td>Not reported</td>
<td>O₂-air-water</td>
<td>kₐaₑ, mass transfer efficiency</td>
</tr>
<tr>
<td>Blyakher et al.(1967)</td>
<td>0.41</td>
<td>30</td>
<td>0.0014-0.011</td>
<td>1.4-4.0</td>
<td>35</td>
<td>38</td>
<td>180</td>
<td>NH₃-air-water</td>
<td>Efficiency</td>
</tr>
<tr>
<td>Pollock et al.(1967)</td>
<td>Not reported</td>
<td>25.4-51</td>
<td>0.00106-0.0265</td>
<td>2.79</td>
<td>35.5</td>
<td>38</td>
<td>Efficiency of absorption</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Douglas (1964)*</td>
<td>Not reported</td>
<td>26-84</td>
<td>0.0034-0.02</td>
<td>1.16-2.26</td>
<td>30.5×30.5</td>
<td>38</td>
<td>155</td>
<td>NH₃-air-boric acid soln. in water</td>
<td>Kₐₔa, HTU</td>
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<tr>
<td>Douglas et al.(1963)</td>
<td>Not reported</td>
<td>25.4</td>
<td>0.0034-0.02</td>
<td>1.46-2.26</td>
<td>30.5×30.5</td>
<td>38</td>
<td>155</td>
<td>air-water</td>
<td>HTU</td>
</tr>
<tr>
<td>Douglas et al.(1963)</td>
<td>38.1</td>
<td>0.020-0.034</td>
<td>4.72-4.98</td>
<td>30.5×30.5</td>
<td>38</td>
<td>SO₂-air-NaOH</td>
<td>Kₐₔa</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Douglas et al.(1963)</td>
<td>79.3</td>
<td>0.00204-0.034</td>
<td>3.56</td>
<td>30.5×30.5</td>
<td>38</td>
<td>CO₂-air-NaOH</td>
<td>Kₐₔa</td>
<td></td>
<td></td>
</tr>
</tbody>
</table>

The authors who excluded plenum mass transfer are labeled with *
Chapter 4

4 Experimental Setup and Procedures

4.1 Experimental setup

The schematic diagram of the experimental set up is shown in Figure 4-1. It consists of a 44.7 cm diameter, 213 cm high Perspex column. A gas distribution section (called plenum) of 45 cm diameter and 122 cm height is attached under the Perspex column. The plenum is made of SS-304L. It also serves as air seal and a liquid collector for the determination of liquid holdup.

Different types of gas distributors were tried for proper gas distribution above grid. These were

i. Standard inlet where there was no additional distribution device and the inlet pipe was cut at 45 degrees from bottom at its outlet.

ii. Orifice baffle where direction of major portion of the air was diverted along the walls of the column tangentially while remaining gas was allowed to enter as in standard inlet as described by Wehrli et al. (2003)

iii. Perforated pipe where the inlet pipe was perforated from bottom of its outlet.

iv. Inverted cone type where an elbow was provided at air inlet pipe in centre of the column. The other, upward end of the elbow was provided with inverted cone to change direction of gas from center to whole column and to prevent liquid enter into the inlet pipe through elbow This type of gas distributor was used by Inayat (1995).

When none of the above distributors provided acceptable gas distribution due to the limited plenum height available, a novel type of gas distributor was installed to achieve acceptable gas distribution at the support grid situated between Perspex and SS cylinders. Plenum section contains two vertical plates welded to both sides of the column at a position 90 degree from inlet air nozzle (Figure 4-2). Air inlet pipe in the column contains a T shaped section at its end. This T shaped section is provided with
two adjustable sliding plates at each side perpendicular to the inlet air. Air entering into the column was divided into four paths with the help of these sliding plates in the T shaped section. Horizontal plates were fixed in vertical plates at both outlets of the T shaped section for better air distribution. Required air distribution was achieved by adjusting the gap between the sliding plates.

Liquid was showered from a SS shower connected with the inlet pipe. To prevent the liquid flow along the walls of the column, shower was housed in a 6 inches PVC pipe. This arrangement caused the liquid to flow down in the form of ring thus achieving a reasonable liquid distribution.

Support grid was a 16 BWG SS sieve having mesh size 2 and a 74% free area. To remove the moisture entrained by the gas flowing upwards, a mist eliminator was installed at the top most section of the tower. Air flow rates to the column were measured by a calibrated orifice meter installed in the inlet gas pipe whereas the liquid flow rate was measured by a calibrated rotameter. Two hand holes were provided in Perspex column one above the support grid and other near the top of the column to change the packing. During the determination of volumetric gas film transfer coefficients, hand hole at the top was provided with drift type mist eliminator (Figure 4-3) to remove moisture from air. Packing used was made of high-density polyethylene (HDPE) hollow balls of various density and diameters.

4.2 Pressure drop measurements

Pressure drop across the column was measured by calibrated pressure transducers (Model 600D1P12D20, Auto Tran Incorporated) having a range of 0-1 psi. For measuring pressure drops along the column one pressure tap was provided in the Perspex column just below the support grid and others were in the column at various locations. For measuring pressure drop across grid, one pressure tap was provided just below the grid, one just above the grid and one at a distance of 3 cm above the grid. Pressure drop across grid was measured by pressure transducer with range 0-2 inches of water (Dwyer USA model 616W-0) having an accuracy of ± 0.5 %. Pressure drop in the plenum was determined from pressure taps one below the grid and other at the air inlet pipe. Arrangements as shown in Figure 4-4 were made to prevent the liquid
entry into the taps. Velocity head in the gas inlet pipe (Figure 4-1) was added to
determine total pressure drop across plenum.

4.3 Liquid holdup measurements

The total liquid held in the tower including bed, plenum and liquid distributor pipe
was measured by collecting the liquid after simultaneously closing liquid inlet and
outlet valves and butterfly valve located at the discharge side of the blower. The rise
in level of water collected in the plenum was observed from the level glass. To
measure the true liquid holdup in the fluidized bed only, a correction for the liquid
volume held in the liquid distributor pipe as well as in the plenum was needed. This
was achieved by measuring the liquid volume held in the empty column including
liquid distributor and its pipe at all the liquid and gas velocities studied by shut off
valves technique. To measure the liquid collected in the column i.e. bed plus plenum,
liquid in the liquid distributor along with its pipe was needed. This was measured by
removing the liquid distributor outside of the column and measuring the volume
collected after closing pump discharge valve at each liquid flow rate. This amount of
liquid was subtracted from that collected in case of column with bed or empty column
to get the holdup in column or plenum respectively. Range of the variables at which
liquid holdup was measured are given in Table 4-1 and specification of packing
in Table 4-2.

Table 4-1: Range of Variables studied for liquid holdup

<table>
<thead>
<tr>
<th>Variables</th>
<th>Ranges</th>
</tr>
</thead>
<tbody>
<tr>
<td>Diameter of column $D_c$ (m)</td>
<td>0.447</td>
</tr>
<tr>
<td>Superficial liquid velocity $u_l$ (m/s)</td>
<td>0.0-0.014</td>
</tr>
<tr>
<td>Superficial gas velocity $u_g$ (m/s)</td>
<td>1.8-3.6</td>
</tr>
<tr>
<td>Static bed height $H_o$ (m)</td>
<td>0.15, 0.25, 0.35, 0.45</td>
</tr>
<tr>
<td>Fractional free area of support $f$</td>
<td>0.74</td>
</tr>
</tbody>
</table>
Table 4-2: Specifications of the packings used

<table>
<thead>
<tr>
<th>Packing</th>
<th>Packing dia (mm)</th>
<th>Packing weight (grams)</th>
<th>Packing density (kg/m³)</th>
</tr>
</thead>
<tbody>
<tr>
<td>P1</td>
<td>38</td>
<td>5.17</td>
<td>180</td>
</tr>
<tr>
<td>P2</td>
<td>38</td>
<td>7.75</td>
<td>270</td>
</tr>
<tr>
<td>P3</td>
<td>38</td>
<td>10.20</td>
<td>354</td>
</tr>
<tr>
<td>P4</td>
<td>38</td>
<td>12.70</td>
<td>442</td>
</tr>
<tr>
<td>P5</td>
<td>38</td>
<td>1.74</td>
<td>547</td>
</tr>
<tr>
<td>P6</td>
<td>25</td>
<td>2.90</td>
<td>354</td>
</tr>
<tr>
<td>P7</td>
<td>45</td>
<td>16.90</td>
<td>354</td>
</tr>
</tbody>
</table>

4.4 Volumetric gas film transfer coefficient

Volumetric gas film transfer coefficients were studied for liquid velocities 0.004-0.012 m/s and gas velocities 0.45-3.6 m/s with the same other parameters as mentioned in table 4.1 and table 4.2. For the determination of volumetric gas film transfer coefficients, the whole column was wrapped with 18 mm thick polyurethane insulation having a thermal conductivity 0.02 W/m-K. The TCA was operated as an adiabatic humidification tower. Water was re-circulated through the tower while air flowed continuously through the tower in a once through mode. Steady state water and air temperature at the inlet and outlet of the TCA bed were measured. The inlet and outlet air and water temperatures were determined with thermistors having an accuracy of ± 0.15 ⁰C and a least count 0.1 ⁰C. Mercury thermometers with range -5 to 50 ⁰C and a least count 0.1 ⁰C were also installed for determination of temperatures to check the reliability of thermistors. Air outlet temperature was measured at the exit of an insulated drift mist eliminator as shown in Figure 4-3.

Photograph of the setup during experiments for determination of $k_{ga}$ are given in Figure 4-6.
Figure 4-1: Experimental setup

1- Fan
2- Butterfly valve
3- Gas inlet pipe
4- Pressure tap
5- Orifice plate
6- Temperature indicator
7- Mixing tank
8- Ball valve
9- Pump
10- Liquid inlet pipe
11- Liquid outlet pipe
12- Level pipe
13- Globe valve
14- SS plenum
15- Air distributor
16- Support grid
17- Hand hole
18- Rotameter
19- Packing
20- Perspex column
21- SS shower
22- PVC pipe
23- Drift mist eliminator
24- Baffle
25- Flexible pipe to mixing tank
26- SS sieves
27- Mist eliminator
28- Nipple

Figure 4-2: Gas distributor

29- Plenum wall
30- Vertical plate
31- Hole for bolt
32- Slot
33- Sliding plate
34- Inlet air pipe
35- Tee
36- Horizontal plate
Figure 4-3: Mist eliminator

Air and moisture from TCA column

37-Baffle plate
38-Perspex flange
39-Perspex pipe
40-115.3mmx140mm box
41-Drain pipe

Figure 4-4: Pressure drop measurement arrangement

42-Flexible tube
43-Bellows
44-Pressure tap
45-Pressure transducer
Figure 4-5: Photo of experimental set up

Figure 4-6: Photo of experimental setup with insulation
Chapter 5

5 Mathematical Modeling

5.1 Introduction

Liquid and gas film transfer coefficients are determined to predict the mass transfer characteristics of the equipment. A number of mass transfer theories are available for the prediction of gas absorption in the gas absorption equipments. The most common is the two film theory. There are two types of systems for determination of mass transfer coefficients i.e. by physical absorption and by chemical absorption. These have been mentioned in chapter 3. The theoretical detail of these systems have been given by Astarita (1967), Danckwerts (1970) and Sherwood et al. (1975). Theoretical detail related to air-water system is given here. In the two film theory, all the mass transfer resistance lies in the liquid and gas films adjacent to gas-liquid interface as shown in Figure 5-1.

Figure 5-1: Liquid and gas films for mass transfer

Mass transfer through the films can be given as follows.

$$Ra = k_g a (P - P_f) = E k_c a (C^* - C^0)$$  \hspace{1cm} (5.1)

Where,

R = Molar rate of diffusion of diffusing component per unit area

P = Partial pressure of diffusing component in the bulk of the gas phase
P_i = Partial pressure of diffusing component at the interface.
C^* = Molar concentration at interface
C_0 = Molar concentration in bulk of liquid.
k_g = Gas film transfer coefficient
k_l = Liquid film transfer coefficient
a = Interfacial area per unit volume
k_{la} = Volumetric liquid film transfer coefficient
k_{ga} = Volumetric gas film transfer coefficient
E = Enhancement factor which is ratio of absorption of gas with chemical reaction to that with physical absorption only.

5.2 Humidification and dehumidification in TCA

Consider a counter current flow TCA cooling tower as shown in Figure 5-2. A relationship for determination of volumetric gas film mass transfer coefficient can be derived as shown by Norman (1961). Air enters through the bottom of the column with humidity H_{G1}, temperature T_{G1}, enthalpy i_{G1} and leaves at the top of the column with humidity H_{G2}, temperature T_{G2}, and enthalpy i_{G2}. Liquid enters the top of the column with mass velocity L_1, Temperature T_{L1} and leaves from the bottom of the column with mass velocity L_2 and temperature T_{L2}. G' is the mass velocity of air.
based on vapor free air. At a distance H from the bottom of the tower, temperature and humidity of air are $T_G$ and $H_G$ respectively whereas the temperature of water is $T_L$. The temperature and humidity at the gas liquid interface are $T_i$ and $H_i$ respectively. In cooling towers, the temperature of air is normally less than the temperature of water to be cooled. In such cases, the surface temperature $T_i$ is greater than the gas temperature. Cooling is possible even if the air temperature is greater than the water temperature as long as the adiabatic saturation temperature of the air is lower than the desired water temperature. However, interface temperature $T_i$ will always be lower than the liquid bulk temperature, otherwise cooling is not possible. For a short element of tower $dH$, neglecting the change in liquid flow rate, the enthalpy balance gives

$$G'd_{i_G} = Lc_LT_i$$  \hspace{1cm} (5.2)

Where $d_{i_G}$ is the change in the enthalpy of air along with vapors and $dT_L$ is the change in the temperature of liquid respectively. Overall enthalpy balance for the column is

$$G'(i_{G2} - i_{G1}) = Lc_L(T_{L1} - T_{L2})$$  \hspace{1cm} (5.3)

Above equation can be written for any point in the column as follows

$$G'(i_{G2} - i_G) = Lc_L(T_{L1} - T_L)$$  \hspace{1cm} (5.4)

Equation (5.4) is the equation of operating line of the column. It can be written as

$$i_G = \frac{Lc_L}{G'}T_L + \frac{i_{G2}}{G'}T_{L1}$$  \hspace{1cm} (5.5)

Thus the operating line is a straight line having a slope of $Lc_L/G'$ on the plot of enthalpy of air versus temperature of the liquid. Equation of the equilibrium line is

$$i_{G,sat} = c_s(T_G - T_0) + \lambda_0 H_s$$  \hspace{1cm} (5.6)

Where,

$i_{G,sat}$ = Enthalpy of saturated air, J/kg
$T_0$ = Reference temperature, $0^\circ$C
$\lambda_0$ = Latent heat of vaporization at temperature $T_0$, J/kg
$H_s$ = Humidity of saturated air, kg/kg
The rate of sensible heat transfer from water to the interface in the section of bed height $dH$ is

$$Lc\,dT_L = h_a\,c_e\,(T_L - T_i)\,dH$$  \hspace{1cm} (5.7)

Where

$h_a\,c_e =$ liquid side heat transfer coefficient, W/m$^3$-K

The rate of sensible heat transfer from the interface to the air in the section of bed height $dH$, is

$$G'c\,dT_G = h_a\,(T_i - T_G)\,dH$$  \hspace{1cm} (5.8)

Where,

$e =$ humid heat, J/kg

$h =$ heat transfer coefficient, W/m$^2$-K

$a_e = $ interfacial area based on expanded bed volume, m$^2$/m$^3$

Assuming that the air is very dilute, the rate of evaporation of water vapor through the gas film is

$$G'dy = k'a_e\,(y_i - y)\,dH$$  \hspace{1cm} (5.9)

Assuming that the humidity of air is very low, $y$ becomes proportional to $H$. So equation (5.9) can be written as

$$G'd\,H = k'a_e\,(\tilde{H} - \tilde{H})\,dH$$  \hspace{1cm} (5.10)

where

$k'a_e =$ volumetric mass transfer coefficient kg/(m$^3$-s)

By multiplying equation (5.10) by $\lambda_0$, we can get the rate of energy transfer i.e.

$$G'\lambda_0\,d\tilde{H} = k'a_e\lambda_0\,(\tilde{H} - \tilde{H})\,dH$$  \hspace{1cm} (5.11)

Adding equations (5.8) and (5.11), we have

$$G'(\lambda_0\,d\tilde{H} + c_e\,dT_G) = [k'a_e\lambda_0\,(\tilde{H} - \tilde{H}) + h_a\,(T_i - T_G)]\,dH$$  \hspace{1cm} (5.12)

For the air water system, by Lewis relation, $h = c_c\,k'$. Substituting value of hinto equation(5.12), we get

$$G'(\lambda_0\,d\tilde{H} + c_e\,dT_G) = k'a_e[\lambda_0\,(\tilde{H} - \tilde{H}) + c_c\,(T_i - T_G)]\,dH$$  \hspace{1cm} (5.13)

Since
\[ \text{di}_G = \lambda_e \text{dT} + c_d \text{dT}_G \]  

(5.14)

So equation (5.13) can be written as

\[ G' \text{di}_G = k'a_e (i_i - i_G) \text{dH} \]  

(5.15)

Above equation can be integrated to get

\[ \int_{i_{i0}}^{i_{i'}} \frac{\text{di}_G}{i_i - i_G} = \frac{k'a_e H_s}{G'} = N_G \]  

(5.16)

where

\( N_G = \text{number of gas film transfer unit based on enthalpy driving force.} \)

Equation (5.15) is based on the volumetric gas film transfer coefficient. It can be modified for the overall volumetric mass transfer coefficient as follows

\[ G' \text{di}_G = K'a_e (i_{G*} - i_G) \text{dH} \]  

(5.17)

where

\( K'a_e = \text{overall volumetric mass transfer coefficient,} \)

\[ \frac{1}{K'a_e} = \frac{1}{k'a_e M_B} + \frac{m}{h_a a_e} = \text{Reciprocal of overall volumetric mass transfer coefficient.} \]

\[ m = \frac{\text{di}_{G*}}{dT} = \text{Slope of the equilibrium line} \]

\( i_{G*} = \text{Enthalpy of gas in equilibrium with liquid at temperature} \ T_L \).

Integrating equation (5.17) gives

\[ \int_{i_{G*}}^{i_{i'}} \frac{\text{di}_G}{i_{G*} - i_G} = N_{OG} = \frac{H}{H_{OG}} \]  

(5.18)

Where,

\( H_{OG} = \frac{G'}{K'a_e} = \text{Height of transfer unit} \)

\( N_{OG} = \text{Number of transfer units.} \)

Thus the total active height of the column can be determined by equation (5.18)
5.3 Mathematical modeling for adiabatic humidification

When air is brought in contact with water for long time, air and water attain certain temperature called the adiabatic saturation temperature. For the determination of gas phase transfer coefficient, the column used should be well insulated and the air should be passed through the column in once through mode whereas water should be recirculated. After some time, water attains certain steady state temperature called the adiabatic saturation temperature. Once the steady state has been achieved; the enthalpy of the air does not change. In such a case, the inlet air temperature is always greater than the water bulk temperature and the interface temperature $T_i$ is less than the air temperature.

Heat transfer from the bulk of the gas to the interface is

$$G'c_v dT_g = h_a \left(T_g - T_i\right) dH$$ (5.19)

All the terms in equation (5.19) have already been defined in equation (5.8)

As the water temperature becomes constant at the steady state condition, the interface temperature is same as the bulk temperature of water i.e. the adiabatic saturation temperature. So equation (5.19) can be written as

$$G'c_v dT_g = h_a \left(T_g - T_s\right) dH$$ (5.20)

Where, $T_s$ is the adiabatic saturation temperature of the inlet air.
For air-water system, adiabatic saturation temperature and wet bulb temperatures are the same (Norman 1961).

Integration of equation (5.20) gives

\[
\int_{T_{G1}}^{T_{G2}} \frac{dT_G}{T_G - T_s} = \int_0^H \frac{h_a}{c_v G'} dH
\]

(5.21)

\[
\ln \left[ \frac{(T_{G1} - T_s)}{(T_{G2} - T_s)} \right] = \frac{h_a H_e}{G' c_v} = N_h
\]

(5.22)

Mass transfer from interface to the bulk of the gas as in equation (5.10) is

\[
G'd\mathcal{H} = k'a_e (\bar{\mathcal{H}}_i - \mathcal{H}) dH
\]

(5.23)

Rearranging and integrating the above equation, we get

\[
\ln \left[ \frac{(\bar{\mathcal{H}}_i - \mathcal{H})}{(\bar{\mathcal{H}}_i - \mathcal{H}_i)} \right] = \frac{k'a_e H_e}{G'} = N_G
\]

(5.24)

From Lewis relation

\[
h = c_v k'
\]

(5.25)

Substituting this value of \( h \) in equation (5.22), we get

\[
N_h = \frac{c_v k'a_e H_e}{G' c_v} = \frac{k'a_e H_e}{G'} = N_G
\]

(5.26)

Thus, the number of heat transfer units is the same as the number of gas phase transfer units.

Now

\[
k_g a_e = \frac{k'a_e}{\rho G RT} = \frac{k'a_e}{\rho G RT}
\]

(5.27)

Where,

\( P \)= pressure of the gas, atm.
\( M \)=molecular wt. of the gas.
\( k_g a_e \)= volumetric mass transfer coefficient, kmoles/(m\(^3\)-s-atm)
\( \rho G \)= density of the gas
\( R \)= gas constant

From equations (5.22), (5.26)and (5.27), we obtain the following equations for \( k_g a_e \)
As it is easier to measure accurate temperature than humidity, hence equation (5.29) can be used to determine volumetric gas phase transfer coefficient \( k_{ga} \). With the above general available interfacial area definition, one can now define three different ways of interfacial area as based on static bed volume, expanded bed volume or liquid volume retained in the bed.

\[
k_{ga} = \frac{G'N_G}{H_2\rho_GRT} = \frac{G'}{H_2\rho_GRT} \ln \left[ \frac{(H_1 - H_G)}{(H_2 - H_G)} \right]
\]

(5.28)

\[
k_{ga} = \frac{G'N_h}{H_2\rho_GRT} = \frac{G'}{H_2\rho_GRT} \ln \left[ \frac{(T_{G1} - T_e)}{(T_{G2} - T_e)} \right]
\]

(5.29)
Chapter 6

6 Hydrodynamics Results and Discussion

6.1 Introduction

Pressure drop, Expanded Bed height and liquid holdup are regarded as key hydrodynamic parameters of any mass transfer equipment. Therefore they are supposed to be the key role players in determining and explaining the mass transfer performance of any mass transfer equipment. Generally from the liquid holdup, mass transfer efficiency of the equipment can be predicted while pressure drop and bed expansion can be used to determine the capital and operating cost of the equipment. Moreover, the knowledge of bed expansion is necessary for determining the minimum height required above the support grid in a single bed arrangement and minimum height required between two consecutive beds in a multistage arrangement. Effects of liquid and gas velocities, static bed height, diameter and density of packing on pressure drop, liquid holdup and expanded bed height have been reported in this chapter. Correlations have been developed for the pressure drop, expanded bed height and liquid holdup from the experimental data obtained in this study.

The experimental data reported is presented for varying liquid and gas velocities, static bed heights and packing density and diameters. The trends and magnitudes are then compared with the already published data. Efforts have been made to identify the factors which led to the vast degree of discrepancies found in literature regarding the hydrodynamic performance of a TCA. Lastly the correlations developed from the current study data are applied on the already published data by fixing certain conditions based on the consensus available in literature. This effort thus resulted in making a part of the already available data to be presented by a single correlation and hence consensus can be developed.

6.2 Pressure Drop

Pressure drop data obtained in the current study is presented in Figure 6-1 to Figure 6-10. In the following sections the parametric dependences have been discussed.
6.2.1 The effect of liquid and gas velocities

The pressure drop data for varying liquid and gas velocities is presented in Figure 6-1 to Figure 6-4. One can make the following observations from these graphs.

1. Pressure drop increases with increase in gas and liquid velocities. This rise is pretty sharp for low gas velocities but for higher gas velocities this rise is quite small.

2. This behavior is valid for both type 1 and type 2 fluidization.

3. The effect of liquid velocity on the rise in pressure drop for low gas velocities is different for the two types of fluidizations. For type 1 fluidization liquid velocity seems to have considerable effect on pressure drop in this region. Whereas for type 2 fluidization liquid velocity seems to have little effect in this region.

4. Similarly liquid velocities do have significant effects on pressure drop for both types of fluidizations for high gas velocities. However, again this effect is more noticeable for type 1 as compared to type 2 fluidization.

The above mentioned observations can be explained by considering the observation made during these experiments. In the region where the pressure drop rises sharply, the bed is not fluidized. This is in fact the region where the bed behaves like a packed bed. The difference in behavior for the two types of fluidization in this region is due to the different densities of the packing. The two types of fluidizations are determined on the basis of the packing densities. (O'Neill et al. 1972). As a bed of lighter packing can be fluidized to a higher degree for a given gas velocity as compared to the heavier packing, therefore one can see that the pressure drop gradient through the bed is higher for heavier packing as compared to the lighter packing. Similarly the varying liquid velocities affect more when the lighter packing are fluidized as compared to the heavier packing.

Moreover as the liquid velocity increases, more liquid is retained in the bed for a given gas velocity, hence occupying more volume of the column. This causes an increase in interstitial gas velocity which in turn causes more frictional drag, hence more pressure drop.

This sharp increase in pressure drop with gas velocity behavior should change when the bed becomes fully fluidized. At low gas velocity before fluidization, there is much lower area for gas to flow as there is a minimum free space available between the packing. On the other hand, the increase in gas velocity should lift the packing, causing the bed to expand. Therefore at higher gas flow when the bed is fully fluidized more and more free area should be available. Hence as the bed will be fully fluidized the interstitial gas velocities should be significantly reduced, thus reducing the drag as well as friction across the packings. Hence
one can expect a considerably less steep rise in pressure drop against the same bed. This is exactly what is found experimentally.

From this data one can also point out the minimum gas velocities that would result in a fully fluidized bed. Comparing the pressure drop magnitudes for the two types of fluidizations, one can see that pressure drop for type 1 fluidization is almost 30% less as compared to the pressure drop in type 2 fluidization for the whole region of fully fluidized beds. This in turn would suggest that from pressure drop point of view type 1 fluidization is more desirable. However, only the effect of gas and liquid velocities do not completely control the hydrodynamic characteristics of a TCA. Other factors such as packing density, packing diameter and static bed height also affect the hydrodynamics of TCA and hence are now discussed.

6.2.2 The effect of packing density

The experimental results of the effect of packing density on the pressure drop across the bed of TCA are presented in Figure 6-5. The following observations can be made from these figures.

1. The pressure drop increases with increasing packing density. The same sharp rise in pressure drop is observed initially which ultimately settles into a relatively flat pressure drop curve at gas velocities greater than 1 m/s.
2. For lesser dense packing the rise in pressure drop with increasing gas velocity is relatively gradual. However, as the packing density is increased this rise becomes lesser and lesser gradual, ultimately for packing density greater than 354 kg/m$^3$ there is almost a spontaneous change in the pressure drop characteristic around a gas velocity of 1 m/s.
3. For higher density packing the pressure drop almost remains constant for gas velocities greater than around 1 m/s. However for lesser density packing the pressure drop curve become almost constant for gas velocities greater than 3 m/s.

These observations were very much expected, considering the discussion made earlier. The increase in pressure drop with packing density is primarily due to extra weight of the packing causing less bed expansion hence resulting in more interstitial gas velocity. However, for lower density packings the rise in pressure drop is much more gradual once the bed is fully fluidized as compared to the heavier packing. For a packing density of 180 kg/m$^3$ the pressure drop rise from 1 m/s gas velocity (point where the bed is almost fully fluidized) to 3 m/s is around 30%. Whereas for the same gas velocity range the rise in pressure drop is
almost negligible for packing density greater than or equal to 354 kg/m$^3$. If one draws a graph between packing density and pressure drop for a particular $u_l$ but for a range of $u_g$ from 1 m/s to 3 m/s (assuming that for all packing densities the pressure drop for fully fluidized beds is almost independent of gas velocity), one gets a linear relationship between packing density and pressure drop. Hence any increase in packing density will correspondingly increase the pressure drop according to the slope of this density pressure drop curve. The slight non-linearity at the lower end of the graph is where there is type 1 fluidization, where it has been assumed that the pressure drop is independent of the gas velocity in the fully fluidized beds, as assumption not very much in line with the experimental data.

These observations suggest that for type 1 fluidization, the increase in flow rate from a particular TCA would result in a 30% more head loss as compared to the type 2 fluidization (after the bed is fully fluidized). Hence for type 2 fluidization one can increase the flow rate through a particular TCA bed without any increase in pressure drop. However, it should be kept in mind that the pressure drop for type-2 fluidization is significantly more as compared to the type-1. It is worth mentioning that in most of the industrial applications packings used are in type-2 range. This is due to the fact that the mechanical strength of the type-1 packings is far less due to the thin walls. Hence packing densities less than 300 kg/m$^3$ are avoided in industrial applications as they tend to develop cracks and holes over longer periods of operation, thus get settled down in beds, leading to excessive pressure drops and reducing the mass transfer performance as well.

Even though the pressure drop for type-2 fluidization is more than type-1 but the range of gas flow rate available with constant pressure drop is very large.

Now does the packing diameter have any effect on the pressure drop characteristics’ of TCA and if it has, then what would be the optimum packing size. This is now discussed in the following section.

6.2.3 The effect of packing diameter

The data obtained to see the effect of packing diameter on the pressure drop characteristics of TCA is presented in Figure 6-6 and Figure 6-2 to Figure 6-4. The following inferences can be made from these graphs.

1. A decrease in pressure drop with increasing diameter of packing before the bed is fluidized
2. Once the bed is fully fluidized this trend reverses and pressure drop marginally increases with increasing diameter of packing for all liquid and gas velocities

53
To explain this behavior one has to consider the fact that the number of packings available in particular static bed height is far more for the smaller diameter packing as compared to the larger diameter packing. Hence one can expect the number density of smaller diameter packings will be much larger per unit static bed height or expanded bed height as compared to the larger diameter packing. Hence one can expect that the surface area for contact of gas and liquid with solid balls should increase with decreasing packing diameter. This in turn should increase the pressure drop due to more drag and friction before the fluidization starts. However, as the fluidization starts, the bed expands creating more voids. This expansion in the bed is more for smaller diameter packing as compared to larger diameter packings (Figure 6-24). Therefore, the larger the bed expansion lesser is the pressure drop as more free area is available for the gas to flow. This is the reason why the pressure drop is almost independent of the packing diameter when the bed becomes fully fluidized.

Hence from the TCA design point of view as far as the pressure drop characteristics are concerned one can suggest the pressure drop does not depend on the packing diameter. Therefore the optimum packing diameter should depend on other such as bed expansion and/or mass transfer.

6.2.4 The effect of static bed height

The results of the effect of static bed height are presented in Figure 6-7 to Figure 6-9. The presented results can be summarized as:

1. Pressure drop increases with increasing static bed height for all gas velocities.
2. Pressure drop also increases with increasing liquid velocity for all static bed heights
3. This behavior is valid for both type 1 and type 2 fluidizations
4. For same static bed height, pressure drop is more in type 2 fluidization as compared to type 1 for all gas and liquid velocities
5. The increase in pressure drop for type 1 is more gradual even after bed is fully fluidized as compared to type 2
6. The pressure drop per unit static bed height decreases with increasing static bed height

All these observations can be explained in light of the discussion made above. Pressure drop should be more for more static bed height as the weight of the packing increases. However, when the bed is fully fluidized the pressure drop rise becomes almost negligible for type 2 fluidizations as compared to the type 1 fluidizations. Reasons are same as discussed above.
Hence from these observations the optimum static bed height should be dictated by other hydrodynamic parameters besides the pressure drop performance.

### 6.2.5 The effect of grid

No measurable pressure drop across grid was observed at all liquid and gas velocities studied when there was no packing in the column. To further study the effect of grid on pressure drop, pressure drop across the grid was measured with packing in the column having static bed height 25cm, density 354 kg/m$^3$ and diameter 38 mm. One pressure tap was below the grid and other was 3 cm above the grid. Pressure drop across the grid with this static bed height of 3 cm is shown in Figure 6-10. It increases with increasing liquid velocity up to a certain gas velocity then it becomes almost independent of liquid velocity since above certain gas velocity, 3 cm height above grid becomes fully flooded. Even with this fully flooded grid, maximum pressure drop across this 3 cm high packing is 223 N/m$^2$ which is equivalent to 2.3 cm of water head. Hence one can conclude that the contribution of grid will be more if the grid free area is small. Similarly the pressure drop contribution of constant free area grid would decrease as the liquid velocity is increases. However even for same grid free area the pressure drop contribution increases with increasing gas velocity.

### 6.2.6 Comparison with published results

A comparison of measured pressure drop was made with that calculated from liquid holdup in Figure 6-11 both for type 1 and type 2 modes of operation of TCA. Measured pressure drop was found to be more than that calculated from liquid holdup due to contribution of frictional losses in grid and the bed.

Although, there are number of correlations available in literature for pressure drop in TCA, but one cannot find any consensus in the predicted pressure drop from these correlations. Pressure drop calculated by the correlations of various authors available in the literature, compared with those calculated using the correlations developed in the present study both for type 1 and type 2 operations of TCA are presented in Figure 6-12, Figure 6-13, Figure 6-14 and Figure 6-15. Separate correlations for type 1 and type 2 fluidization were developed in present study and earlier by Vunjak-Novakovic et al.(1987b) and Soundarajan and Krishnaiah(1998a). In all the correlations used to predict the pressure drop in Fig 6-12 to 6-15 only Vunjak et.al,(1987b), Soundarajan and Krishnaiah(1998a) and Paterson and Clift (1987) developed correlations for $\Delta P$ prediction from liquid holdup measurements while all other investigators (Aksel' rod and Yakovenko 1969; Balabekov et al. 1971; Barile and
Meyer 1971; Handle 1976; Uchida et al. 1977; Rama et al. 1983; Lyashuk 2001) developed
correlations from the data of experimental measurements of pressure drop. However only
Paterson and Clift (1987) used a 45.7 cm diameter column and measured liquid holdup by
tracer technique and concluded that the pressure drop calculated by liquid hold up was
overestimated. Other investigators (Aksel'rod and Yakovenko 1969; Barile and Meyer 1971;
Handle 1976; Uchida et al. 1977; Rama et al. 1983; Vunjak-Novakovic et al. 1987b;
Soundarajan and Krishnaiah 1998a) worked with columns having diameters from 14 to 30
cm. Congregation of packing at the walls of smaller diameter columns and maldistribution of
liquid and gas may be the reasons of discrepancies among the pressure drop by different
correlations.

The pressure drop in the plenum depends on the design of the plenum; hence it should
vary for different designs. For acceptable gas distribution, either a relatively long plenum is
required or a good gas distributor is installed. In the present study, it was observed that the
smallest diameter and lightest packing (25 mm diameter) required very good gas distribution
for uniform expansion of the bed. Pressure drop for different liquid and gas velocities was
determined in the plenum and found to be independent of liquid velocity.

An attempt is made to develop a dimensionless correlation from the data obtained in this
study. However, the coefficient and powers in the correlation has been tuned to accommodate
as much as possible the reported data of earlier workers. Since the main objective of this
study was to identify the reasons of discrepancies, therefore efforts were not made to develop
a universal correlation that would accommodate all the parameters and would be applicable
to the experimental conditions of all the earlier workers. Therefore this is suggested as one of
the future recommendations. The correlations developed in the present study are as follows.

For bed Type 1

\[ \Delta p = 210.07 Fr_t^{-0.94} We_t^{-0.40} \hat{Re}_l^{1.92} \hat{Re}_g^{-1.8} \left( \frac{\rho_p}{\rho_l} \right)^{0.94} \left( \frac{H_0}{D_c} \right)^{0.85} \rho_g u_g^2 \]  

\[ (6.1) \]

MRD=7%

For bed Type 2

\[ \Delta p = 0.026 Fr_t^{-1.3} We_t^{-0.72} \hat{Re}_l^{2.85} \hat{Re}_g^{-1.92} \left( \frac{\rho_p}{\rho_l} \right)^{1.3} \left( \frac{H_0}{D_c} \right)^{0.5} \rho_g u_g^2 \]  

\[ (6.2) \]

MRD=5.9%
The data were fitted for liquid velocities 0.004-0.012 m/s and gas velocities 1.8-3.6 m/s and mean relative deviation (MRD) was calculated using the relation

$$\text{MRD}=\frac{1}{N} \left( \sum_{i=1}^{N} \frac{\Delta p_{\text{pred}} - \Delta p_{\text{exp}}}{\Delta p_{\text{exp}}} \right) \times 100$$

(6.3)

Where $\Delta p_{\text{pred}}$ and $\Delta p_{\text{exp}}$ are pressure drops predicted by correlations and by experimental data respectively.

In order to see how the above correlation performs for the data produced by earlier workers Figure 6-16 and Figure 6-17 are presented. An error band of ±30 % is placed to accommodate the experimental discrepancies.

As the emphasis in this study was to somehow identify the vast amount of discrepancies and their reasons reported in the literature for the hydrodynamic characteristics of the TCA, hence an attempt has been made in this regard. A careful review of the literature suggests that the following are the various parameters which need to be defined before any reasonable analysis can be done.

1. The column diameter should not be too small as the small diameter columns contribute towards the wall effects, which then makes the operation of any TCA unstable. In this regard another fact to be considered is that for most of the industrial applications the column diameter would be in several feet or more rather than few inches. Hence, the data obtained in column diameters less than 15 cm should have some degree of wall effects and this effect would increase as the column diameter becomes lesser and lesser, making these reported data not so reliable.

2. In literature it has been reported that the column diameter to packing diameter ratio should not be less than 10, as wall effects leads to the congregation of packing at the column wall. This congregation of packing can happen any time and cannot be predicted. Once this happens then there is no fluidization in the column. This ultimately would lead to severe degradation of the TCA performance.

3. The static bed height to column diameter ration should be less than one. For this ratio greater than or close to one leads to bed pulsation and non-uniform fluidization. Even for these ratios much less than unity resulted in very different hydrodynamic and mass transfer behavior as only a few layers of packing is present for large diameter columns. Hence the data obtained for very small static
bed heights where the ration is less than 0.3 and for ratios close to or greater than 1 may have erratic behavior. However, this behavior is less pronounced in small diameter columns mentioned above. This may be the reason why higher static bed heights has been studied but in small diameter columns. One cannot find such static bed height studies in column diameters greater than say 40 cm, as the bed behavior is very erratic and pulsating.

4. The grid free area less than 70 % also results in significant hydrodynamic and mass transfer effects in a TCA column. Most of the earlier studies have been limited to grid free areas less than 60 %. It has been reported in literature that the hydrodynamic and mass transfer effects of grid becomes negligible for grid free areas greater than 70 %.

5. During this study it was observed that uniform fluidization was quite difficult to achieve with small diameter packing namely the 25mm balls. A considerable amount of effort was put in to design a gas distributor to achieve uniform fluidization for these balls which ultimately resulted in large pressure drop in the plenum section. This in turn resulted in a very significant amount of liquid hold up as well as volumetric gas film mass transfer coefficients in the gas distribution section.

Considering the above mentioned parameters this study was carried out for the following conditions.

1. $D_c \geq 15$ cm hence, in this study the column diameter of 44.7 cm was used
2. $D_c/d_p \geq 10$ hence, in this study a ratio of 10, 11.8 to 17.9 was used
3. $H_o/D_c \leq 1$ hence, in this study a ratio of 0.33, 0.56, 0.78 and 1 was used (the ratio of 1 was used in a very limited number of experiments as bed pulsation made the experiments very difficult)

For the analysis of the data reported in literature, the correlation developed in this study was used to predict the pressure drop in the TCA and then these values were compared with the values either obtained from the experimental data reported or from the correlations proposed by various authors. The result is plotted in Figure 6-16 and Figure 6-17 for type I and type 2 respectively. The details of the criteria mentioned above for various workers whose data for type-1 and 2 fluidization used in this comparison study are represented in Table 6.1. These results are presented for type-1 and type-2 fluidization in Figure 6-16 and Figure 6-17. Similarly the data of those workers is plotted in Table 6.1 which do not correspond to the above mentioned criteria.
Considering the type 1 fluidization one can see that most of the data of the earlier workers presented lie within the ±30% band. The only noticeable violations are that of Uchida et al 1977 and some data points of Wozniak 1977 and Soundarajan and Krishnaiah 1998. Now let us consider the Table 6.1 to see how the above mentioned criteria are met by these workers. Uchida et al did not mention the grid free area used in their study. They also used very small diameter balls generally not used in industrial applications. More or less the standard packing diameters used are 25 mm or 38 mm, whereas they used 12.5 mm and 10 mm packing. It was observed in this study that smaller the diameter of the ball more difficult is to have stable and even fluidization in the column. Keeping this observation in mind it seems that due to non-uniform fluidization the pressure drop measured by Uchida et al is less as compared to what is predict by the current correlation.

Regarding the data points relating to the experimental studies of Wozniak 1977, he used the static bed height to column diameter ratio of 1.45 which is much larger than mentioned in literature to effect stable and uniform fluidization. Hence one can expect that the data obtained for this ratio more than 1 will have to some degree non uniform and pulsating fluidization. In case of non-uniform fluidization one can expect lower values of pressure drop as a significant portion of gas flows through the bed without making significant contact with the packing. Similarly one can reason for the data points of Soundarajan and Krishnaiah 1998 as he used a static bed height to column diameter ration of 1, column diameter to packing diameter ratio of 9.4 (which is quite close to the recommendation that this ratio should be greater than 10) and column diameter of only 11.3 cm which should cause wall effects.

The best data as per the above mentioned criteria are that of Shackley(2000) and Micconet(1982) and their data fall quite well. The relatively larger variation for the data of Micconet is due to their use of column to packing diameter ration of 7.9 quite significantly less than 10.

Considering the case of type-2 fluidization the Figure 6-17 is presented. One can see that some points of Inayat 1995 and Strumillo and Kudra 1977 are beyond the ± 30% window. For reasons consider the various ratios they used in their study. Inayat 1995 used a column to packing diameter ratio of 8.8 which is less than the recommended ratio of 10. As all the other criteria of Inayat 1995 are met hence one can see that his data is satisfactorily predicted by the correlation. The case of Strumillo and Kudra 1977 can also be considered by noting that they used a packing density of 770 kg/m³ which in fact is beyond the packing density range used (354 and 442 kg/m³) for which this correlation was established. Hence, one can very well imagine that their data should not be well predicted by the current correlation. However,
the data of Wozniak (1977) is quite satisfactorily predicted, even though they used a static bed height to column diameter ratio of 1.45 which is large as compared to the recommendation and one could expect bed pulsations. If there were some bed pulsations in their experiments then one can expect a lower measured pressure drop, which may be the reason why their data is well predicted. Therefore it may be pointed out that even though the correlation satisfactorily predicts the experimental data yet the data is not reliable for the said geometry, as the geometry itself cannot be scaled up using this data. Hence caution is recommended in the use of this correlation for geometries which do not meet the above mentioned criteria. Even though he used a column diameter of 20cm which is quite within the recommendation whereas the grid free area uses is significantly less than the recommended 70%. This in fact should have caused additional pressure drop which is not seen in their data. This further strengthens the argument that there were bed pulsations in their experimental study.

6.2.7 Conclusions

Pressure drop increases with liquid velocity for both types of operations of TCA. Pressure drop increases sharply at low gas velocities. Pressure drop increases with density and height of packing but pressure drop per unit static bed height decreases with increasing static bed height. At lower gas velocities, pressure drop for smaller diameter packing is more but for higher gas velocities, it is less than that for larger diameter packing. Pressure drop across plenum is comparable to that in the bed for higher gas velocities.

For pressure drop measurements the grid free area, static bed height to column diameter ratio and packing density greater than 450 kg/m$^3$ seems to be the main parameters that generate discrepancy among the pressure drop data reported in literature. Rest of the criteria mentioned above more or less do not strongly affect the pressure drop characteristic of TCA. Hence, the correlations developed in the current study can be used as a sort of universal correlation to predict the pressured drop characteristics of any TCA provided the grid free area is greater than 70 % and packing density is less than say 400 kg/m$^3$ for all column diameters, packing diameters, static bed heights and liquid and gas velocities reported in the literature.

6.3 Expanded Bed Height

Bed expansion data obtained in this study is represented in dimensionless form as ratio of the expanded bed height to the static bed height called the reduced bed height is plotted in Figure
6-20 to Figure 6-27. The various parameters which affect this ratio are now presented and discussed.

6.3.1 Effect of gas and liquid velocities

The static bed begins to expand once the superficial gas velocity reaches the minimum fluidization velocity. The expansion of the bed with increase in gas velocity is due to the upward force of the gas which tends to lift the packing. During the experiments, it was observed that the bed expansion begins at lower gas velocities when the liquid velocity is increased. No change in liquid holdup was observed with an increase in gas velocity once the bed is fully fluidized. This means that as the bed expands with increasing gas velocity, there is practically no increase in the interstitial gas velocity. Thus the reduced bed height ($H/H_0$) increases with increase in gas velocity. The effect of liquid and gas velocities is shown in Figure 6-20 to Figure 6-23.

However, there is a slight increase in bed expansion with an increase in the liquid flow rate as can be observed from Figure 6-20 and Figure 6-21. With an increase in the liquid velocity the amount of liquid in the column increases which helps the increase the bed expansion. The increase in bed expansion with increase in liquid flow rate is more prominent at higher gas flow rates.

The effect of liquid and gas velocities on bed expansion are more pronounced in case of type-2 fluidization as compared to type-1. Moreover, this increasing effect is even more noticeable for higher liquid and gas velocities.

6.3.2 Effect of packing diameter

There is a pronounced effect of packing size on bed expansion. Up to 20 % decrease of bed expansion can be observed in Figure 6-24 when the packing size is changed from 25 mm to 45 mm. Although the apparent density of the packing is same for all the packing used (354 kg/m³) however, the weight of the smaller packing is less as compared to the bigger packing. Hence, one can why the lighter but smaller packing is more easily fluidized as compared to the larger one.

6.3.3 Effect of packing density

The effect of packing density on the reduced bed height is presented in Figure 6-25. Considering the above argument that light weight packing is easier to fluidize, hence packing
with same diameter but increasing density will fluidize to a smaller height as compared to the packing with lesser density.

6.3.4 Effect of static bed height

The effects of increasing static bed height on the reduced bed height are given in Figure 6-26 and Figure 6-27.

There is a slight decreasing effect of static bed height on the reduced bed height for type 1 fluidization. However this effect is only present at the static bed height of 0.15 m, beyond this static bed height as the static bed height increases there is practically no increase in reduced bed height. This initial drop in reduced bed height can be explained by considering that as the static bed height increases the increase in weight of the bed is more than the force generated by the upward moving gas. Hence the relative increase in the downward force reduced the reduced bed height.

For type 2 fluidization, there is practically negligible effect of increasing static bed height on the reduced bed height.

Following correlation is developed for the bed expansion from the data obtained in this study.

\[
\frac{H}{H_0} = 1.704 u_i^{0.13} u_g^{0.544} H_0^{-0.1055} \rho_p^{-0.2426} d_p^{-0.38}
\]

The data were fitted for liquid velocities 0.004-0.012 m/s and gas velocities 1.8-3.6 m/s and above all correlations fit in the data within maximum error of ±15 % and mean relative deviation (MRD) of 5%
Figure 6-1 Effect of gas and liquid velocities on pressure drop for type 1

Figure 6-2 Effect of gas and liquid velocities on pressure drop for type 2
Figure 6-3: Effect of liquid and gas velocities on pressure drop for 25 mm dia packing

Figure 6-4: Effect of liquid and gas velocities on pressure drop for 45 mm dia packing
Figure 6-5 Effect of density of packing on pressure drop

Figure 6-6 Effect of diameter of packing on pressure drop
Figure 6-7 Effect of height of packing on pressure drop for type 1 TCA operation

Figure 6-8 Effect of height of packing on pressure drop for type 2 TCA operation
Figure 6-9: Effect of pressure drop per unit static bed height of packing on pressure drop

Figure 6-10: Effect of pressure drop across grid
Figure 6-11 Comparison of pressure drop by direct measurement with that calculated from liquid holdup

Figure 6-12 Comparison with literature correlations for type 1 TCA operation at constant gas velocity
Figure 6-13: Comparison with literature correlations for type 1 TCA operation at constant liquid velocity.

Figure 6-14: Comparison with literature correlations for type 2 TCA operation at constant gas velocity.
Figure 6-15: Comparison with literature correlations for type 2 TCA operation at constant liquid velocity.

Figure 6-16: Pressure drop from literature vs. present study correlation for type 1 fluidization
Figure 6-17: Pressure drop from literature vs. present study correlation for type2 fluidization

Table 6-1: Parameters used in graphs for the comparison of pressure drop with present study correlation

<table>
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<tr>
<th>Author</th>
<th>$H_0$</th>
<th>$\rho_p$</th>
<th>$D_c$</th>
<th>$d_p$</th>
<th>$f$</th>
<th>$D_c/d_p$</th>
<th>$H_0/D_c$</th>
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<td>243</td>
<td>0.113</td>
<td>0.012</td>
<td>0.728</td>
<td>9.42</td>
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<td>170</td>
<td>0.195</td>
<td>0.0125, 0.01</td>
<td>N/A</td>
<td>15.6-19.5</td>
<td>0.64-0.97</td>
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<td>(Balabekov et al. 1971)</td>
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<td>356</td>
<td>0.175</td>
<td>0.022</td>
<td>0.6</td>
<td>7.95</td>
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<td>0.46</td>
<td>0.038</td>
<td>0.74</td>
<td>12.11</td>
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<td>(Inayat 1995)</td>
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<td>327, 162</td>
<td>0.22</td>
<td>0.025, 0.045</td>
<td>0.72</td>
<td>8.8, 4.9</td>
<td>0.48</td>
</tr>
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<td>(Barile and Meyer 1971)</td>
<td>0.15</td>
<td>160</td>
<td>0.286</td>
<td>0.019</td>
<td>0.82</td>
<td>15.2</td>
<td>0.52</td>
</tr>
<tr>
<td>(Rama et al. 1983)</td>
<td>0.12</td>
<td>53</td>
<td>0.15</td>
<td>0.038</td>
<td>0.7</td>
<td>3.95</td>
<td>0.8</td>
</tr>
<tr>
<td>(El-Dessouky 1993)</td>
<td>0.3, 0.4, 0.5, 0.2, 770, 0.2, 0.015, 0.65, 13.33, 0.42, 0.100, 0.29, 0.19, 0.60, 10.53, 1.45</td>
<td>0.2</td>
<td>0.022</td>
<td>0.6</td>
<td>7.95</td>
<td>0.69</td>
<td>0.46</td>
</tr>
<tr>
<td>(Strumillo et al. 1974)</td>
<td>0.85</td>
<td>770</td>
<td>0.2</td>
<td>0.0127</td>
<td>0.708</td>
<td>15.75</td>
<td>1.5, 2, 2.5</td>
</tr>
<tr>
<td>(Bruce et al. 2002)</td>
<td>0.107</td>
<td>110</td>
<td>0.11</td>
<td>0.016</td>
<td>0.70</td>
<td>7.06</td>
<td>0.95</td>
</tr>
<tr>
<td>(Wozniak 1977)</td>
<td>0.29</td>
<td>266, 442</td>
<td>0.200</td>
<td>0.019</td>
<td>0.60</td>
<td>10.53</td>
<td>1.45</td>
</tr>
</tbody>
</table>
Figure 6-18: Pressure drop across plenum without incorporating velocity head in air inlet pipe

Figure 6-19: Pressure drop across plenum by adding velocity head in air inlet pipe into measured pressure drop
Figure 6-20: Effect of liquid velocities on expanded bed height for type 1 TCA operation

Figure 6-21: Effect of liquid velocities on bed expansion for type 2 TCA operation
Figure 6-22: Effect of gas velocities on expanded bed height for type 1 TCA operation

Figure 6-23: Effect of gas velocities on bed expansion for type 2 TCA operation
Figure 6-24 Effect of diameter of packing on bed expansion

Figure 6-25 Effect of density of packing on bed expansion
Figure 6-26: Effect of gas velocity on bed expansion for different static bed heights for type 1

Figure 6-27: Effect of gas velocity on bed expansion for different static bed heights for type 2
Figure 6-28 Comparison of present correlation with literature correlations

6.4 Liquid holdup

Liquid hold is a very important hydrodynamic parameter of TCA. It controls two main characteristics of any TCA device. On one side it controls the pressure drop whereas on the other side it plays a very significant role in the mass transfer characteristics of a TCA. Hence, in literature it occupies a central position for study. Similarly in this study a significant part of the research was focused to determine the liquid hold up characteristics of TCA and efforts has also been made to sort out the reasons of discrepancies among the reported literature. In order to exactly specify the amount of liquid held in the TCA bed which contributes both towards pressure drop and the volumetric mass transfer coefficients, all necessary measures were taken to remove any sort of errors. (For details chapter 4 for this). Hence, the true liquid holdup as per its approved definition used in literature, only that volume of liquid was measured and reported which during operation remained in the bed and contributed towards the bed hydrodynamic and mass transfer performance. Therefore, the liquid held in the gas distribution section, liquid distributor and the accompanying piping within the column was also measured and was subtracted from the total volume of liquid held in the column. This volume of liquid is the true liquid contributing towards the hydrodynamics and mass transfer.
characteristics of TCA bed. As mentioned in detail in chapter 1, it was observed that the major portion of the concerned literature does not mention whether the above mentioned correction in liquid hold up measurement has been applied or not. Hence while comparing the liquid hold up data of this study with that of literature care has been taken to account for this observation. The liquid holdup based on static bed height volume was calculated using the following equation.

\[ \varepsilon_{l,sl} = \frac{V_l}{A_s H_s} \]

### 6.4.1 Effect of gas and liquid velocities

Liquid holdup in the bed increases with an increase in the liquid velocity. However, no significant effect of gas velocity was observed both for type 1 and type 2 fluidization (Figure 6-29 and Figure 6-30). The data for liquid hold up reported in this thesis is for superficial gas velocities greater than or equal to 1.8 m/s. This velocity is well above the minimum fluidization velocity. Hence the data reported is for fully fluidized beds. In the fully fluidized beds there is sufficient free area available for the gas to flow without any significant hindrance. Thus the liquid flowing down causes negligible drag for the gas flowing upwards. Therefore one can expect no significant rise in liquid hold up with increasing gas velocities. However the liquid holdup should increase with an increase in the liquid velocity as now more liquid is present in the bed for the entire range of gas velocities.

Whereas the liquid holdup in the column which includes the plenum and liquid distributor increases with an increase in both the liquid and gas velocities for both types 1 and 2 fluidization modes of TCA operation (Figure 6-31 & Figure 6-32). This increase is due to two factors:

1. Due to the liquid volume held in the plenum section, which increases with increasing gas velocities. This increase in the liquid volume held in the limited space of the plenum is due to an increase in turbulence and re-circulations within the plenum with increasing gas velocity. Moreover, the liquid volume held in the plenum depends on its design, which will vary from one design to another.

2. The liquid holdup increases with an increase in the liquid velocity since more liquid is present in the column at higher liquid velocities for the entire range of gas velocities.
6.4.2 Effect of packing density

Liquid holdup increases with an increase in the packing density (Figure: 6-33). This behavior can be explained by considering the fact that the higher density packing has more weight, thus one can expect less expansion of the bed. If for a particular gas and liquid flow rate the bed expands less than the interstitial gas velocity must increase. This increase in interstitial velocity should in turn result in an increased liquid holdup. Figure: 6-33 also shows that the gradient of liquid hold up with increasing density is steeper when density is increased from 270 to 354 kg/m³. This is the region where the mode of fluidization changes from Type 1 to Type 2. However this effect is reduced as the liquid flow rate is increased.

6.4.3 Effect of packing diameter

Little effect of diameter of packing on liquid holdup was observed (Figure 6-34). The data reported is for fully fluidized beds, i.e. gas velocities are well above minimum fluidization velocities. At lower gas velocities the liquid holdup in TCA using smaller diameter packing is expected to be more as compared with those having larger diameter due to high surface area of contact causing friction and low free area in the packing for gas flow when the bed is not fully fluidized. However, as the bed is fully fluidized, the effect of friction and free area are diminished.

6.4.4 Effect of static bed height

The effects of static bed height on the liquid holdup are presented in Figure 6-35 and Figure 6-36 for type-1 and type-2 fluidizations respectively. For type-1 fluidization one can see that the effect of static bed height is more pronounced as the static bed height changes from 0.15m to 0.25m. However, as the static bed height further increases the drop in liquid holdup becomes less and less significant. Whereas for type-2 fluidization the decrease in liquid holdup with an increase in the static bed height is very uniform and even the slope of the liquid holdup curves remain almost same as the liquid velocity increases.

It was observed during the hydrodynamic study of the TCA that the liquid held on the grid becomes lesser and lesser as either the packing density or static bed height is increased. Which means that even for grid free area greater than 70% there is some contribution of grid towards liquid holdup and hence towards mass transfer characteristics also. Once this effect is diminished as the weight of the bed increases the liquid hold up behavior becomes more and more uniform. This observation is in line with the results produced in Figure 6-35 and Figure
Another comment can be made here that by increasing the static bed height one can eliminate the contribution of the grid, whereas deeper beds can and do cause bed pulsations which would then cause unpredictable liquid holdup. Similarly these bed pulsations can also be controlled by decreasing the column diameter, where the presence of walls suppresses the bed pulsation. However, for larger diameter columns one can predict that these wall effects will not be that significant, hence there will be bed pulsations. This point should be remembered when the liquid holdup data generated in this study of large diameter column is compared with relatively small diameter columns.

Finally the liquid holdup for the whole column including the effect of plenum section with increasing static bed height is presented in Figure 6-37 and Figure 6-38 for type-1 and type-2 respectively. Pronounced decreasing effect of static bed height on liquid holdup can be observed both for type 1 and type 2 TCA operations when liquid held in plenum is not subtracted from total liquid held. Here one can only make one comment that this is not the behavior of the TCA bed, it is the effect of the bed and the plenum section. Hence, it does not represent the true characteristics of any TCA bed.

6.4.5 Comparison with published results

The experimental data of liquid holdup obtained in this study is in agreement in terms of trends with those of Chen and Douglas (1968), Kito et al. (1978), Vunjak et al. (1987a) and Bruce et. al (2004) who have reported no significant effect of gas velocity on liquid holdup. However, Rama et al. (1983), Hekmat-Nazemi (1992), Inayat (1995), Gimenes and Handley (1998), Soundarajan and Krishnaiah (1998a) and Lyashuk (2001) reported an increase in the liquid holdup with gas velocity for the types of fluidizations they studied which can be evaluated from their ranges of variables mentioned in Table 2-1. However, there is consensus in literature that for type 2 fluidization, there is an increase in liquid holdup until the bed is fully fluidized.

A comparison of the predicted liquid holdup from the correlations available in literature with the correlation developed in this study is given in Figure 6-40 to Figure 6-46. One can see that the liquid holdup predicted by the present study correlation is much lower as compared to that of the earlier investigators. However, the liquid holdup for the column and for the column plus plenum is presented separately to signify the effect of the liquid held in the plenum. Considering all this, now one can suggest the possible reasons for the discrepancies in the reported liquid holdup in TCA. It is the plenum liquid holdup and the liquid held in the liquid distributor and piping which has been included by earlier...
investigators. The level of discrepancy among the reported data of liquid hold up in this study and by earlier investigators who have applied the plenum liquid holdup correction lie within the experimental variations.

Gel'perin et al. (1968) and Bruce et al. (2004) predicted no increase of liquid holdup with an increase in density of packing. Vunjak-Novakovic (1987a) predicted increase in liquid holdup for type 2 mode of fluidization while Kito et al. (1978), Soundarajan and Krishnaiah (1998a) predicted significant increase in liquid holdup with increase in packing density for both types of fluidizations. The findings of this study agree with the findings of Kito et al. (1978), Soundarajan and Krishnaiah (1998a).

No effect of diameter of packing on liquid holdup was observed (Figure 6-34). The data reported is for fully fluidized beds, i.e. gas velocities are well above minimum fluidization velocities. It has earlier been investigated by (Chen and Douglas 1968; Gel'perin et al. 1968; Kito et al. 1978; Vunjak-Novakovic et al. 1987a) that liquid holdup decreases with increasing packing diameter however, in the present study (Figure 6.44), no significant change in liquid holdup was observed.

Correlations developed by regression analysis of the data with respective mean relative deviation (MRD) are presented below. Effect of grid free area f was not studied hence these correlations are for grid free area 0.74.

MRD was calculated by the formula

$$\frac{1}{N} \sum_{i=1}^{N} \left[ \frac{\varepsilon_{i,\text{pred.}} - \varepsilon_{i,\text{exp.}}}{\varepsilon_{i,\text{exp.}}} \right] \times 100$$  \hspace{1cm} (6.4)

Where $\varepsilon_{i,\text{pred.}}$ and $\varepsilon_{i,\text{exp.}}$ are liquid holdup predicted by correlations and by experimental data respectively.

Type 1 bed only

$$\varepsilon_{i,\text{st}} = 2069 \hat{R} e_l^{-0.77} F r_l^{0.59} W e_l^{0.27} \left( \frac{\rho_p}{\rho_l} \right)^{0.44} \left( \frac{H_0}{D_c} \right)^{-0.10}$$  \hspace{1cm} (6.5)

MRD = 5.5%

Type 2 bed only

$$\varepsilon_{i,\text{st}} = 325612 \hat{R} e_l^{-1.15} F r W e_l^{0.42} \left( \frac{\rho_p}{\rho_l} \right)^{0.20} \left( \frac{H_0}{D_c} \right)^{-0.15}$$  \hspace{1cm} (6.6)

MRD = 6.4%

Plenum only
\[ h_l = 0.0739 u_g^{1.21} u_i^{0.81} \]  \hspace{1cm} (6.7)

MRD=5%

Liquid holdup in the plenum depends on liquid velocity, gas velocity, height of plenum and type of gas distributor housed in the plenum. Although liquid held in plenum will vary from system to system, however, correlation for liquid held in plenum was also developed as in above equation (6.7).

Figure 6-47 and Figure 6-48 show the relative comparison of the correlation developed in this study with some of the selected experimental data present in the literature. To fully comprehend the data produced in these plots. Please consider the Table 6-2 which shows the summary of the literature data used for comparison. It is also mentioned in this table that which of the earlier workers have applied the correction for the liquid holdup in the plenum section and liquid distributors and for which the literature is silent.

The criteria set above to compare the experimental data of various earlier workers with the correlations developed in this study for liquid holdup is used again. With these results an effort is made to find out the reasons of discrepancies in the reported liquid holdup data. Figure 6-47 shows the comparison of the experimental data of earlier workers plotted against the liquid holdup predicted by the correlation developed in this study for Type-1 fluidization. One can see that some of the data of Inayat 1995, Shackley 2000, Soundarajan and Krishnaiah(1998a), Vunjak et.al 1987, Rama et.al 1983 and Gimenes and Handley(1999) fall within the ±30% band, where as some additional data of these workers along with, Kito et al 1978 and Bruce et.al 2004 lie within the ±40% band.

Considering the criteria set above the data reported by Shackley 2000, Gelperin et.al 1968, and Peterson 1987 is suitable to compare. In fact the data of Shackley 2000 is the best reported data for comparison. One can see that the data reported by those workers who meet the criteria set above have good agreement with the current data and the respective correlation. See the conditions met by the data of Shackley and Gimenes and how well it is predicted by the current correlation. The data reported by Rama et al 1983 is quite well predicted by the correlation. They used the RTD technique for measuring the liquid holdup. However, they used extremely light packing as the apparent density is only 53 kg/m³ and Dc/dp ratio is only 3.95. They applied the liquid holdup correction for the distributor sections.

Another very important point to note is for Type 1 fluidization, where the current correlation consistently under predicts the liquid holdup. This is particularly true for
Soundarajan at al 1998, Gelperin et.al 1968 and Bruce 2004. Literature concerning these workers is silent about the liquid hold up contribution of the gas and liquid distributors. This seems to be one of the major reasons why the liquid hold up is under predicted by the current correlation. However, the case of Kito at al(1978) where the liquid hold up correction has been applied yet the current liquid holdup is over predicted. Reason may lie in the wall effects as in this particular case the Dc/dp ratio is only 5.13 as compared to the recommended value of 10. Similarly the Ho/Dc is one where the recommended value is less than one, moreover the column diameter is only 10 cm.

Similarly the data of Vunjak et.al 1987 is over predicted. Here, the Ho/Dc ratio is 2.14 far above the recommended value of one. Here one can expect the correlation needs a correction factor if needs to be applied for Ho/Dc ratio of more than one.Further experimental studies need to be carried to justify the use of RTD in TCA type systems.

The reasons why the data of Inayat 1995 who also applied the correction for liquid hold up lie also in the Dc/dp ratio less than 10 and in the case of type-1 studies he used a ratio of only 4.9.

Figure 6-48 shows the performance of the equation developed for the type-2 fluidization in predicting the liquid hold up data reported in literature. The data of Shackley 2000, Gelperin et.al 1968, Inayat 1995, Vunjak et.al 1987 is very well predicted by the correlation. However, the data of Peterson 1987, Soundarajan and Krishnaiah 1998 and Kito et al 1978 is very much of the ±30% band. The reasons for the so much under prediction for the case of Peterson 1987 may be the use of RTD technique to experimentally determine the liquid holdup. The use of RTD in TCA needs to be studied further as one can see that the data reported using this technique is much higher as compared to the others’ data on similar systems. In TCA there is extensive back mixing of the liquid phase and presumably no in the continuous gas phase. In the RTD technique it is assumed that there is plug flow and its use for CSTR type systems where the response has a time delay depending upon the size of the system needs to be studied further.

The data of Soundarajan and Krishnaiah 1998 is over predicted. Even though they have not mentioned that they applied the distributor corrections yet there data is over predicted. They have used a column diameter of only 11.3 cm which is less than the recommended criteria, their Ho/Dc ratio is not less than one and Dc/dp ratio is also less than 10 though close. It seems somewhat odd, if the data of Bruce et al 2004 is considered, who also did not applied the distributor correction. The experimental conditions of Soundrajan(1998) and Bruce et.al(2004) are very similar(Table 6-2). The only major difference is the packing
density, Soundarajan and Krishnaiah (1998) used a density of 599 kg/m³ whereas Bruce et.al 2004 used 835 kg/m³. Yet there experimental data reported results are so different from each other that question is raised on the validity of the data. Hence the reasons for the variation of their data from that of the predicted values lie in the validity of their data. However it seems that the data produced by Bruce et.al 2004 is correct as they have not applied the distributor corrections and hence there measured values are more than those predicted by the current correlation. Another point to be mentioned here is the density used to plot the data of Bruce et.al is 835 kg/m³ which is almost twice that of the density used in the current study. The correlation presented here is valid for the ranges of parameters used in this study. Extrapolation far beyond this range is not recommended.

A few data points of Inayat 1995 are also over predicted by the correlation, whereas he has applied the distributor corrections. The reason may be the ratio Dc/dp of 8.8 used which is less than 10 the recommended value.

6.4.6 Conclusions

Based on the experimental results, it can be concluded that liquid holdup increases with decreasing static bed height, increasing density of packing and liquid flow rates. However, there is negligible effect of packing diameter and gas velocity on liquid holdup. The significant quantitative differences among the liquid holdup data reported in literature from correlations by the earlier investigators may be due to the reason that they have not included the effect of liquid volume held in either liquid distributor pipe or plenum or both. Since the liquid volume held in the liquid and gas distributor sections depends on its design, hence it would not be possible to apply some sort of a correction strategy to all the reported data. The use of Dc < 15 cm, Ho/Dc ≥ 1 and Dc/dp< 10 lead to wall effects which generate the experimental data that is quite different from the one generated by experiments where these criteria is met. However, more or less consensus exists in the qualitative trends of the effect of various parameters on the liquid holdup in a TCA. However, based on the above analysis one can recommend that the correlations and data reported with the correction of liquid and gas distribution sections are more reliable for TCA design considerations as can be seen how well this correlation predicts other workers data.
Figure 6-29: Effect of liquid and gas velocities on liquid holdup for type 1 fluidization

Figure 6-30: Effect of liquid and gas velocities on liquid holdup for type 2
Figure 6-31 Effect of liquid and gas velocities on liquid holdup with plenum for type 1

Figure 6-32 Effect of liquid and gas velocities on liquid holdup with plenum for type 2
Figure 6-33 Effect of density of packing on liquid holdup

Figure 6-34 Effect of diameter of packing on liquid holdup
Figure 6-35 Effect of static bed height on liquid holdup for type 1

Figure 6-36 Effect of static bed height on liquid holdup for type 2
Figure 6-37 Effect of static bed height on liquid holdup for type 1 with plenum

Figure 6-38 Effect of static bed height on liquid holdup with plenum for type 2
Figure 6-39: Liquid holdup in empty column

Figure 6-40: Comparison of effect of liquid velocity on liquid holdup for type 1
Figure 6-41 Comparison of effect of liquid velocity on liquid holdup for type 2

Figure 6-42 Comparison of effect of gas velocity on liquid holdup for type 1
Figure 6-43 Comparison of effect of gas velocity on liquid holdup for type 2

Figure 6-44 Comparison of effect of diameter of packing on liquid holdup type 2
Figure 6-45 Comparison of effect of static bed height on liquid holdup for type 1

Figure 6-46 Comparison of effect of static bed height on liquid holdup for type 2
Figure 6-47: Liquid holdup from literature vs. present study correlation for type1 fluidization

Figure 6-48: Liquid holdup from literature vs. present study correlation for type2 fluidization
Table 6-2: Parameters used in graphs for the comparison of liquid holdup with present study correlation

<table>
<thead>
<tr>
<th>Author</th>
<th>$H_0$</th>
<th>Packing density</th>
<th>$D_c$</th>
<th>$d_p$</th>
<th>$f$</th>
<th>$D_c/d_p$</th>
<th>$H_0/D_c$</th>
<th>Method used</th>
<th>Plenum correction</th>
</tr>
</thead>
<tbody>
<tr>
<td>Soundarajan and Krishnaiah(1998a)</td>
<td>0.113</td>
<td>243, 599</td>
<td>0.113</td>
<td>0.012</td>
<td>0.728</td>
<td>9.42</td>
<td>1.00</td>
<td>Shut off valve</td>
<td>N</td>
</tr>
<tr>
<td>Inayat (1995)</td>
<td>0.105</td>
<td>162, 327</td>
<td>0.22</td>
<td>0.025, 0.045</td>
<td>0.72</td>
<td>8.8, 4.9</td>
<td>0.48</td>
<td>Shut off valve</td>
<td>Y</td>
</tr>
<tr>
<td>Rama et al (1983)</td>
<td>0.12</td>
<td>53</td>
<td>0.15</td>
<td>0.038</td>
<td>0.7</td>
<td>3.95</td>
<td>0.8</td>
<td>RTD</td>
<td>Y</td>
</tr>
<tr>
<td>Paterson and Clift(1987)</td>
<td>0.146</td>
<td>432</td>
<td>0.457</td>
<td>0.038</td>
<td>0.615</td>
<td>12.03</td>
<td>0.32</td>
<td>RTD</td>
<td>Y</td>
</tr>
<tr>
<td>Vunjak et.al 1987</td>
<td>0.3</td>
<td>379, 203</td>
<td>0.14</td>
<td>0.01, 0.025</td>
<td>0.78</td>
<td>14, 5.6</td>
<td>2.14</td>
<td>Shut off valve</td>
<td>Y</td>
</tr>
<tr>
<td>Gelperin et.al (1968)</td>
<td>0.26</td>
<td>160</td>
<td>0.37</td>
<td>0.006-0.035</td>
<td>0.7</td>
<td>10</td>
<td>0.70</td>
<td>Shut off valve</td>
<td>N</td>
</tr>
<tr>
<td>Kito et al (1978)</td>
<td>0.1</td>
<td>170</td>
<td>0.1</td>
<td>0.0195</td>
<td>0.712</td>
<td>5.13</td>
<td>1.00</td>
<td>Shut off valve</td>
<td>Y</td>
</tr>
<tr>
<td>Bruce et al (2004)</td>
<td>0.107</td>
<td>215, 835</td>
<td>0.113</td>
<td>0.012</td>
<td>0.7</td>
<td>9.42</td>
<td>0.95</td>
<td>Shut off valve</td>
<td>N</td>
</tr>
<tr>
<td>Shackley (2000)</td>
<td>0.1-0.2</td>
<td>168</td>
<td>0.46</td>
<td>0.038</td>
<td>0.67</td>
<td>12.1</td>
<td>0.22-0.43</td>
<td>pressure drop</td>
<td>Y</td>
</tr>
</tbody>
</table>

Y= Either the authors have mentioned plenum subtraction or they have used shutter above plenum for quick closing

N=Either the authors have not mentioned about plenum contribution or their set up does not show such effort.

Note: Kito et.al 1978 and Vunjak et.al 1987 used shutters below support grid but did not mention about the liquid present in liquid distributor or its pipe
Chapter 7

7 Mass Transfer Results and Discussions

7.1 Introduction

In this chapter, results obtained for the volumetric gas film transfer coefficients measurement are presented and discussed in detail. As there is a significant contribution of pressure drop and liquid holdup in plenum section as mentioned in chapter 6, therefore, logic does not permit to ignore the effect of plenum in the mass transfer studies. In Table 3-2, the authors who have taken into account the effect of plenum in their mass transfer studies are indicated.

7.2 Presentation of results

In order to represent the mass transfer characteristics of the TCA fluidized bed, the effect of plenum measured is presented and discussed. This is followed by the presentation of the volumetric gas film mass transfer coefficient experimental data. Then the volumetric gas film mass transfer coefficient data for the TCA bed only is presented and discussed. Lastly the same data is presented in terms of the operational efficiency of the TCA bed as a few of the earlier workers have also done.

In this presentation and discussion of the mass transfer data the following equations and nomenclature is used

The various equations (derived in chapter 5 and defined in terms of various heights) used to calculate these quantities are given below.

\[
(k_g a)_x = \frac{G^N_{G,x}}{(H_y)P_G RT} = \frac{G^N_{G,x}}{(H_y)P_G M_G}
\] (7.1)

Where

\(G, \rho_G, M_G, P_G, T_G\) = Superficial mass velocity, density, molecular wt., pressure and temperature of air and

\(H_y\) the height in which mass transfer takes place, where \(y = s\) = static bed height, \(y = e\) = expanded bed height, \(y = p\) = plenum height.
(k_g a)_x are the gas side volumetric mass transfer coefficients, where x = c = whole column (bed plus plenum), x = p = plenum only and x = b = bed only,

\[ N_{G,x} \] are the number of gas phase transfer units for x = c = whole column (bed plus plenum), x = p = plenum only and x = b = bed only

Similarly the number of gas phase transfer units for column and plenum are given as;

\[
N_{G,x} = \ln \left[ \frac{T_{G_1, x} - T_{s, x}}{T_{G_2, x} - T_{s, x}} \right] \tag{7.2}
\]

In above equations, \( T_G \) is temperature of gas and \( T_s \) is adiabatic saturation temperature. Where x can have values as mentioned above and 1 is for inlet and 2 for outlet

Details of the equations based on how the mass transfer data can be represented are given in Table 7-1

Table 7-1: Equations used in volumetric gas film transfer coefficient and operational efficiency

<table>
<thead>
<tr>
<th>Sr.No</th>
<th>Quantity</th>
<th>Equation</th>
</tr>
</thead>
<tbody>
<tr>
<td>1</td>
<td>Volumetric gas film mass transfer coefficient of bed based on static bed volume</td>
<td>((k_g a)<em>{b,x} = \frac{G N</em>{G,b}}{(H_s)P_G M_G} ) \tag{7.3}</td>
</tr>
<tr>
<td>2</td>
<td>Volumetric gas film mass transfer coefficient of plenum based on plenum volume</td>
<td>((k_g a)<em>{p} = \frac{G N</em>{G,p}}{(H_p)P_G M_G} ) \tag{7.4}</td>
</tr>
<tr>
<td>3</td>
<td>Volumetric gas film mass transfer coefficient of column based on static bed volume</td>
<td>((k_g a)<em>{c,x} = \frac{G N</em>{G,c}}{(H_s)P_G M_G} ) \tag{7.5}</td>
</tr>
<tr>
<td>4</td>
<td>Operational mass transfer efficiency of column</td>
<td>((\eta_{op})<em>c = (k_g a)</em>{c,p} \frac{(H_s + H_p)}{\Delta p u_g} ) \tag{7.6}</td>
</tr>
<tr>
<td>5</td>
<td>Operational mass transfer efficiency of bed</td>
<td>((\eta_{op})<em>b = (k_g a)</em>{b} \frac{H_s}{\Delta p u_g} ) \tag{7.7}</td>
</tr>
<tr>
<td>6</td>
<td>Operational mass transfer efficiency of plenum</td>
<td>((\eta_{op})<em>p = (k_g a)</em>{p} \frac{H_p}{\Delta p u_g} ) \tag{7.8}</td>
</tr>
</tbody>
</table>

The basis of the calculations for different types of mass transfer coefficients is that the number of transfer units of the column is the sum of the transfer units of plenum and bed i.e.

\[ N_{G,c} = N_{G,p} + N_{G,b} \] \tag{7.9}
7.3 Mass transfer in the plenum

As discussed above the mass transfer characteristics of TCA should not include the effects of mass transfer occurring in the plenum. Hence, to quantify the mass transfer characteristics of TCA, which are due to the violent turbulent motion of the fluidized packing, the contribution of plenum must be quantified and subtracted from that of the whole column. As most of the studies reported in literature generally are small scale equipment where due to the space restrictions, the gas distribution sections are designed in such a way to give an even distribution of gas velocity across the column diameter. In doing so the pressure drop considerations are generally ignored. In the vast majority of the literature reported on TCA there is no mention of the design of the plenum and also, there is no mention of whether the correction for the mass transfer occurring in the plenum has been made or not. Only very few studies reported in literature has discussed this and tried to quantify the contribution of the plenum.

Hence in this study the mass transfer in the plenum was studied separately and results are plotted in Figure 7-1. In plenum section, there is a significant increase in \((k_{ga})_{p}\) with gas velocity. However, the increase with liquid velocity is relatively less significant at low gas velocities but becomes significant at high gas velocities. As discussed earlier that this behavior of plenum is design dependent and any logical explanation should discuss the design of the plenum. As the mass transfer studies of plenum is not the goal of this study hence the plenum was designed primarily taking into account the even distribution of gas in the column leading to as flat as possible radial velocity profile. This was ensured by noting that the bed fluidizes evenly.

Considering the performance of the plenum one can consider it to be a relatively good mass transfer device, hence, mass transferred in the plenum is not negligible as compared to mass transferred in the column.

Therefore it is pointed out that mass transfer in the column as generally reported in most of the literature in terms of \((k_{ga})_{c,s}\) is not a true representation of mass transfer in a TCA bed.
7.4 Volumetric Mass transfer coefficients in the TCA bed based on the static bed volume

7.4.1 The effect of gas and liquid velocities

The volumetric gas film mass transfer coefficients of the TCA bed based on static bed volume i.e. \( (k_{g\alpha})_{b,s} \) have been plotted from Figure 7-2 to Figure 7-13. The following can be concluded from these plots.

1. There is practically little effect of liquid velocity on \( (k_{g\alpha})_{b,s} \) for all the gas velocities studied.
2. The gas velocity has a significant effect on \( (k_{g\alpha})_{b,s} \) for all liquid velocities.
3. \( (k_{g\alpha})_{b,s} \) passes through a maxima with increasing gas velocities.
4. These above mentioned effects of liquid and gas velocities on \( (k_{g\alpha})_{b,s} \) are valid for both type 1 and type 2 fluidizations.

Considering Figure 7-1 (the plenum behavior) and Figure 9-1 to Figure 9-4 (the column plus plenum behavior based on static bed height) one can conclude that the slight effect of liquid velocity on \( (k_{g\alpha})_{c,s} \) (As can be seen in appendix) is due to the contribution of the plenum. Therefore it can be concluded that for TCA, \( (k_{g\alpha})_{b,s} \) is almost independent of the liquid velocity. Moreover, this inference is further strengthened by another argument that the volumetric gas film transfer coefficients should not depend on the liquid velocities as the mass transfer resistance lie on the gas side. Hence any increased turbulence on the liquid side should not improve the transfer coefficients, whereas the increased turbulence on the gas side should improve the transfer coefficients.

As far as the maxima of \( (k_{g\alpha})_{b,s} \) is concerned, it can be explained that at relatively higher gas velocities may be there is a channeling effect on the gas side. Once this particular gas velocity is reached then gas tends to flow through the bed without making an increased contact with the liquid. This passing of \( (k_{g\alpha})_{b,s} \) through a maxima is observed both in type 1 and type 2 fluidizations. However, the velocities at which they occur are fluidization type dependent. For type 2 (packing density 354 kg/m³) the gas velocity is around 3 m/s, for type 1 (packing density 180 kg/m³) it is around 2.5 m/s and for the borderline case (packing density 270 kg/m³) the velocity seems to be more than 2.5 m/s.

One can make another conclusion that the \( (k_{g\alpha})_{b,s} \) represents the magnitude of the mass transfer occurring in the TCA bed. Therefore, the results shown in Figure 7-2 to Figure 7-13 represent the magnitude of mass transfer occurring in a TCA bed. These figures can also be interpreted as there is little effect of liquid velocity on the mass transfer occurring in the bed.
However, for the same conditions the hydrodynamic characteristics of the TCA bed should also be kept in mind. The hydrodynamic characteristics shown in the previous chapter shows that the liquid velocity significantly increases the liquid hold up, expanded bed height and pressure drop. This effect is more pronounced for type 2 fluidizations as compared to type1. Hence, there is no significant effect of liquid velocity on \((k_ga)_{b,s}\) but significant increasing effect on pressure drop and liquid holdup. This further suggests that for the most economical operation of a TCA should be at conditions where

- pressure drop is minimum (to conserve energy)
- expanded bed height is minimum (to conserve space)
- liquid hold up is minimum (to conserve the liquid used)

However, the packing density, diameter of the packing and the static bed height also affect the \((k_ga)_{b,s}\) and are now discussed.

7.4.2 Effect of packing density on \((k_ga)_{b,s}\)

Effects of packing density on \((k_ga)_{b,s}\) are given in Figure 7-12 and Figure 7-13.

The following inferences can be drawn from these experimental results.

- The volumetric mass transfer coefficient increases significantly with increasing packing density. This increase is more noticeable at higher gas flow rates.
- This trend is valid for both type 1 and 2 fluidizations.
- There seems to be a maxima for packing densities less than or equal to 354 kg/m\(^3\) which means that one can say that for type-1 fluidization \((k_ga)_{b,s}\) passes through a maxima at high gas velocities. This effect becomes less and less pronounced as the packing density increases.
- All liquid velocities studied in this work show a negligible effect on \((k_ga)_{b,s}\) for all packing densities. Even though a slight decrease in \((k_ga)_{b,s}\) is very much noticeable for all gas velocities and all packing densities as the liquid velocities are increased.

The above mentioned inferences can be explained by considering the hydrodynamic characteristics of the bed. Figure 6-25 for expanded bed height and Figure 6-33 for the liquid hold up show the effect of packing density on bed expansion and liquid hold up respectively. Expanded bed height decreases as the packing density increases whereas the liquid hold up increases for a particular static bed height, packing diameter and liquid and gas velocities. Now if the bed expands less and there is more liquid holdup in the bed then naturally the free volume available for the gas to flow upwards will decrease. This decrease in flow area should
significantly increase the interstitial gas velocity which should also result in an increase in the degree of turbulence on the gas side.

This logical reasoning dictates that the increase in the interstitial gas velocity should result in an increase in the \((k_{ga})_{b,s}\). This leads to a conclusion that the higher the density of the packing higher will be the mass transfer coefficients. However, it should be kept in mind to consider the effect of packing density on the bed pressure drop. Figure 6-5 reveals this aspect. This figure shows that the pressure drop is significantly increased if either the packing density or the liquid velocity is increased. Hence the optimum packing density should be the one which gives maximum mass transfer per unit pressure drop. This is where the concept of operation mass transfer efficiency reveals its worth. This parameter will be discussed in the later part of this chapter.

### 7.4.3 Effect of packing size on \((k_{ga})_{b,s}\)

The effects of packing diameter on the volumetric gas film mass transfer coefficients are presented in Figure 7-5 to Figure 7-7. The following inferences can be made from these plots for a constant static bed height and packing density.

- \((k_{ga})_{b,s}\) increases with a decrease in packing diameter
- For a particular diameter packing the liquid velocity has no effect on \((k_{ga})_{b,s}\)
- For a particular diameter packing \((k_{ga})_{b,s}\) increases with an increase in the gas velocity

The above mentioned effect of packing diameter is very much expected. Again, the hydrodynamic characteristics of the TCA bed help in explaining this mass transfer behavior. If Figure 6-24 and Figure 6-34 are reviewed again then one can conclude that the TCA bed made of lower diameter ball expands more as compared to the larger diameter balls. Similarly the bed composed of smaller diameter balls have a slightly greater liquid hold up as compared to the larger diameter balls. This suggests that for the same expanded bed height the more free volume should be available for the gas to flow upwards in case of small diameter packing as compared to the larger diameter packing. This should in fact suggest that there should more turbulence in a less expanded bed as compared to a more expanded bed. Moreover, this argument should also lead to a conclusion that the \((k_{ga})_{b,s}\) should be more for a less expanded bed i.e. large diameter packing as compared to the more expanded bed of small diameter packing.

However, the results presented in the above mentioned Figure 7-5 to Figure 7-7 show a very different picture. This discrepancy can be resolved by considering the following table:
Table 7-2: No. of balls per unit height of expanded bed height at various liquid and gas velocities

<table>
<thead>
<tr>
<th>Sr. No</th>
<th>d_p m</th>
<th>Number of balls per meter expanded bed height, ( u_g = 2.4 \text{ m/s}, u_l = 0.008 \text{ m/s} )</th>
<th>Number of balls per meter expanded bed height, ( u_g = 3.0 \text{ m/s}, u_l = 0.008 \text{ m/s} )</th>
<th>Number of balls per meter expanded bed height, ( u_g = 3.6 \text{ m/s}, u_l = 0.008 \text{ m/s} )</th>
</tr>
</thead>
<tbody>
<tr>
<td>1</td>
<td>0.025</td>
<td>6770</td>
<td>5960</td>
<td>5100</td>
</tr>
<tr>
<td>2</td>
<td>0.038</td>
<td>2270</td>
<td>2120</td>
<td>1870</td>
</tr>
</tbody>
</table>

The number of balls per unit expanded bed height is significantly more for 25 mm packing as compared to the 38 mm packing. As the number of packing increases per unit expanded bed height logic suggests that the number of collisions occurring between the balls will be much larger in case of 25 mm packing as compared to the 38 mm packing. This increased number of collisions should contribute significantly towards the increased degree of turbulence on the gas as well as liquid side. However, it is the increased turbulence on the gas side that should contribute positively towards the increase in \((k_g a)_{b,s}\). Hence one can conclude that the smaller diameter packing should result a higher \((k_g a)_{b,s}\) in a TCA bed as compared to the larger diameter packing.

Figure 6-6 reveals that the pressure drop is almost independent of the packing diameter hence the choice of packing for a particular set of gas and liquid velocities, static bed height and packing density should not be dictated by the pressure drop conditions but rather by the mass transfer performance. Hence for such a set of process conditions the smaller diameter packing should provide more mass transfer.

7.4.4 Effect of static bed height on \((k_g a)_{b,s}\)

The effects of the static bed height on the volumetric gas film mass transfer coefficients of the TCA bed i.e.\((k_g a)_{b,s}\) have been plotted from Figure 7-8 to Figure 7-11. The following interpretations can be made from these plots.

- \((k_g a)_{b,s}\) increases as the static bed height decreases for a given set of liquid and gas velocities, packing density and packing diameter
- This increase in \((k_g a)_{b,s}\) behavior is valid for both type 1 and 2 fluidizations
- The increasing liquid velocity has no effect on \((k_g a)_{b,s}\) for a given set of gas velocity, static bed height, packing diameter and packing density
- This behavior is also valid for both type 1 and 2 fluidizations
- \((k_g a)_{b,s}\) passes through a maxima with increasing gas velocity for both type 1 and 2 fluidizations. However, the gas velocity at which this maximum occurs is different for
the two fluidizations types. For type 1 this occurs around a gas velocity of 2.6 m/s whereas for type two it occurs at around 3.1 m/s

- The magnitude of mass transfer and mass transfer coefficients is larger for type 2 as compared to type 1
- The performance of static bed height of 0.25 and 0.31 m are not very different from each other, however, the performance of static bed height of 0.15m is significantly better and becomes even more pronounced as the gas velocity increases.

The mass transfer coefficient is more for a shallow bed as compared to the deeper bed. However if we calculate the mass transfer occurring at a particular liquid velocity (0.006 m/s) and a gas velocity (2.5 m/s) by taking the product of \((k_{g}a)_{b,s}\) and volume of the static bed then the following table results.

<table>
<thead>
<tr>
<th>Sr. No</th>
<th>(H_0) m</th>
<th>((k_{g}S)_{b,s})</th>
<th>Mass transfer Kmoles/(m(^3)-s atm)</th>
</tr>
</thead>
<tbody>
<tr>
<td>1</td>
<td>0.15</td>
<td>0.66</td>
<td>0.015</td>
</tr>
<tr>
<td>2</td>
<td>0.25</td>
<td>0.465</td>
<td>0.018</td>
</tr>
<tr>
<td>3</td>
<td>0.35</td>
<td>0.375</td>
<td>0.021</td>
</tr>
</tbody>
</table>

Now if one tries to see how much the first 15 cm of the bed in a bed of 25 or 35 cm height would perform as compared to the static bed height of only 15 cm in terms of mass transfer, then the following data can be produced.

<table>
<thead>
<tr>
<th>Sr. No</th>
<th>(H_0) m</th>
<th>Mass transfer (Kmoles/s atm)</th>
<th>Mass transfer contribution in terms of first 0.15 m bed height (Kmoles/s atm)</th>
<th>% age contribution as compared to 0.15 cm bed height</th>
</tr>
</thead>
<tbody>
<tr>
<td>1</td>
<td>0.15</td>
<td>0.015527</td>
<td>0.015527</td>
<td>100</td>
</tr>
<tr>
<td>2</td>
<td>0.25</td>
<td>0.018234</td>
<td>0.01094</td>
<td>70.5</td>
</tr>
<tr>
<td>3</td>
<td>0.35</td>
<td>0.020584</td>
<td>0.0088217</td>
<td>57</td>
</tr>
</tbody>
</table>

This table shows that the first 15 cm of a 25 cm bed contributes only 70.5 % as compared to a static bed height of 15 cm. Similarly if the bed height is further increased to 35 cm then the contribution of the first 15 cm of the bed reduces to 57% as compared to a static bed height of 15 cm. This means as the static bed height is increased the total mass transfer does increase but the mass transfer per unit static bed height decrease. The reason for this decrease lies in the hydrodynamic performance of the bed. When a bed of only 15 cm height is fluidized then one can expect that the area closer to the support grid will have lesser number
of balls hence the chance of liquid being splashed and broken into droplets will be much larger as compared to a deeper bed, due to the presence of relatively larger number of packing. This behavior was also observed when these experiments were carried out.

Therefore one can conclude that in shallow beds for a particular set of operating conditions the contribution of mass transfer due to this splashing and breaking of liquid at the support grid should be more as compared to a deeper bed. However one can reason that the mass transfer contribution due to the bed only would not be significantly different for different heights as the bed expansion characteristics for different static bed heights are not very different (Figure 6-26 and Figure 6-27). This behavior of increased mass transfer for shallow beds has also been reported in the very limited literature available. One can further predict that this contribution of support grid will be increased as the bed becomes shallower and shallower. However, one can also predict there may be a maxima as the bed becomes more and more shallow. Hence, there may be a particular minimum static bed height where the mass transfer will be maximum (due to the maximum contribution of the support grid) and any further decrease in static bed height will significantly decrease the mass transfer of the bed only even though the mass transfer contribution of support grid would be maximum.

7.5 Operational efficiencies of Bed, $\eta_{op,b}$

The operational mass transfer efficiency of the bed as defined above in Table 7-1 has been calculated from the $(k_g \alpha)_{c,s}$ (for the column) and $(k_g \alpha)_p$ (for the plenum section). This has been done to exclude the effect of the plenum section. These operational mass transfer efficiencies for the plenum and column are presented in Figure 9-13 to Figure 7-14 whereas, these efficiencies of the bed only are presented in Figure 7-15 to Figure 7-26.

7.5.1 The effect of gas and liquid velocities

Figure 7-15 to Figure 7-18 show the effect of liquid and gas velocities on the $\eta_{op,b}$ for both type 1 and 2 fluidizations and for packing diameter of 25, 38 and 45 mm. The following inferences can be made for the operational mass transfer efficiency of the TCA bed.

- The $\eta_{op,b}$ of the bed decreases as the gas velocity increases for both type 1 and 2 fluidizations for all packing diameter.
• For type 1 fluidization $\eta_{op,b}$ also decreases significantly with increasing liquid velocity. This decreasing effect becomes less significant as the gas velocity increases. This effect also decreases as the packing density increases.

• For type 2 fluidization the effect of liquid velocity is almost insignificant. However, the effect of gas velocity is quite significant.

As by definition the operational mass transfer efficiency is the ratio of the magnitude of mass transfer achieved per unit work expanded in fluidizing the bed. In-fact all mass transfer process required that the liquid and gas phase should be so interacted to achieve maximum mass transfer contact area as well as maximum mass transfer coefficients. In order to achieve this, the two phases need to be well intermixed at the expense of energy in terms of pressure drop through the mass transfer device. AS TCA is also one of the concepts to achieve both these targets of increased contact area and increased mass transfer coefficients. Since gas velocities through TCA are quite high hence one can expect considerable pressure drop.

From operation point of view one should hope to achieve maximum mass transfer in TCA at the expanse of minimum pressure drop. Hence ideally the $\eta_{op,b}$ should be as high as possible for a TCA. Seeing the behavior as shown in the above mentioned graphs one can conclude that $\eta_{op,b}$ is maximum at low gas and liquid velocities for type 1 fluidization. However, for type 2 fluidization $\eta_{op,b}$ is again maximum at low gas velocities but it is independent of the liquid velocity.

One should also point out here that one of the biggest advantages of TCA is their high throughput in terms of gas flow rates which in turn reduces the equipment size. However, if the above mentioned results are considered then the best efficiency occurs at low gas velocities and hence low throughput, therefore large column diameters. If on the other hand large gas throughputs are to be ensured then one has to pay in terms of high pressure drop and possible large column heights.

Finally one can conclude that the optimum flow rate should be worked out by optimizing the equipment size in terms of its capital cost and its operational cost. One should opt for an $\eta_{op}$ that gives a minimum equipment capital cost.

The above graphs also show some effects of the packing density and packing size on operational efficiency. These are now further presented and discussed

### 7.5.2 Effect of packing density

The effects of packing density are shown in Figure 7-25 and Figure 7-26 for various gas and liquid velocities. The graphs show that higher the packing density lower is the $\eta_{op,b}$. 
Similarly the $\eta_{op,b}$ also decreases with increasing gas velocity as well as liquid velocity. It is quite easy to comprehend this behavior as all the three parameters i.e. gas velocity, liquid velocity and packing density increase the pressure drop if one or more of them are increased. Again the maximum mass transfer efficiency occurs at lowest gas velocity, lowest liquid velocity and lowest packing density. The above discussion regarding the optimum choice is very much valid here also.

7.5.3 Effect of packing size

The effects of packing diameter are shown in Figure 7-19 and Figure 7-20. With increasing gas velocity the $\eta_{op,b}$ decreases but is independent of liquid velocity. However, $\eta_{op}$ is more for smaller diameter packing as compared to the large diameter packing. Similarly $\eta_{op,b}$ drops rapidly for the small diameter packing as compared to the large diameter packing. This behavior is valid for type-2 fluidization. Moreover, $\eta_{op,b}$ becomes same for all packings studied for gas velocity equal to more than 3.0 m/s. $\eta_{op,b}$ is a combination of mass transfer and pressure drop hence the combined behavior will be same as these two individual parameter vary with increasing gas and liquid velocity as the packing diameter increases.

7.5.4 Effect of static bed height

Consider Figure 6-7, Figure 6-8, Figure 7-8 to Figure 7-11 showing the behavior of pressure drop and (kga)$_{b,s}$ changing with increasing gas velocity for increasing static bed heights. These figures show these behaviors for both type-1,2 fluidizations. As the pressure drops are higher for larger static bed heights and (kga)$_{b}$’s are more for smaller static bed heights, therefore one can very easily predict that $\eta_{op,b}$ will be higher for small static bed heights. This is exactly represented in Figure 7-21 to Figure 7.24. Since pressure initially increases with increasing gas velocity till minimum fluidization velocity is reached. After this pressure drop becomes almost independent of gas velocity. On the other hand kga increases with increasing gas velocity for both type-1 and 2 fluidizations. $\eta_{op,b}$ behaves in a similar manner, being highest at lowest gas and liquid velocities for the smallest static bed height and vice versa for largest static bed height. It should be remembered that fully developed fluidization occurs beyond a gas velocity of 1.8 m/s. The $\eta_{op,b}$ curves with increasing gas velocity are much more flat for type 2 fluidization as compared to type-1 fluidization. Therefore $\eta_{op,b}$ will almost remain almost same with increasing gas velocity for type-2 fluidization as compared to type-1 fluidization. Hence more room for gas velocity variation with consistent mass transfer performance in case of type-2 fluidization.
7.5.5 Correlation development

Using dimensional analysis correlation for volumetric gas film mass transfer coefficients developed using the experimental mass transfer data of this study. The two correlations developed and presented below are for type-1 and 2.

Correlation for type-1

$$
(k_g a)_{b,s} = 0.006(u_g D_c)^{-1} \bar{Re}_g^{-0.07} F_{Re}^{1.39} \left( \frac{\rho_p}{\rho_g} \right)^{0.53} \left( \frac{H_o}{D_c} \right)^{-0.35}
$$

MRD=8.4 %

Correlation for type-2

$$
(k_g a)_{b,s} = 4.24 \times 10^{-7} (u_g D_c)^{-3} \bar{Re}_g^{-0.41} F_{Re}^{1.39} \left( \frac{\rho_p}{\rho_g} \right)^{0.53} \left( \frac{H_o}{D_c} \right)^{-0.35}
$$

MRD=9.1 %

7.6 Comparison with literature

It is not possible to compare the present results exactly with the literature data as the conditions used in literature differ from each other and with present study. However, an attempt has been made to compare the results with literature data. Only Inayat (1995) and Guerriere et al. (1995) have subtracted the mass transferred in plenum from column to get the mass transferred in the bed. Other investigators did not mention about mass transferred in plenum. Results in present study are consistent in trend with respect to gas velocities with those of Inayat(1995), Guerriere et al.(1995) and Barile et al.(1975) and with respect to liquid velocities with those of Inayat (1995), Guerriere et al.(1995) and El-Dessouky (1993). Only Guerriere et al.(1995) has mentioned about the maxima of $k_g a$ with increase in gas velocity after which the values of $k_g a$ decrease with increase in gas velocity. Results of present study can be compared quantitatively in Figure 7-27 and Figure 7-28. The variables for data of Figure 7-27 are mentioned in Table 7.6 while those of Figure 7-28 are mentioned in Table 7.1. For data in Table 7.6 and Table 7.7, liquid and gas velocities are taken which are close to 0.006 m/s and 2.6 m/s respectively. Results of present study are higher than those of Guerriere et al.(1995), El-Dessouky (1993) and NH3 absorption data of Douglas (1964) while it is lower than Inaya (1995), Barile et al.(1975) and Dehumidification data of Douglas (1964). Values of $k_g a$ by Barile et al.(1975) and by Douglas(1964) for dehumidification are of the same order as of $(k_g a)_{c,s}$ by present study. Experimental values of $k_g a$ by El-Dessouky (1993) are highly dependent on water inlet temperature. The values from data by El-
Dessouky (1993) mentioned in Table 7.7 are for inlet water temperature 324 K when experiments for water cooling were carried for determination of tower characteristic.

The correlations developed above are presented in Figure 7-29 and Figure 7-30 to see how the experimental data of other workers fit if their data is also represented in a dimensionless form. This attempt is again to see why there are discrepancies in the kga data reported in literature. Before doing a comparison it is mentioned to remember which of the earlier workers have considered the contribution of the kga in the plenum while determining the kga of the TCA bed. One should expect that the kga reported in literature who has not considered the contribution of plenum should be higher as compared to those who have considered its contribution.

Figure 7-29 shows that data of Inayat 1995 and Barile 1975 for two different static bed heights. Inayat used a H0/Dc ratio less than 10 for his studies and applied the plenum correction therefore one could expect the wall effects in his data. Yet his data is within ±30 band. Barile 1975 used a 28.6 cm column hence one should not expect the wall effects, however he used two static bed heights, one less than 1 and the other equal to 1. His most of the data is within ±30 band, yet most of his data is on the higher side as he has not applied the plenum correction.

Figure 7-30 shows the performance of the correlation for type-2 fluidization data. For type-2 fluidization the data of the current study has a relatively larger spread as compared to the type-1 fluidization. Hence in this case a ±40 % band is placed for the comparison with other workers data. Considering the spread of the experimental data of this study one could say this fit of others data is not that bad. However, these two plots suggest that the mass transfer characteristics of the TCA can be represented in a dimensionless form. The spread of the data reported in literature also suggests that more detailed study in this case is required.
7.7 Graphs for mass transfer in TCA

Figure 7-1: Effect of liquid and gas velocities on \((k_g a)_p\) in plenum

Figure 7-2: Effect of liquid and gas velocities on \((k_g a)_{b,s}\) for density 180 kg/m³
Figure 7-3: Effect of liquid and gas velocities on $(kga)_{b,s}$ for density 270 kg/m$^3$

Figure 7-4: Effect of liquid and gas velocities on $(kga)_{b,s}$ for density 354 kg/m$^3$, dia 38 mm
Figure 7-5: Effect of liquid and gas velocities on \((k_{ga})_{h,s}\) for density 354 kg/m\(^3\), dia 25 mm.

Figure 7-6: Effect of packing size on \((k_{ga})_{h,s}\) for density 354 kg/m\(^3\) at different liquid velocities.
Figure 7-7: Effect of packing size on \((k_{ga})_{h,a}\) for density 354 kg/m\(^3\) at different gas velocities

Figure 7-8: Effect of static bed height on \((k_{ga})_{h,a}\) at different gas velocities for density 180 kg/m\(^3\)
Figure 7-9: Effect of static bed height on \((k_a g)_{b, s}\) at different liquid velocities for density 180 kg/m³.

Figure 7-10: Effect of static bed height on \((k_a g)_{b, s}\) for density 354 kg/m³ for different gas velocities.
Figure 7-11: Effect of static bed height on \( (k_a)_{b,s} \) for density 354 kg/m\(^3\) at different liquid velocities.

Figure 7-12: Effect of packing density on \( (k_a)_{b,s} \) at different liquid velocities.
Figure 7-13: Effect of packing density on \((k_g a)_{h,s}\) at different gas velocities

Figure 7-14: Effect of liquid and gas velocities on \(\eta_{op,p}\) for plenum
Figure 7-15: Effect of liquid and gas velocities on $\eta_{op,b}$ for density 180 kg/m$^3$.

Figure 7-16: Effect of liquid and gas velocities on $\eta_{op,b}$ for density 270 kg/m$^3$. 
Figure 7-17: Effect of liquid and gas velocities on $\eta_{\text{op,b}}$ for density 354 kg/m$^3$, dia 38 mm

Figure 7-18: Effect of liquid and gas velocities on $\eta_{\text{op,b}}$ for density 354 kg/m$^3$, dia 25 mm
Figure 7-19: Effect of packing size on $\eta_{op,b}$ for density 354 kg/m$^3$ at different gas velocities

Figure 7-20: Effect of packing size on $\eta_{op,b}$ for density 354 kg/m$^3$ at different liquid velocities
Figure 7-21: Effect of static bed height on $\eta_{\text{op,b}}$ at different gas velocities for density 180 kg/m$^3$

Figure 7-22: Effect of static bed height on $\eta_{\text{op,b}}$ at different liquid velocities for density 180 kg/m$^3$
Figure 7-23: Effect of static bed height on $\eta_{op,b}$ for density 354 kg/m$^3$ for different gas velocities.

Figure 7-24: Effect of static bed height on $\eta_{op,b}$ for density 354 kg/m$^3$ at different liquid velocities.
Figure 7-25: Effect of packing density on $\eta_{op,b}$ at different gas velocities

Figure 7-26: Effect of packing density on $\eta_{op,b}$ at different liquid velocities
Table 7-5: Comparison of $k_{ga}$ with literature

<table>
<thead>
<tr>
<th>Authors</th>
<th>$u_l$ (m/s)</th>
<th>$u_s$ (m/s)</th>
<th>$f$</th>
<th>$H_0$ (m)</th>
<th>$d_p$ (mm)</th>
<th>$D_c$ (cm)</th>
<th>$\rho_p$ (kg/m$^3$)</th>
<th>$k_{ga}$ (kmoles/m$^3$-s-atm)</th>
<th>System used</th>
</tr>
</thead>
<tbody>
<tr>
<td>Present study (kga)b, s Type1</td>
<td>0.006</td>
<td>2.60</td>
<td>0.74</td>
<td>0.25</td>
<td>38</td>
<td>44.7</td>
<td>180</td>
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Figure 7-27: Literature comparison with effect of gas velocity on $k_g a$

Table 7-6: Conditions for readings of Figure 7-27

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<th>$D_c$ (cm)</th>
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<td>10.5</td>
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Figure 7-28: Literature comparison with effect of liquid velocity on $k_g a$

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<td>15, 25</td>
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<td>354</td>
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<td>0.74</td>
<td>25</td>
<td>38</td>
<td>44.7</td>
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Figure 7-29: \((k_g)_{0.5}\) from literature vs. present study correlation for type1 fluidization.

Figure 7-30: \((k_g)_{0.5}\) from literature vs. present study correlation for type2 fluidization

Table 7-8: Parameters used in graphs for the comparison of mass transfer coefficient with present study correlation
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<tr>
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<th>$D_c$</th>
<th>$d_p$</th>
<th>$f$</th>
<th>$D_c/d_p$</th>
<th>$H_0/D_c$</th>
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<td>0.019</td>
<td>0.82</td>
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<td>0.52, 1.04</td>
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</table>
Chapter 8

8 Conclusions and Future Recommendations

8.1 Conclusion

The hydrodynamic and mass transfer characteristics reported in this study and in the literature have qualitatively same trends.

Among the hydrodynamic characteristics, the vast amount of discrepancies in pressure drop are removed once the information available is classified according to the following conditions and reported in a dimensionless form.

1. The column diameter should be ≥ 15 cm
2. The column diameter to particle diameter ratio should Dc/dp ≥ 10
3. The static bed height to column diameter ratio H0/Dc should be ≤ 1
4. The grid free area should be more than 70%
5. There should a minimum static bed to where the contribution of grid free area towards the mass transfer performance of TCA is maximum yet the pressure drop is minimum.

The vast amount of discrepancies is simply because of the reason that due to so many parameters which affect the performance of TCA, it is almost impossible to represent the tower characteristics in a single equation unless the tool of dimensionless analysis is used.

The discrepancies in the liquid hold up are also removed once the above given conditions are met and necessary corrective measures are to account for the contribution of the gas and liquid distributors. Hence it can be concluded that the main reason for the discrepancies in the hydrodynamic characteristics lie in the liquid hold up studies, where in most of the literature reported the contribution of the gas and liquid distributors had been ignored.

As in the case of pressure drop the liquid hold up data can also be represented by a dimensionless correlation which may represent the vast amount of liquid holdup data reported in literature with in a ± 30% error band.

Similarly the volumetric mass transfer coefficients can also be represented by a single dimensionless correlation and for this also necessary corrective measure for the gas distribution section need to be incorporated along with the above mentioned criteria.
The operational mass transfer efficiency is a good design parameter to be considered which represents both the mass transfer performance per unit pressure drop for a TCA.

The small diameter columns data may not be used for the design of industrial scale TCA towers as the data of these small columns may not be reliable due to the wall affects.

### 8.2 Future Recommendations

1. More detailed study for gas phase transfer coefficients need to be evaluated in a pilot scale Turbulent Contact Absorber using other techniques e.g. Absorption of NH$_3$ in sulphuric acid solutions and result may be compared with the present study.

2. Beside the volumetric gas film mass transfer coefficients, the liquid film mass transfer coefficient should also be measured for a large scale TCA.

3. Comprehensive universal type correlations for the hydrodynamic and mass transfer characteristics of TCA may be developed using the detailed experimental data present in literature using dimensional analysis on the same lines as done in this study. In doing so necessary correction factors may be applied for the cases where various ratios such as ($H_0/D_c$, $D_c/d_p$ etc) are not as per the recommended values.

4. Mist eliminator may be designed for minimum entrainment of liquid with the gas

5. To study the effect of grid free area on mass transfer in a turbulent contact absorber.
References


Linek, V. (1981). "Chemical engineering use of catalyzed sulfite oxidation kinetics for the
determination of mass transfer chatacteristics of gas-liquid contactors." Chemical
scrubbing data from pilot plant spray and TCA scrubber." Ind Eng Chem Process
scrubber." The Chemical Engineering Journal 22:
Miconnet, M., P. Guigon, et al. (1982). "the scrubbing of acid gases in columns with fixed
or mobile packings." International Chemical Engineering 22(1): 133-141.
towers with fluidized packings." Canadian Journal of Chemical Engineering 50:
595-601.
Paterson, A. H. J. (1979). Liquid holdup and Gas Absorption in Turbulent Bed Contactors,
University of Cambridge. Ph.D.
contactors by a tracer technique." Canadian Journal of Chemical Engineering. 65:
10-17.
density packing particles of different shapes." Canadian Journal of Chemical
Contact Absorber." International Journal of Chemical Engineering and Applications
Shackley, I. (2000). The Design and Operation of Turbulent Bed Contactors. Department of
Chemical and Process Engineering University of Sheffield. Ph.D: 249.
columns. University of UMSIT. Ph.D.
Soundarajan, K. and K. Krishnaiah (1994). "Hydrodynamics of Turbulent Bed Contactor -


Appendix

9 Figures of mass transfer in the whole column including plenum

Figure 9-1: Effect of liquid and gas velocities on \((k_a)_{c,s}\) for density 180 kg/m³

Figure 9-2: Effect of liquid and gas velocities on \((k_a)_{c,s}\) for density 270 kg/m³
Figure 9-3: Effect of liquid and gas velocities on \((k_a)_{c,s}\) for density 354 kg/m³, dia 38 mm

Figure 9-4: Effect of liquid and gas velocities on \((k_a)_{c,s}\) for density 354 kg/m³, dia 25 mm
Figure 9-5: Effect of packing size on \((k_a a)_{\text{cs}}\) for density 354 kg/m\(^3\) at different gas velocities

Figure 9-6: Effect of packing size on \((k_a a)_{\text{cs}}\) for density 354 kg/m\(^3\) at different liquid velocities
Figure 9-7: Effect of static bed height on \((k_{ga})_{cs}\) at different gas velocities for density 180 kg/m³

Figure 9-8: Effect of static bed height on \((k_{ga})_{cs}\) at different liquid velocities for density 180 kg/m³
Figure 9-9: Effect of static bed height on $(k_g a)_{c,s}$ for density 354 kg/m$^3$ for different gas velocities.

Figure 9-10: Effect of static bed height on $(k_g a)_{c,s}$ for density 354 kg/m$^3$ at different liquid velocities.
Figure 9-11: Effect of packing density on $(k_a)_{c,s}$ at different gas velocities

Figure 9-12: Effect of packing density on $(k_a)_{c,s}$ at different liquid velocities
Figure 9-13: Effect of liquid and gas velocities on $\eta_{\text{op,c}}$ for density 180 kg/m$^3$

Figure 9-14: Effect of liquid and gas velocities on $\eta_{\text{op,c}}$ for density 270 kg/m$^3$
Figure 9-15: Effect of liquid and gas velocities on $\eta_{op,c}$ for density 354 kg/m$^3$, dia 38 mm

Figure 9-16: Effect of liquid and gas velocities on $\eta_{op,c}$ for density 354 kg/m$^3$, dia 25 mm
Figure 9-17: Effect of packing size on $\eta_{\text{op,c}}$ for density 354 kg/m$^3$ at different gas velocities

Figure 9-18: Effect of packing size on $\eta_{\text{op,c}}$ for density 354 kg/m$^3$ at different liquid velocities
Figure 9-19: Effect of static bed height on $\eta_{\text{op,e}}$ at different gas velocities for density 180 kg/m$^3$.

Figure 9-20: Effect of static bed height on $\eta_{\text{op,e}}$ at different liquid velocities for density 180 kg/m$^3$. 

143
Figure 9-21: Effect of static bed height on $\eta_{op,c}$ for density $354 \text{ kg/m}^3$ for different gas velocities.

Figure 9-22: Effect of static bed height on $\eta_{op,c}$ for density $354 \text{ kg/m}^3$ at different liquid velocities.
Figure 9-23: Effect of packing density on $\eta_{op,c}$ at different gas velocities

Figure 9-24: Effect of packing density on $\eta_{op,c}$ at different liquid velocities
Sample calculations for mass transfer experiments

**k_{ga} Sample Calculations**

Diameter of packing= 38 mm
Density of packing=180 kg/m³
Air velocity, \( u_g = 2.61 \) m/s
Water velocity, \( u_l = 0.006 \) m/s
Air mass velocity, \( G' = 3.23 \) kg/m²-s
Static bed height=0.25 m
Expanded bed height=0.42 m
Plenum height for mass transfer above liquid level=1.07 m
Air inlet temperature=26.2 °C
Air outlet temperature= 16.5 °C
Water outlet temperature= 15.1 °C
Number of transfer units for column as from equation 7.2
\( N_{G,c} = \ln \frac{26.2-15.1}{16.5-15.1} = 2.07 \)

From equation 7.1
\[
(k_{ga})_{c,p} = \frac{3.23 \times 2.07 \times 101300}{1.2 \times (0.42 + 1.07) \times 8314 \times (\frac{26.2 + 16.5}{2} + 273)}
\]

\( = 0.15 \) k-moles/(m³-s-atm.)

Similarly no. of transfer units for plenum were determined from equation 7.2 in separate experiments for same water and air flow rates. Since it was not possible to achieve exactly same velocities so \( N_{G,p} \) was interpolated for each velocity for same liquid flow rate. In present case, \( N_{G,p} \) at \( u_l = 0.006 \) m/s and \( u_g = 1.94 \) was 1.28 and at \( u_l = 0.006 \) m/s and \( u_g = 2.69 \) was 1.32. So for desired air velocity 2.61 m/s, it can be interpolated to get the value of \( N_{G,p} \) as follows
\( N_{G,p} = (2.61-1.94) \times (1.32-0.128)/(2.69-1.94)+1.28 = 1.316 \) k-moles/(m³-s-atm)

From equation 7.3 & 7.9
\( (k_{ga})_{b,s} = 3.23 \times (2.07-1.316)/(0.25 \times 1.013 \times 29) = 0.33 \) k-moles/(m³-s-atm)
\( (k_{ga})_{b,e} = 0.33 \times 0.25/0.42 = 0.196 \) kmoles/(m³-s-atm)

**\( \eta_{op} \) Sample Calculations**

Operational efficiency of column and bed can be calculated from equations 7.6 and 7.7 as follows.

For the same parameters as in \( k_{ga} \) sample calculations,
H_c = 0.42 m, H_p = 1.07 m, u_g = 2.6 m/s,
Pressure drop across plenum, $\Delta p_p = 632 \text{ N/m}^2$
Pressure drop across bed, $\Delta p_b = 499 \text{ N/m}^2$
Pressure drop across column, $\Delta p_c = 632 + 499 = 1131 \text{ N/m}^2$
As has been calculated in previous section,

$(kga)_{c, p, c} = 0.15 \text{ kmoles/} (\text{m}^3 \cdot \text{s} \cdot \text{atm})$,

$(kga)_{b, c} = 0.196 \text{ kmoles/} (\text{m}^3 \cdot \text{s} \cdot \text{atm})$

So, from equations 7.6 & 7.7

$\eta_{op, c} = 0.15 \times (0.42 + 1.07) / (1131 \times 2.6) = 7.6 \times 10^{-5} \text{ kmoles/} (\text{N} \cdot \text{m})$

$\eta_{op, b} = 0.196 \times 0.42 / (499 \times 2.6) = 6.3 \times 10^{-5} \text{ kmoles/} (\text{N} \cdot \text{m})$