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Heat transfer to a stationary and moving sphere immersed in a fluidized bed

Chetan Jitendra Desai

Iowa State University

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Heat transfer to a stationary and moving sphere immersed in a fluidized bed

Desai, Chetan Jitendra, Ph.D.

Iowa State University, 1989
Heat transfer to a stationary and moving sphere immersed in a fluidized bed

by

Chetan Jitendra Desai

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1989
# TABLE OF CONTENTS

NOMENCLATURE .......................................................... xxxii

1 INTRODUCTION ......................................................... 1
   1.1 Liquid Like Behavior of a Fluidized Bed ..................... 2
   1.2 Brief History of Fluidization .................................. 3
   1.3 Industrial Applications of Fluidized Beds .................. 4
   1.4 Heat Transfer in Fluidized Beds .............................. 4
       1.4.1 Particle-to-particle heat transfer ..................... 4
       1.4.2 Particle-to-gas heat transfer ....................... 6
       1.4.3 Bed-to-surface heat transfer ....................... 6
   1.5 Present Investigation ....................................... 8
   1.6 Significance of Present Investigation ..................... 10
   1.7 Literature Survey ............................................ 11
       1.7.1 Superficial gas velocity ............................. 12
       1.7.2 Particle diameter ................................... 13
       1.7.3 Size of the heat transfer surface ..................... 14
       1.7.4 Static pressure of the bed .......................... 16
       1.7.5 Particle sphericity ................................ 17
1.7.6 Position of the heat transfer surface .................. 18
1.7.7 Gas conductivity ........................................ 18
1.7.8 Baffled fluidized bed .................................. 19
1.7.9 Bed diameter ............................................. 19
1.7.10 Bed temperature ....................................... 20
1.7.11 Heat transfer surface temperature .................. 21
1.7.12 Particle thermophysical properties .................. 21
1.7.13 Local heat transfer coefficient on a tube or sphere perimeter ........ 22
1.7.14 Orientation of heat transfer surface ................. 25
1.7.15 Moisture ............................................... 25
1.7.16 Mixed particle size ..................................... 25
1.7.17 Packed bed height ..................................... 26
1.7.18 Gas convective component ............................ 27
1.7.19 Vibrated bed ........................................... 28
1.7.20 Vibrating plate ......................................... 31
1.7.21 Freely moving sphere .................................. 32

1.8 Theoretical Models ...................................... 35
1.8.1 Conductive heat transfer through the gas boundary layer .. 35
1.8.2 Unsteady heat conduction by particles ................. 37
1.8.3 Packet theory ........................................... 41
1.8.4 Packet model including contact resistance ............. 46
1.8.5 Residence time study .................................... 55

2 EXPERIMENTAL SET UP AND PROCEDURE ............... 57
<table>
<thead>
<tr>
<th>Section</th>
<th>Title</th>
<th>Page</th>
</tr>
</thead>
<tbody>
<tr>
<td>2.1</td>
<td>Introduction</td>
<td>57</td>
</tr>
<tr>
<td>2.2</td>
<td>Experimental Setup</td>
<td>57</td>
</tr>
<tr>
<td>2.2.1</td>
<td>Fluidized bed</td>
<td>58</td>
</tr>
<tr>
<td>2.2.2</td>
<td>Flowmeters</td>
<td>61</td>
</tr>
<tr>
<td>2.2.3</td>
<td>Heater Cup</td>
<td>61</td>
</tr>
<tr>
<td>2.2.4</td>
<td>Sphere Driving Mechanisms</td>
<td>64</td>
</tr>
<tr>
<td>2.2.5</td>
<td>Copper sphere</td>
<td>66</td>
</tr>
<tr>
<td>2.2.6</td>
<td>Fluidized bed material</td>
<td>67</td>
</tr>
<tr>
<td>2.2.7</td>
<td>Data acquisition system</td>
<td>67</td>
</tr>
<tr>
<td>2.3</td>
<td>Experimental Procedure</td>
<td>71</td>
</tr>
<tr>
<td>2.3.1</td>
<td>Linearly downward moving sphere</td>
<td>74</td>
</tr>
<tr>
<td>2.3.2</td>
<td>Oscillating sphere</td>
<td>76</td>
</tr>
<tr>
<td>2.4</td>
<td>Method of Evaluation of Average Heat Transfer Coefficient</td>
<td>76</td>
</tr>
<tr>
<td>3</td>
<td>RESULTS AND DISCUSSION</td>
<td>84</td>
</tr>
<tr>
<td>3.1</td>
<td>Minimum Fluidization Velocity and Classification of Glass Powder</td>
<td>84</td>
</tr>
<tr>
<td>3.2</td>
<td>Stationary Sphere</td>
<td>94</td>
</tr>
<tr>
<td>3.2.1</td>
<td>Role of Particle and Gas Convective Heat Transfer</td>
<td>107</td>
</tr>
<tr>
<td>3.2.2</td>
<td>Heat transfer coefficient as a function of average sphere temperature</td>
<td>111</td>
</tr>
<tr>
<td>3.3</td>
<td>Heat Transfer from Linearly Downward Moving Sphere</td>
<td>118</td>
</tr>
<tr>
<td>3.3.1</td>
<td>Industrial figure of merit</td>
<td>133</td>
</tr>
<tr>
<td>3.4</td>
<td>Heat Transfer from Oscillating Spheres</td>
<td>135</td>
</tr>
<tr>
<td>3.4.1</td>
<td>Comparison with linearly downward moving sphere</td>
<td>139</td>
</tr>
</tbody>
</table>
12 APPENDIX D: COMPUTER PROGRAMS  
13 APPENDIX C: EXPERIMENTAL DATA
# LIST OF TABLES

<table>
<thead>
<tr>
<th>Table 2.1: List of variables used in the study of heat transfer from a linearly downward moving sphere</th>
<th>75</th>
</tr>
</thead>
<tbody>
<tr>
<td>Table 2.2: List of variables used in the study of heat transfer from an oscillating sphere</td>
<td>77</td>
</tr>
<tr>
<td>Table 2.3: Maximum Biot number and response time for immersed copper sphere-fluidized bed system</td>
<td>78</td>
</tr>
<tr>
<td>Table 3.1: Maximum particle Reynolds number, minimum fluidization and terminal velocity for each glass particle-air system</td>
<td>90</td>
</tr>
<tr>
<td>Table 3.2: Summary of group property [Geldart, 1986]</td>
<td>92</td>
</tr>
<tr>
<td>Table 3.3: Average velocity of the sphere oscillating at various frequencies and amplitudes</td>
<td>137</td>
</tr>
<tr>
<td>Table 4.1: Coefficients for the equation 4.5 and corresponding range of application for all glass particle size</td>
<td>198</td>
</tr>
<tr>
<td>Table 4.2: Coefficients for equation 4.6 and the corresponding range of application for 5-44 μm glass particle system</td>
<td>204</td>
</tr>
<tr>
<td>Table 4.3: Coefficients for equation 4.6 and the corresponding range of application for 126-147 μm glass particle system</td>
<td>205</td>
</tr>
</tbody>
</table>
Table 4.4: Coefficients for equation 4.6 and the corresponding range of application for 355-420 \( \mu \text{m} \) glass particle system ........................................ 206

Table 10.1: List of rotameters used in the experiment ........................................ 271

Table 13.1: Heat transfer coefficient for stationary spheres immersed in a fluidized bed of 355 - 420\( \mu \text{m} \) glass particle ........................................ 312

Table 13.2: Heat transfer coefficient for stationary spheres immersed in a fluidized bed of 126 - 147\( \mu \text{m} \) glass particle ........................................ 313

Table 13.3: Heat transfer coefficient for stationary spheres immersed in a fluidized bed of 5 - 44\( \mu \text{m} \) glass particle ........................................ 314

Table 13.4: Heat transfer coefficient for linearly downward moving 1.0 cm diameter sphere immersed in a fluidized bed of 355 - 420\( \mu \text{m} \) glass particle ........................................ 315

Table 13.5: Heat transfer coefficient for linearly downward moving 1.4 cm diameter sphere immersed in a fluidized bed of 355 - 420\( \mu \text{m} \) glass particle ........................................ 316

Table 13.6: Heat transfer coefficient for linearly downward moving 2.0 cm diameter sphere immersed in a fluidized bed of 355 - 420\( \mu \text{m} \) glass particle ........................................ 317

Table 13.7: Heat transfer coefficient for linearly downward moving 1.0 cm diameter sphere immersed in a fluidized bed of 126 - 147\( \mu \text{m} \) glass particle ........................................ 318
Table 13.8: Heat transfer coefficient for linearly downward moving 1.4 cm diameter sphere immersed in a fluidized bed of 126 – 147μm glass particle ................. 319

Table 13.9: Heat transfer coefficient for linearly downward moving 2.0 cm diameter sphere immersed in a fluidized bed of 126 – 147μm glass particle ................. 320

Table 13.10: Heat transfer coefficient for linearly downward moving 1.0 cm diameter sphere immersed in a fluidized bed of 5 – 44μm glass particle .................. 321

Table 13.11: Heat transfer coefficient for linearly downward moving 1.4 cm diameter sphere immersed in a fluidized bed of 5 – 44μm glass particle .................. 322

Table 13.12: Heat transfer coefficient for linearly downward moving 2.0 cm diameter sphere immersed in a fluidized bed of 5 – 44μm glass particle .................. 323

Table 13.13: Heat transfer coefficient for a copper sphere of 1.0 cm diameter, oscillating at constant peak-to-peak amplitude of 6.9 cm immersed in a fluidized bed of 5 – 44μm glass particle . 324

Table 13.14: Heat transfer coefficient for a copper sphere of 1.0 cm diameter, oscillating at constant peak-to-peak amplitude of 4.0 cm immersed in a fluidized bed of 5 – 44μm glass particle . 325
<table>
<thead>
<tr>
<th>Table</th>
<th>Description</th>
</tr>
</thead>
<tbody>
<tr>
<td>13.15</td>
<td>Heat transfer coefficient for a copper sphere of 1.0 cm diameter, oscillating at constant peak-to-peak amplitude of 1.8 cm immersed in a fluidized bed of 5 – 44μm glass particle.</td>
</tr>
<tr>
<td>13.16</td>
<td>Heat transfer coefficient for a copper sphere of 1.4 cm diameter, oscillating at constant peak-to-peak amplitude of 6.9 cm immersed in a fluidized bed of 5 – 44μm glass particle.</td>
</tr>
<tr>
<td>13.17</td>
<td>Heat transfer coefficient for a copper sphere of 1.4 cm diameter, oscillating at constant peak-to-peak amplitude of 4.0 cm immersed in a fluidized bed of 5 – 44μm glass particle.</td>
</tr>
<tr>
<td>13.18</td>
<td>Heat transfer coefficient for a copper sphere of 1.4 cm diameter, oscillating at constant peak-to-peak amplitude of 1.8 cm immersed in a fluidized bed of 5 – 44μm glass particle.</td>
</tr>
<tr>
<td>13.19</td>
<td>Heat transfer coefficient for a copper sphere of 2.0 cm diameter, oscillating at constant peak-to-peak amplitude of 6.9 cm immersed in a fluidized bed of 5 – 44μm glass particle.</td>
</tr>
<tr>
<td>13.20</td>
<td>Heat transfer coefficient for a copper sphere of 2.0 cm diameter, oscillating at constant peak-to-peak amplitude of 4.0 cm immersed in a fluidized bed of 5 – 44μm glass particle.</td>
</tr>
<tr>
<td>13.21</td>
<td>Heat transfer coefficient for a copper sphere of 2.0 cm diameter, oscillating at constant peak-to-peak amplitude of 1.8 cm immersed in a fluidized bed of 5 – 44μm glass particle.</td>
</tr>
</tbody>
</table>
Table 13.22: Heat transfer coefficient for a copper sphere of 1.0 cm diameter, oscillating at constant peak-to-peak amplitude of 6.9 cm immersed in a fluidized bed of 126 – 147μm glass particle 333

Table 13.23: Heat transfer coefficient for a copper sphere of 1.0 cm diameter, oscillating at constant peak-to-peak amplitude of 4.0 cm immersed in a fluidized bed of 126 – 147μm glass particle 334

Table 13.24: Heat transfer coefficient for a copper sphere of 1.0 cm diameter, oscillating at constant peak-to-peak amplitude of 1.8 cm immersed in a fluidized bed of 126 – 147μm glass particle 335

Table 13.25: Heat transfer coefficient for a copper sphere of 1.4 cm diameter, oscillating at constant peak-to-peak amplitude of 6.9 cm immersed in a fluidized bed of 126 – 147μm glass particle 336

Table 13.26: Heat transfer coefficient for a copper sphere of 1.4 cm diameter, oscillating at constant peak-to-peak amplitude of 4.0 cm immersed in a fluidized bed of 126 – 147μm glass particle 337

Table 13.27: Heat transfer coefficient for a copper sphere of 1.4 cm diameter, oscillating at constant peak-to-peak amplitude of 1.8 cm immersed in a fluidized bed of 126 – 147μm glass particle 338

Table 13.28: Heat transfer coefficient for a copper sphere of 2.0 cm diameter, oscillating at constant peak-to-peak amplitude of 6.9 cm immersed in a fluidized bed of 126 – 147μm glass particle 339
Table 13.29: Heat transfer coefficient for a copper sphere of 2.0 cm diameter, oscillating at constant peak-to-peak amplitude of 4.0 cm immersed in a fluidized bed of 126 – 147μm glass particle 340

Table 13.30: Heat transfer coefficient for a copper sphere of 2.0 cm diameter, oscillating at constant peak-to-peak amplitude of 1.8 cm immersed in a fluidized bed of 126 – 147μm glass particle 341

Table 13.31: Heat transfer coefficient for a copper sphere of 1.0 cm diameter, oscillating at constant peak-to-peak amplitude of 6.9 cm immersed in a fluidized bed of 355 – 420μm glass particle 342

Table 13.32: Heat transfer coefficient for a copper sphere of 1.0 cm diameter, oscillating at constant peak-to-peak amplitude of 4.0 cm immersed in a fluidized bed of 355 – 420μm glass particle 343

Table 13.33: Heat transfer coefficient for a copper sphere of 1.0 cm diameter, oscillating at constant peak-to-peak amplitude of 1.8 cm immersed in a fluidized bed of 355 – 420μm glass particle 344

Table 13.34: Heat transfer coefficient for a copper sphere of 1.4 cm diameter, oscillating at constant peak-to-peak amplitude of 6.9 cm immersed in a fluidized bed of 355 – 420μm glass particle 345

Table 13.35: Heat transfer coefficient for a copper sphere of 1.4 cm diameter, oscillating at constant peak-to-peak amplitude of 4.0 cm immersed in a fluidized bed of 355 – 420μm glass particle 346
Table 13.36: Heat transfer coefficient for a copper sphere of 1.4 cm diameter, oscillating at constant peak-to-peak amplitude of 1.8 cm immersed in a fluidized bed of 355 — 420μm glass particle.

Table 13.37: Heat transfer coefficient for a copper sphere of 2.0 cm diameter, oscillating at constant peak-to-peak amplitude of 6.9 cm immersed in a fluidized bed of 355 — 420μm glass particle.

Table 13.38: Heat transfer coefficient for a copper sphere of 2.0 cm diameter, oscillating at constant peak-to-peak amplitude of 4.0 cm immersed in a fluidized bed of 355 — 420μm glass particle.

Table 13.39: Heat transfer coefficient for a copper sphere of 2.0 cm diameter, oscillating at constant peak-to-peak amplitude of 1.8 cm immersed in a fluidized bed of 355 — 420μm glass particle.
LIST OF FIGURES

<table>
<thead>
<tr>
<th>Figure</th>
<th>Description</th>
<th>Page</th>
</tr>
</thead>
<tbody>
<tr>
<td>1.1</td>
<td>Classification of fluidized bed applications according to predominating mechanism [Geldart 1986]</td>
<td>5</td>
</tr>
<tr>
<td>1.2</td>
<td>Defluidized and gas film zone, Gelperin et al. [1963]</td>
<td>24</td>
</tr>
<tr>
<td>1.3</td>
<td>Laminar boundary layer thickness between two layers of particles close to the heat transfer wall [Levenspiel and Walton 1954]</td>
<td>36</td>
</tr>
<tr>
<td>1.4</td>
<td>Packet theory model</td>
<td>43</td>
</tr>
<tr>
<td>2.1</td>
<td>Schematic of the experimental setup</td>
<td>59</td>
</tr>
<tr>
<td>2.2</td>
<td>Photograph of the experimental set up</td>
<td>60</td>
</tr>
<tr>
<td>2.3</td>
<td>Schematic of the fluidized bed</td>
<td>62</td>
</tr>
<tr>
<td>2.4</td>
<td>Schematic of a motor driven gear mechanism for controlling sphere motion in linear direction inside the fluidized bed</td>
<td>63</td>
</tr>
<tr>
<td>2.5</td>
<td>Schematic of a sliding mechanism for controlling oscillatory sphere motion within the fluidized bed</td>
<td>65</td>
</tr>
<tr>
<td>2.6</td>
<td>Photograph of 5-44 μm glass particles, magnification factor=320</td>
<td>68</td>
</tr>
</tbody>
</table>
Figure 2.7: Photograph of 126-147 μm glass particles, magnification factor=64 ........................................... 69
Figure 2.8: Photograph of 355-420 μm glass particles, magnification factor=64 ........................................... 70
Figure 2.9: Computer controlled data acquisition system ................................. 72
Figure 2.10: Typical thermocouple output versus time for 2 cm diameter stationary sphere ........................................... 82
Figure 2.11: Typical thermocouple output versus time for 2 cm diameter linearly downward moving sphere at 3.0 cm/s .............. 83
Figure 3.1: Bed-pressure drop vs. superficial air velocity for 5-44 μm glass particles ........................................... 85
Figure 3.2: Bed-pressure drop vs. superficial air velocity for 126-147 μm glass particles ........................................... 86
Figure 3.3: Bed-pressure drop vs. superficial air velocity for 355-420 μm glass particles ........................................... 87
Figure 3.4: Diagram for powder classification into groups [Geldart, 1986] 93
Figure 3.5: Heat transfer coefficient vs. superficial air velocity for 5-44 μm size glass particles and various size copper spheres ... 95
Figure 3.6: Heat transfer coefficient vs. superficial air velocity for 126-147 μm size glass particles and various size copper spheres 96
Figure 3.7: Heat transfer coefficient vs. superficial air velocity for 355-420 μm size glass particles and various size copper spheres 97
Figure 3.8: Heat transfer coefficient vs. superficial air velocity for a copper sphere diameter of 1.0 cm and various size glass particles.

Figure 3.9: Heat transfer coefficient vs. superficial air velocity for a copper sphere diameter of 1.4 cm and various size glass particles.

Figure 3.10: Heat transfer coefficient vs. superficial air velocity for a copper sphere of 2.0 cm diameter and various size glass particles.

Figure 3.11: Scatter in heat transfer coefficient as a function of the superficial air velocity for a copper sphere of 1.4 cm diameter and 126-147 μm glass particles.

Figure 3.12: Contribution of gas convection heat transfer to particle convection heat transfer for a 1.0 cm copper sphere immersed in a fluidized bed of 5-44 μm glass particles.

Figure 3.13: Contribution of gas convection heat transfer to particle convection heat transfer for a 1.0 cm copper sphere immersed in a fluidized bed of 126-147 μm glass particles.

Figure 3.14: Contribution of gas convection heat transfer to particle convection heat transfer for a 1.0 cm copper sphere immersed in a fluidized bed of 355-420 μm glass particles.

Figure 3.15: Heat transfer coefficient vs. average sphere temperature for a temperature step of 5 K and for a 1.0 cm copper sphere immersed in a fluidized bed of 126-147 μm glass particles.
Figure 3.16: Heat transfer coefficient vs. average sphere temperature for a temperature step of 10 K and for a 1.0 cm copper sphere immersed in a fluidized bed of 126-147 μm glass particles.

Figure 3.17: Heat transfer coefficient vs. average sphere temperature for a temperature step of 15 K and for a 1.0 cm copper sphere immersed in a fluidized bed of 126-147 μm glass particles.

Figure 3.18: Heat transfer coefficient vs. average sphere temperature for a temperature step of 20 K and for a 1.0 cm copper sphere immersed in a fluidized bed of 126-147 μm glass particles.

Figure 3.19: Heat transfer coefficient vs. average sphere temperature for a temperature step of 30 K and for a 1.0 cm copper sphere immersed in a fluidized bed of 126-147 μm glass particles.

Figure 3.20: Heat transfer coefficient vs. average sphere temperature for a 1.0 cm copper sphere immersed in a packed bed of 126-147 μm glass particles.

Figure 3.21: Heat transfer coefficient for a 1.0 cm copper sphere in linearly downward motion (various speeds) versus superficial air velocity for a fluidized bed of 5-44 μm glass particles.

Figure 3.22: Heat transfer coefficient of a 1.4 cm copper sphere in linearly downward motion (various speeds) versus superficial air velocity for a fluidized bed of 5-44 μm glass particles.
Figure 3.23: Heat transfer coefficient of a 2.0 cm copper sphere in linearly downward motion (various speeds) versus superficial air velocity for a fluidized bed of 5-44 μm glass particles . . 123

Figure 3.24: Heat transfer coefficient of a 1.0 cm copper sphere in linearly downward motion (various speeds) versus superficial air velocity for a fluidized bed of 126-147 μm glass particles 125

Figure 3.25: Heat transfer coefficient of a 1.4 cm copper sphere in linearly downward motion (various speeds) versus superficial air velocity for a fluidized bed of 126-147 μm glass particles 126

Figure 3.26: Heat transfer coefficient of a 2.0 cm copper sphere in linearly downward motion (various speeds) versus superficial air velocity for a fluidized bed of 126-147 μm glass particles 127

Figure 3.27: Heat transfer coefficient of a 1.0 cm copper sphere in linearly downward motion (various speeds) versus superficial air velocity for a fluidized bed of 355-420 μm glass particles 128

Figure 3.28: Heat transfer coefficient of a 1.4 cm copper sphere in linearly downward motion (various speeds) versus superficial air velocity for a fluidized bed of 355-420 μm glass particles 129

Figure 3.29: Heat transfer coefficient of a 2.0 cm copper sphere in linearly downward motion (various speeds) versus superficial air velocity for a fluidized bed of 355-420 μm glass particles 130
Figure 3.30: Industrial figure of merit, heat transfer to operating cost ratio, normalized to the best stationary sphere operation of the bed

Figure 3.31: Heat transfer coefficient of a 1.0 cm copper sphere in oscillating motion (various frequencies) with the peak-to-peak amplitude of 1.8 cm versus superficial air velocity for a fluidized bed of 5-44 μm glass particles

Figure 3.32: Heat transfer coefficient of a 1.0 cm copper sphere in oscillating motion (various frequencies) with the peak-to-peak amplitude of 4.0 cm versus superficial air velocity for a fluidized bed of 5-44 μm glass particles

Figure 3.33: Heat transfer coefficient of a 1.0 cm copper sphere in oscillating motion (various frequencies) with the peak-to-peak amplitude of 6.9 cm versus superficial air velocity for a fluidized bed of 5-44 μm glass particles

Figure 3.34: Heat transfer coefficient of a 1.4 cm copper sphere in oscillating motion (various frequencies) with the peak-to-peak amplitude of 1.8 cm versus superficial air velocity for a fluidized bed of 5-44 μm glass particles

Figure 3.35: Heat transfer coefficient of a 1.4 cm copper sphere in oscillating motion (various frequencies) with the peak-to-peak amplitude of 4.0 cm versus superficial air velocity for a fluidized bed of 5-44 μm glass particles
Figure 3.36: Heat transfer coefficient of a 1.4 cm copper sphere in oscillating motion (various frequencies) with the peak-to-peak amplitude of 6.9 cm versus superficial air velocity for a fluidized bed of 5-44 μm glass particles ............. 149

Figure 3.37: Heat transfer coefficient of a 2.0 cm copper sphere in oscillating motion (various frequencies) with the peak-to-peak amplitude of 1.8 cm versus superficial air velocity for a fluidized bed of 5-44 μm glass particles ............. 150

Figure 3.38: Heat transfer coefficient of a 2.0 cm copper sphere in oscillating motion (various frequencies) with the peak-to-peak amplitude of 4.0 cm versus superficial air velocity for a fluidized bed of 5-44 μm glass particles ............. 151

Figure 3.39: Heat transfer coefficient of a 2.0 cm copper sphere in oscillating motion (various frequencies) with the peak-to-peak amplitude of 6.9 cm versus superficial air velocity for a fluidized bed of 5-44 μm glass particles ............. 152

Figure 3.40: Heat transfer coefficient of a 1.0 cm copper sphere in oscillating motion (various frequencies) with the peak-to-peak amplitude of 1.8 cm versus superficial air velocity for a fluidized bed of 126-147 μm glass particles ............. 153
Figure 3.41: Heat transfer coefficient of a 1.0 cm copper sphere in oscillating motion (various frequencies) with the peak-to-peak amplitude of 4.0 cm versus superficial air velocity for a fluidized bed of 126-147 μm glass particles .......................... 154

Figure 3.42: Heat transfer coefficient of a 1.0 cm copper sphere in oscillating motion (various frequencies) with the peak-to-peak amplitude of 6.9 cm versus superficial air velocity for a fluidized bed of 126-147 μm glass particles .......................... 155

Figure 3.43: Heat transfer coefficient of a 1.4 cm copper sphere in oscillating motion (various frequencies) with the peak-to-peak amplitude of 1.8 cm versus superficial air velocity for a fluidized bed of 126-147 μm glass particles .......................... 156

Figure 3.44: Heat transfer coefficient of a 1.4 cm copper sphere in oscillating motion (various frequencies) with the peak-to-peak amplitude of 4.0 cm versus superficial air velocity for a fluidized bed of 126-147 μm glass particles .......................... 157

Figure 3.45: Heat transfer coefficient of a 1.4 cm copper sphere in oscillating motion (various frequencies) with the peak-to-peak amplitude of 6.9 cm versus superficial air velocity for a fluidized bed of 126-147 μm glass particles .......................... 158
Figure 3.46: Heat transfer coefficient of a 2.0 cm copper sphere in oscillating motion (various frequencies) with the peak-to-peak amplitude of 1.8 cm versus superficial air velocity for a fluidized bed of 126-147 μm glass particles .................. 159

Figure 3.47: Heat transfer coefficient of a 2.0 cm copper sphere in oscillating motion (various frequencies) with the peak-to-peak amplitude of 4.0 cm versus superficial air velocity for a fluidized bed of 126-147 μm glass particles .................. 160

Figure 3.48: Heat transfer coefficient of a 2.0 cm copper sphere in oscillating motion (various frequencies) with the peak-to-peak amplitude of 6.9 cm versus superficial air velocity for a fluidized bed of 126-147 μm glass particles .................. 161

Figure 3.49: Heat transfer coefficient of a 1.0 cm copper sphere in oscillating motion (various frequencies) with the peak-to-peak amplitude of 1.8 cm versus superficial air velocity for a fluidized bed of 355-420 μm glass particles .................. 162

Figure 3.50: Heat transfer coefficient of a 1.0 cm copper sphere in oscillating motion (various frequencies) with the peak-to-peak amplitude of 4.0 cm versus superficial air velocity for a fluidized bed of 355-420 μm glass particles .................. 163
Figure 3.51: Heat transfer coefficient of a 1.0 cm copper sphere in oscillating motion (various frequencies) with the peak-to-peak amplitude of 6.9 cm versus superficial air velocity for a fluidized bed of 355-420 \( \mu \text{m} \) glass particles 164

Figure 3.52: Heat transfer coefficient of a 1.4 cm copper sphere in oscillating motion (various frequencies) with the peak-to-peak amplitude of 1.8 cm versus superficial air velocity for a fluidized bed of 355-420 \( \mu \text{m} \) glass particles 165

Figure 3.53: Heat transfer coefficient of a 1.4 cm copper sphere in oscillating motion (various frequencies) with the peak-to-peak amplitude of 4.0 cm versus superficial air velocity for a fluidized bed of 355-420 \( \mu \text{m} \) glass particles 166

Figure 3.54: Heat transfer coefficient of a 1.4 cm copper sphere in oscillating motion (various frequencies) with the peak-to-peak amplitude of 6.9 cm versus superficial air velocity for a fluidized bed of 355-420 \( \mu \text{m} \) glass particles 167

Figure 3.55: Heat transfer coefficient of a 2.0 cm copper sphere in oscillating motion (various frequencies) with the peak-to-peak amplitude of 1.8 cm versus superficial air velocity for a fluidized bed of 355-420 \( \mu \text{m} \) glass particles 168
<table>
<thead>
<tr>
<th>Figure</th>
<th>Description</th>
</tr>
</thead>
<tbody>
<tr>
<td>3.56</td>
<td>Heat transfer coefficient of a 2.0 cm copper sphere in oscillating motion (various frequencies) with the peak-to-peak amplitude of 4.0 cm versus superficial air velocity for a fluidized bed of 355-420 μm glass particles. 169</td>
</tr>
<tr>
<td>3.57</td>
<td>Heat transfer coefficient of a 2.0 cm copper sphere in oscillating motion (various frequencies) with the peak-to-peak amplitude of 6.9 cm versus superficial air velocity for a fluidized bed of 355-420 μm glass particles. 170</td>
</tr>
<tr>
<td>3.58</td>
<td>Heat transfer coefficient of a 1.0 cm oscillating copper sphere moving at various equivalent average sphere velocities versus superficial air velocity for a fluidized bed of 5-44 μm glass particles. 171</td>
</tr>
<tr>
<td>3.59</td>
<td>Heat transfer coefficient of a 1.0 cm oscillating copper sphere versus average sphere velocity for various superficial air velocities. 172</td>
</tr>
<tr>
<td>3.60</td>
<td>Heat transfer coefficient of a 1.0 cm oscillating copper sphere moving at various equivalent average sphere velocities versus superficial air velocity for a fluidized bed of 126-147 μm glass particles. 173</td>
</tr>
<tr>
<td>3.61</td>
<td>Heat transfer coefficient of a 1.0 cm oscillating copper sphere versus average sphere velocity for various superficial air velocities. 174</td>
</tr>
</tbody>
</table>
Figure 3.62: Heat transfer coefficient of a 1.0 cm oscillating copper sphere moving at various equivalent average sphere velocities versus superficial air velocity for a fluidized bed of 355-420 μm glass particles .................................................. 175

Figure 3.63: Heat transfer coefficient of a 1.0 cm oscillating copper sphere versus average sphere velocity for various superficial air velocities .............................................................. 176

Figure 3.64: Heat transfer coefficient of a 1.0 cm oscillating and linearly downward moving copper sphere versus superficial air velocities for various average velocities and 5-44 μm glass particles 177

Figure 3.65: Heat transfer coefficient of a 1.0 cm oscillating and linearly downward moving copper sphere versus superficial air velocities for various average velocities and 126-147 μm glass particles .................................................. 178

Figure 3.66: Heat transfer coefficient of a 1.0 cm oscillating and linearly downward moving copper sphere versus superficial air velocities for various average velocities and 355-420 μm glass particles .................................................. 179

Figure 3.67: Heat transfer coefficient for various sphere diameters versus superficial air velocity for the same linear average velocity and 5-44 μm glass particles .................................................. 180
Figure 3.68: Heat transfer coefficient for various sphere diameters versus superficial air velocity for the same linear average velocity and 126-147 \( \mu \text{m} \) glass particles \hspace{1cm} 181

Figure 3.69: Heat transfer coefficient for various sphere diameters versus superficial air velocity for the same linear average velocity and 355-420 \( \mu \text{m} \) glass particles \hspace{1cm} 182

Figure 4.1: Experimental and predicted heat transfer coefficients as a function of superficial air velocity for the fluidized bed of 5-44 \( \mu \text{m} \) glass particles and stationary sphere \hspace{1cm} 185

Figure 4.2: Experimental and predicted heat transfer coefficients as a function of superficial air velocity for the fluidized bed of 126-147 \( \mu \text{m} \) glass particles and stationary sphere \hspace{1cm} 186

Figure 4.3: Experimental and predicted heat transfer coefficients as a function of superficial air velocity for the fluidized bed of 355-420 \( \mu \text{m} \) glass particles and stationary sphere \hspace{1cm} 187

Figure 4.4: Experimental and predicted heat transfer coefficients as a function of XSCORR \hspace{1cm} 189

Figure 4.5: Experimental and predicted heat transfer coefficients (correlation I, equation 4.3) as a function of superficial air velocity for the linearly downward moving sphere in a fluidized bed of 5-44 \( \mu \text{m} \) glass particles \hspace{1cm} 190
Figure 4.6: Experimental and predicted heat transfer coefficients (correlation I, equation 4.3) as a function of superficial air velocity for the linearly downward moving sphere in a fluidized bed of 126-147 μm glass particles

Figure 4.7: Experimental and predicted heat transfer coefficients (correlation I, equation 4.3) as a function of superficial air velocity for the linearly downward moving sphere in a fluidized bed of 355-420 μm glass particles

Figure 4.8: Experimental and predicted heat transfer coefficients (correlation II, equation 4.5) as a function of superficial air velocity for the linearly downward moving sphere in a fluidized bed of 5-44 μm glass particles

Figure 4.9: Experimental and predicted heat transfer coefficients (correlation II, equation 4.5) as a function of superficial air velocity for the linearly downward moving sphere in a fluidized bed of 126-147 μm glass particles

Figure 4.10: Experimental and predicted heat transfer coefficients (correlation II, equation 4.5) as a function of superficial air velocity for the linearly downward moving sphere in a fluidized bed of 355-420 μm glass particles
Figure 4.11: Experimental and predicted heat transfer coefficients (correlation III, equation 4.6) as a function of superficial air velocity for the linearly downward moving 1.0 cm diameter sphere in a fluidized bed of 5-44 \(\mu m\) glass particles. 200

Figure 4.12: Experimental and predicted heat transfer coefficients (correlation III, equation 4.6) as a function of superficial air velocity for the linearly downward moving 1.4 cm diameter sphere in a fluidized bed of 5-44 \(\mu m\) glass particles. 201

Figure 4.13: Experimental and predicted heat transfer coefficients (correlation III, equation 4.6) as a function of superficial air velocity for the linearly downward moving 2.0 cm diameter sphere in a fluidized bed of 5-44 \(\mu m\) glass particles. 202

Figure 4.14: Experimental and predicted heat transfer coefficients (correlation III, equation 4.6) as a function of superficial air velocity for the linearly downward moving 1.0 cm diameter sphere in a fluidized bed of 126-147 \(\mu m\) glass particles. 203

Figure 4.15: Experimental and predicted heat transfer coefficients (correlation III, equation) as a function of superficial air velocity for the linearly downward moving 1.4 cm diameter sphere in a fluidized bed of 126-147 \(\mu m\) glass particles. 208
Figure 4.16: Experimental and predicted heat transfer coefficients (correlation III, equation 4.6) as a function of superficial air velocity for the linearly downward moving 2.0 cm diameter sphere in a fluidized bed of 126-147 μm glass particles . . . . . . . . 209

Figure 4.17: Experimental and predicted heat transfer coefficients (correlation III, equation 4.6) as a function of superficial air velocity for the linearly downward moving 1.0 cm diameter sphere in a fluidized bed of 355-420 μm glass particles . . . . . . . . 210

Figure 4.18: Experimental and predicted heat transfer coefficients (correlation III, equation 4.6) as a function of superficial air velocity for the linearly downward moving 1.4 cm diameter sphere in a fluidized bed of 355-420 μm glass particles . . . . . . . . 211

Figure 4.19: Experimental and predicted heat transfer coefficients (correlation III, equation 4.6) as a function of superficial air velocity for the linearly downward moving 2.0 cm diameter sphere in a fluidized bed of 355-420 μm glass particles . . . . . . . . 212

Figure 4.20: Predicted heat transfer coefficients (correlation III, equation 4.6) as a function of the experimental heat transfer coefficient for a fluidized bed of 5-44 μm glass particles . . . . . . . . . 213

Figure 4.21: Predicted heat transfer coefficients (correlation III, equation 4.6) as a function of the experimental heat transfer coefficient for a fluidized bed of 126-147 μm glass particles . . . . . . . . 214
Figure 4.22: Predicted heat transfer coefficients (correlation III, equation 4.6) as a function of the experimental heat transfer coefficient for a fluidized bed of 355-420 μm glass particles .......................... 215

Figure 4.23: Experimental and predicted heat transfer coefficients as a function of superficial air velocity for an oscillating sphere in a fluidized bed of 5-44 μm glass particles .......................... 216

Figure 4.24: Experimental and predicted heat transfer coefficients as a function of superficial air velocity for an oscillating sphere in a fluidized bed of 126-147 μm glass particles .......................... 217

Figure 4.25: Experimental and predicted heat transfer coefficients as a function of superficial air velocity for an oscillating sphere in a fluidized bed of 355-420 μm glass particles .......................... 218

Figure 4.26: Predicted heat transfer coefficients as a function of the experimental heat transfer coefficient for an oscillating sphere in a fluidized bed .................................................. 219

Figure 5.1: Schematic diagram to illustrate the mechanism of heat transfer from a surface to a fluidized bed .......................... 221

Figure 5.2: Experimental and theoretical (equation 5.6) heat transfer coefficient versus the residence time of the emulsion packet for 5-44 μm glass particle fluidized bed near minimum fluidization velocity .................................................. 225
Figure 5.3: Experimental and theoretical (equation 5.6) heat transfer coefficient versus the residence time of the emulsion packet for 126-147μm glass particle fluidized bed near minimum fluidization velocity .............................. 226

Figure 5.4: Experimental and theoretical (equation 5.6) heat transfer coefficient versus the residence time of the emulsion packet for 355-420μm glass particle fluidized bed near minimum fluidization velocity .............................. 227

Figure 5.5: Plot of the radial distance in the bed versus the particulate velocity for 126-147 μm glass particle-fluidized bed system at 3 second .............................. 233

Figure 11.1: Thermocouple wire with the insulation. (a) Thermocouple wires with insulation. (b) equivalent radii .................. 273

Figure 11.2: Thermocouple wires with equivalent radii inside a hollow steel tube .............................. 275
NOMENCLATURE

a constant in the heat capacity of the copper sphere relationship, J/Kg K

$a_1$ radius of cylindrical heater, m

A vibrational amplitude, m

$A_1$ empirical factor

$A_c$ area of contact of a packet with heat transfer surface

$A_f$ area of the float inside a rotameter

$(AK)_{eq}$ equivalent area-conductivity product for the thermocouple wire and insulation

$Ar$ Archimedes number, $(\frac{g d^2 \rho g (\rho_p - \rho)}{\mu^2})$

$A_s$ vibrational amplitude of a sphere, m

$A_{sph}$ surface area of copper sphere, $m^2$

$A_w$ Variable area of the orifice of the rotameter

b constant in the heat capacity of the copper sphere relationship, J/Kg $K^2$

$Bi$ Biot number, $\frac{hL_{ch}}{K_{cu}}$

C perimeter of the outside of the hollow steel tube housing a
thermocouple wire

$C_D$ drag coefficient

$C_g$ specific heat of fluidizing gas ($J/KgK$)

$C_0$ constant

$C_{eff}$ effective heat capacity of the particles, ($J/KgK$)

$C_{m1}$ Constant for heat transfer correlation-II for a moving sphere

$C_{m2}$ Constant for heat transfer correlation-II for a moving sphere

$C_{m3}$ Constant for heat transfer correlation-II for a moving sphere

$C_{m4}$ Constant for heat transfer correlation-II for a moving sphere

$C_{m5}$ Constant for heat transfer correlation-II for a moving sphere

$C_{mm1}$ Constant for heat transfer correlation-III for a moving sphere

$C_{mm2}$ Constant for heat transfer correlation-III for a moving sphere

$C_{mm3}$ Constant for heat transfer correlation-III for a moving sphere

$C_{mm4}$ Constant for heat transfer correlation-III for a moving sphere
<table>
<thead>
<tr>
<th>Symbol</th>
<th>Description</th>
</tr>
</thead>
<tbody>
<tr>
<td>$C_p$</td>
<td>specific heat of the glass particles, $(J/Kg \cdot K)$</td>
</tr>
<tr>
<td>$C_p(x)$</td>
<td>specific heat of emulsion phase at point x, $(J/Kg \cdot K)$</td>
</tr>
<tr>
<td>$C_{sph}$</td>
<td>heat capacity of the copper sphere, $J/Kg \cdot K$</td>
</tr>
<tr>
<td>$C_{s1}$</td>
<td>Constant for heat transfer correlation for a stationary sphere</td>
</tr>
<tr>
<td>$C_{s2}$</td>
<td>Constant for heat transfer correlation for a stationary sphere</td>
</tr>
<tr>
<td>$C_{s3}$</td>
<td>Constant for heat transfer correlation for a stationary sphere</td>
</tr>
<tr>
<td>$C_{s4}$</td>
<td>Constant for heat transfer correlation for a stationary sphere</td>
</tr>
<tr>
<td>$d_c$</td>
<td>diameter of equivalent cylinder, m</td>
</tr>
<tr>
<td>$d_p$</td>
<td>particle diameter, m</td>
</tr>
<tr>
<td>$D_{probe}$</td>
<td>heat transfer probe diameter, m</td>
</tr>
<tr>
<td>$D_r$</td>
<td>distance between two successive layers of particles, m</td>
</tr>
<tr>
<td>$D_{sph}$</td>
<td>heat transfer sphere diameter, m</td>
</tr>
<tr>
<td>$D_t$</td>
<td>fluidized bed diameter, m</td>
</tr>
<tr>
<td>$f_0$</td>
<td>fraction of time that bubbles shrouds the heat transfer surface</td>
</tr>
<tr>
<td>$f_s$</td>
<td>surface heat flux, $W/m^2$</td>
</tr>
<tr>
<td>$F_s$</td>
<td>frequency of sphere oscillation, Hz</td>
</tr>
<tr>
<td>$g$</td>
<td>gravitational acceleration, $m/s^2$</td>
</tr>
<tr>
<td>$G$</td>
<td>superficial mass velocity of gas, $Kg/m^2 \cdot s$</td>
</tr>
<tr>
<td>$h$</td>
<td>heat transfer coefficient for sphere-bed system, $W/m^2 \cdot K$</td>
</tr>
</tbody>
</table>
heat transfer coefficient for the steel tube immersed in
the fluidized bed
heat transfer coefficient through contact resistance, W/m^2 K
conductive heat transfer coefficient, W/m^2 K
equivalent heat transfer coefficient for the thermocouple
and steel tube assembly
gas convective heat transfer coefficient, W/m^2 K
instantaneous heat transfer coefficient, W/m^2 K
heat transfer coefficient for a moving sphere, W/m^2 K
maximum heat transfer coefficient, W/m^2 K
particle convection heat transfer coefficient, W/m^2 K
radiant particle heat transfer coefficient, W/m^2 K
radiation heat transfer coefficient for
sphere-bed system, W/m^2 K
heat transfer coefficient for a stationary sphere, W/m^2 K
total heat transfer coefficient (radiation and convection) for
sphere-bed system, W/m^2 K
instantaneous local heat transfer coefficient, W/m^2 K
local time averaged heat transfer coefficient, W/m^2 K
age distribution function of emulsion packet on a surface
air thermal conductivity, W/m^2 K
thermal conductivity of copper, W/m^2 K
$K_{cn}$ thermal conductivity of constantan

$K_g^*$ thermal conductivity of packed bed, $W/m^2 K$

$K_{eff}$ effective thermal conductivity of the packet, $W/m^2 K$

$K_{ew}$ effective thermal conductivity of the contact zone near the wall, $W/m^2 K$

$K_g$ gas thermal conductivity, $W/m^2 K$

$K_{ins}$ thermal conductivity of thermocouple insulation, $W/m^2 K$

$K_M$ thermal conductivity of two phase boundary layer, $W/m^2 K$

$K_{mf}$ effective thermal conductivity of the packet at minimum fluidization condition, $W/m^2 K$

$K_p$ thermal conductivity of solid particle, $W/m^2 K$

$K_s$ thermal conductivity of steel $W/m^2 K$

$K(x)$ thermal conductivity of emulsion phase at point x, $W/m^2 K$

$l$ length of heater either flat or cylindrical, m

$l_e$ length of emulsion packet, m

$l_{mf}$ height of the bed at incipient fluidization, m

$L_1$ length of the thermocouple wire along with the insulation

$L_2$ width of the thermocouple wire along with the insulation

$L_{ch}$ characteristic length, m

$l_H$ length of a heater, m

$m$ constant

$Nu$ Nusselt number based on sphere diameter

$(Nu_l)_{conv}$ convective Nusselt number based on characteristic length


\( N_u_m \) Nusselt number for a moving sphere

\( N_{umax} \) maximum Nusselt number

\( N_{uo} \) Nusselt number for an oscillating sphere

\( N_{us} \) Nusselt number for a stationary sphere

\( P \) bed pressure

\( P_b \) Static pressure at fluidized bed cm of water

\( P_o \) static pressure at operating condition at rotameter section, cm of water

\( Pr \) Prantle number, \( \left( \frac{C_p \mu g}{K g} \right) \)

\( P_s \) standard atmospheric pressure in cm of Hg

\( Q_r \) air flow rate at rotameter section, cc/s

\( Q_{rc} \) air flow rate at rotameter section at calibration condition, cc/s

\( Q_{ro} \) air flow rate at rotameter section during the operating condition, cc/s

\( Q_{rs} \) air flow rate at rotameter section at standard condition, cc/s

\( Q_{wm} \) air flow rate at wet testmeter section at calibration condition, cc/s

\( q_{cl} \) heat loss due to conduction through the thermocouple wires

\( q_{wi} \) instantaneous rate of heat flux into emulsion packet, \( W/m^2 \)

\( R \) radius of cylinder or sphere, m

\( R_a \) thermal resistance of the packet of emulsion, \( m^2 \text{ K/W} \)

\( Re \) Reynolds number based on sphere diameter

\( Re_{mf} \) Reynolds number at minimum fluidization, \( \left( \frac{d \rho g U_m f}{\mu g} \right) \)

\( Re_l \) Reynolds number based on characteristic length
<table>
<thead>
<tr>
<th>Symbol</th>
<th>Definition</th>
</tr>
</thead>
<tbody>
<tr>
<td>$Re_p$</td>
<td>Reynolds number based on particle diameter</td>
</tr>
<tr>
<td>$R_w$</td>
<td>contact resistance near a wall, $m^2$ K/W</td>
</tr>
<tr>
<td>$r$</td>
<td>$r$ coordinate</td>
</tr>
<tr>
<td>$r_b$</td>
<td>fluidized bed diameter, m</td>
</tr>
<tr>
<td>$r_w$</td>
<td>thermocouple wire radius, m</td>
</tr>
<tr>
<td>$r_1$</td>
<td>radius of the equivalent thermocouple wire, m</td>
</tr>
<tr>
<td>$r_2$</td>
<td>radius of the equivalent thermocouple insulation, m</td>
</tr>
<tr>
<td>$r_3$</td>
<td>inner radius of the hollow steel tube, m</td>
</tr>
<tr>
<td>$r_4$</td>
<td>outer radius of the hollow steel tube, m</td>
</tr>
<tr>
<td>$s$</td>
<td>average replacement of packet at the wall/unit time by means of side mixing</td>
</tr>
<tr>
<td>$S$</td>
<td>stirring factor</td>
</tr>
<tr>
<td>$S_1$</td>
<td>renewal frequency, $s^{-1}$</td>
</tr>
<tr>
<td>$S_L$</td>
<td>expression given by equation</td>
</tr>
<tr>
<td>$t$</td>
<td>time s</td>
</tr>
<tr>
<td>$T_0$</td>
<td>initial temperature of the copper sphere, K</td>
</tr>
<tr>
<td>$T$</td>
<td>temperature of the center of the copper sphere, K</td>
</tr>
<tr>
<td>$T_C$</td>
<td>time constant of emulsion packet, sec</td>
</tr>
<tr>
<td>$T_b$</td>
<td>fluidized bed temperature, K</td>
</tr>
<tr>
<td>$T_e$</td>
<td>temperature of emulsion phase, K</td>
</tr>
<tr>
<td>$T_{gc}$</td>
<td>gas residence time, sec</td>
</tr>
<tr>
<td>$T_{rc}$</td>
<td>temperature at rotameter section at calibrating condition, K</td>
</tr>
<tr>
<td>$T_s$</td>
<td>standard temperature, K</td>
</tr>
</tbody>
</table>
$T_{tb}$ thermocouple temperature at copper sphere end, K

$T_w$ heat transfer wall temperature, K

$T_{w,m}$ temperature at wet testmeter at calibrating condition, K

$U$ superficial gas velocity, m/s

$U_{D_{sph}}$ uncertainty in sphere diameter

$U_{dp}$ uncertainty in particle diameter

$U_g$ gas phase velocity in r direction, m/s

$U_h$ uncertainty in heat transfer coefficient

$U_{mb}$ minimum bubbling velocity, m/s

$U_{mf}$ minimum fluidization velocity, m/s

$U_p$ particle velocity in r direction m/s

$U_t$ terminal velocity of glass particle velocity m/s

$U_U$ uncertainty in superficial air velocity

$U_{Um,f}$ uncertainty in minimum superficial air velocity

$U_{V_{sph}}$ uncertainty in sphere velocity

$U_{X_1}$ uncertainty in $X_1$

$U_{X_2}$ uncertainty in $X_2$

$U_{X_m}$ uncertainty in $X_m$

$U_Z$ uncertainty in Z

$V$ volume of the copper sphere, $m^3$

$V_{avg}$ equivalent average linear velocity of an oscillating sphere, m/s

$V_f$ volume of the float inside the rotameter, $m^3$
$V_g$  \hspace{1cm} \text{gas velocity in y direction, cm/s}
$V_p$  \hspace{1cm} \text{particle velocity in y direction, cm/s}
$V_{sph}$  \hspace{1cm} \text{linear velocity of sphere, cm/s}
$x$  \hspace{1cm} \text{distance from constraining surface, m}
$X_1$  \hspace{1cm} \text{experimental variable}
$X_2$  \hspace{1cm} \text{experimental variable}
$X_m$  \hspace{1cm} \text{experimental variable}
$x_{th}$  \hspace{1cm} \text{thermal penetration distance, m}
$y$  \hspace{1cm} \text{y coordinate}

**Greek**

$\alpha_{eff}$  \hspace{1cm} \text{thermal diffusivity of emulsion packet, m$^2$/s}
$\alpha_M$  \hspace{1cm} \text{thermal diffusivity of two phase boundary layer, m$^2$/s}
$\beta_r$  \hspace{1cm} \text{fluid-to-particle friction coefficient in r direction}
$\beta_y$  \hspace{1cm} \text{fluid-to-particle friction coefficient in y direction}
$\gamma_{bed}$  \hspace{1cm} \text{mean particle concentration}
$\delta$  \hspace{1cm} \text{gas film thickness, m}
$\delta_T$  \hspace{1cm} \text{two phase boundary layer thickness around an immersed cylinder or sphere, m}
$\delta_w$  \hspace{1cm} \text{thickness of the contact resistance zone, m}
$\epsilon$  \hspace{1cm} \text{void fraction}
$\epsilon_{mf}$  \hspace{1cm} \text{void fraction at minimum fluidization}
\( \varepsilon_{sph} \)  emissivity of the copper sphere

\( \Theta \)  absolute bed temperature in K/273

\( \Delta p \)  pressure drop across the bed, cm of water

\( \mu_g \)  viscosity of gas, \( Ns/m^2 \)

\( \nu_g \)  kinematic viscosity of gas, \( m^2/s \)

\( \rho_b \)  fluidized bed density, \( Kg/m^3 \)

\( \rho_{eff} \)  effective density of a fluidized bed, \( Kg/m^3 \)

\( \rho_f \)  density of the float inside the rotameter, \( Kg/m^3 \)

\( \rho_g \)  density of fluidizing gas, \( Kg/m^3 \)

\( \rho_{mf} \)  fluidized bed density at incipient condition, \( Kg/m^3 \)

\( \rho_o \)  density of fluidizing gas at rotameter section under operating condition, \( Kg/m^3 \)

\( \rho_p \)  density of fluidizing particle, \( Kg/m^3 \)

\( \rho_{pb} \)  bulk density of packed bed, \( kg/m^3 \)

\( \rho_{rc} \)  density of fluidizing gas at rotameter section under calibration condition, \( Kg/m^3 \)

\( \rho_{sph} \)  density of copper sphere, \( kg/m^3 \)

\( \rho_{std} \)  density of fluidizing gas at standard condition, \( Kg/m^3 \)

\( \rho_{wm} \)  density of fluidizing gas at wet testmeter section under calibration condition, \( Kg/m^3 \)

\( \rho(x) \)  density of emulsion phase at distance \( x \), \( Kg/m^3 \)

\( \sigma \)  Stefan-Boltzman constant, \( W/m^2 K^4 \)
<table>
<thead>
<tr>
<th>Symbol</th>
<th>Definition</th>
</tr>
</thead>
<tbody>
<tr>
<td>( \tau )</td>
<td>residence time of emulsion packet on a heat transfer surface, s</td>
</tr>
<tr>
<td>( \tau_c )</td>
<td>time constant or response time of the copper sphere-bed system, sec</td>
</tr>
<tr>
<td>( \tau_{rr} )</td>
<td>solid stress in r direction</td>
</tr>
<tr>
<td>( \tau_{yy} )</td>
<td>solid stress in y direction</td>
</tr>
<tr>
<td>( \Psi(\tau) )</td>
<td>fraction of surface occupied by packets of age between ( \tau ) and ( \tau + d\tau )</td>
</tr>
<tr>
<td>( \omega )</td>
<td>frequency of vibration, Hz</td>
</tr>
<tr>
<td>( \Phi_s )</td>
<td>particle sphericity</td>
</tr>
</tbody>
</table>
1 INTRODUCTION

Fluidization is a phenomenon in which solid particles are kept in a floating state by an upward flowing gas or liquid. At a low flow rate, fluid passes through the void spaces between stationary particles. In this condition the bed remains fixed. With an increase in the flow rate, the bed expands slightly. At a critical gas flow rate the solid particles are just suspended in the upward flowing fluid. This condition is called minimum fluidization and the corresponding superficial velocity of the fluid is called minimum fluidization velocity. At this point the drag force exerted by the upward flowing fluid on the particles is just balanced by the weight of the particles. The pressure drop through any section of the fluidized bed equals the weight of the solid particles and fluid in that section. At higher flow rates, a liquid-solid fluidized bed exhibits a uniform expansion from the fixed bed to the hydraulic transport. However, with an increase in flow rate beyond minimum fluidization in the case of a gas-solid system bubbles or large voids begin to appear and movement of the solid particles becomes more vigorous. Such a bed is called bubbling fluidized bed. These bubbles form at the bottom of the bed near the distributor and rise through the bed coalescing and increasing in size. At even higher gas flow rates, the size of the bubbles can grow to the diameter of the bed column and at this stage the bed is said to be a slugging bed. At sufficiently high flow rate entrainment becomes
significant and the solid particles are carried out of the bed along with the fluid flow.

1.1 Liquid Like Behavior of a Fluidized Bed

A gas fluidized bed looks and in many ways behaves like a liquid. Light objects float on the surface of the fluidized bed and can be easily pushed under the bed surface. When a fluidized bed is disturbed, waves appear on the surface and are formed, propagated and reflected from the wall of the container and display interference just like waves on free surface of a liquid. The surface of the bed remains horizontal in a stationary vessel and becomes cylindrical when the vessel is rotated [Gelperin and Einstein 1971]. The kinetic energy of the solid particles in a fluidized bed is a function of the superficial velocity of the fluid being somewhat analogous to the case of a liquid where the kinetic energy of the molecules depends upon temperature of the liquid. Therefore, the parameter of a fluidized bed that is analogous to temperature of a liquid is the superficial velocity. The viscosity of a fluidized bed decreases with an increase in the superficial velocity just like the viscosity of liquid falls with a rise in temperature. Moreover, the viscosity of fluidized bed increases with an increase in the size of the solid particles similar to a liquid where the viscosity rises with increase in molecular size [Gelperin and Einstein 1971]. The transition of a fixed bed to a fluidized bed is similar to the melting of a solid body and entrainment of solid particles from a fluidized bed is similar to evaporation of a liquid. There are indications that near the minimum fluidization velocity, the fluidized bed behaves more like a pseudo-plastic rather than a Newtonian fluid as
occurs in certain liquids close to their melting point [Harrison and Leung 1961]: An analog to the Weissenberg effect in non-Newtonian liquids is observed by the formation of a mound on the free surface of a fluidized bed when it is mixed with a bladed mixer. These analogies between a fluidized bed and liquid exist only in the range of minimum fluidization to minimum bubbling velocity. When bubbling becomes vigorous the properties of a fluidized bed approximate those of a boiling liquid or a liquid through which gas is bubbled.

### 1.2 Brief History of Fluidization

The fluidization technique was first used commercially by Fritz Winkler for the gasification of coal in 1926. A number of similar units were constructed in Germany and Japan in order to supply raw gas for the synthetic chemical industry. During World War II, Standard Oil Corporation development the first fluid bed catalytic cracker (FCC) unit to produce high octane aviation gasoline. In 1944 Dorr-Oliver acquired rights to Esso’s fluidization know-how for use in fields outside of the petroleum industry such as roasting of sulphide ores. Moreover, Dorr-Oliver engineers successfully developed a drying and calcination process using fluidized beds in 1950s. The Sohio process for making acrylonitrile in a fluidized bed was extremely successful in the early 1960s. In early 1970s Union Carbide developed polyethylene synthesis process using fluidized bed technology. Also in the 1970s and 1980s fluidized bed combustion attracted much attention largely due to its relatively low temperature operation (800-900° C) and its ability to absorb \( SO_2 \) through the use of limestone or dolomite [Geldart 1986]. More than 2000 small atmospheric
pressure fluidized bed combustors are in use throughout the world on a variety of duties such as burning plastic waste, providing hot gases for drying grass and raising steam for process use [Highley and Kaye 1983].

1.3 Industrial Applications of Fluidized Beds

Numerous industrial processes make use of fluidized bed technology such as drying, surface coating, coating of pharmaceutical tablets, heat treatment, oil cracking, coal combustion, coal gasification, etc. Figure 1.1 shows many current industrial applications of fluidized beds arranged in five categories according to predominating mechanism [Geldart 1986].

1.4 Heat Transfer in Fluidized Beds

The most remarkable feature of the fluidized bed is its temperature uniformity produced by bubble induced solid mixing generated within the bulk of a bubbling gas fluidized bed. In practice this temperature uniformity exists in both radial and axial directions even in beds of ten meters in diameter [Kunii and Levenspiel 1969]. The heat transfer in a fluidized bed involves three modes of heat flow namely, fluid-to-particle, particle-to-particle and bed-to-surface heat transfer.

1.4.1 Particle-to-particle heat transfer

This mode of heat transfer is caused by the heat flow among particles due to the particle temperature difference. It plays an important role in many processes involving mixing of hot particles and cold particles. Particle-to-particle heat transfer
Figure 1.1: Classification of fluidized bed applications according to predominating mechanism [Geldart 1986]
may be considered as heat transfer through the following paths:

1. Radiation heat transfer.

2. Conduction through points of contacts between particles.

3. Heat absorption or release by the particles upon contacting with the gas serving as the heat transfer medium.

Radiation heat transfer is shown to be negligible below the temperature 600°C and heat conduction through the points of contacts between spherical particles and plane surfaces was experimentally found to be negligible [Botterill et al. 1973]. Therefore, heat transfer from particle-to-particle contact occurs mainly through a thin film of gas serving as the heat transfer medium.

1.4.2 Particle-to-gas heat transfer

Particle-to-gas heat transfer is due to forced convection heat transfer of the fluid to each fluidizing particle in a fluidized bed. Extremely large areas of solid particle surface are available to the fluidizing gas; e.g., a cubic meter volume of 100μm diameter particles has a surface area greater than 30,000m² [Geldart 1986]. Such a large surface area greatly enhances the particle-to-gas heat transfer.

1.4.3 Bed-to-surface heat transfer

The heat transfer between a surface and a fluidized bed may be consider to be composed of three additive components [Botterill, 1975].
1.4.3.1 Particle convection component  The particle convection component $h_{pc}$, is responsible for the high bed-to-surface heat transfer coefficient due to bubble induced particle circulation and very high volumetric heat capacity of the particle (of the order of a thousand fold higher than that of the fluidizing gas). Particles usually stay in the bulk of the bed for a long time to reach to bulk bed temperature by exchanging heat with the fluidizing gas and by conduction through the gas with other particles. Therefore when a packet of particles first arrives at a heat transfer surface, it is at the bulk temperature of the fluidized bed and there exists a high local temperature difference between the packet and the heat transfer surface. This packet of particles resides on a heat transfer surface for a short time and heat is transferred by unsteady conduction [Mickley and Fairbanks 1955].

1.4.3.2 Gas convective component  The interface gas convective component $h_{gc}$, is the heat transferred to the gas phase. This path is significant in the case of large particle fluidized beds and beds operating at high static pressure [Botterill, 1986]. As an example, for 160\,$\mu$m and 2500\,$\mu$m particles, the particle convective components were found to be 90% and 40% respectively [Baskakov and Vitt 1973]. The gas convective component was estimated experimentally by the mass transfer analogy [Baskakov and Suprun 1972].

1.4.3.3 Radiative Component  The radiative heat transfer component $h_{rc}$, becomes significant only at temperatures above 600\,°C. This was experimentally observed by using a gold probe and a copper probe [Botterill 1986]. The radiative heat transfer for an immersed surface in a bubbling fluidized bed must be
considered in two parts. When the surface is in contact with particles, radiative heat transfer mainly depends upon interaction between the surface and the first layer of particles. This layer is closed to the surface temperature than the bulk temperature. Successive layers of the particle do not significantly contribute to this heat transfer mode as each layer is almost opaque to the incident radiation. It was shown that a layer of 1.7 mm particles with the thickness equal to their diameter, lets through only 6% of incident radiation [Botterill 1975]. On the other hand when the heat transfer surface is engulfed by a bubble, the radiative heat transfer takes place between the surface and the particles which are at bulk temperature.

1.5 Present Investigation

The present investigation is directed towards establishing trends useful for the purpose of industrial design in heat transfer to submerged objects in gas fluidized beds through linear and oscillating motion of the object. The general purpose of this investigation is to study the heat transfer from a hot stationary and a moving (linearly downward and oscillating) sphere immersed in a glass-particle bed fluidized by air at room temperature and pressure. First, a systematic investigation was carried out in order to understand the significance of several important variables influencing the heat transfer rate from a stationary copper sphere such as: superficial velocity, glass particle diameter, and sphere diameter. Secondly, the effect of constant vertical downward motion and oscillating motion of the copper sphere (along with previously stated variables) on the heat transfer coefficient was experimentally investigated. In the case of a linear motion, the copper sphere was moved
at a constant velocity of 0.4, 1.1, 1.9, 3.0, 4.6 and 7.5 cm/s. Whereas, in the case of
oscillating motion, the copper sphere was placed about 10 cm above the distributor
plate in the fluidized bed and oscillated at various amplitudes of 1.8, 4.0 and 6.9
cm and at 1.1, 2.0 and 2.85 Hz frequencies. For all of the heat transfer studies,
5-44, 126-147 and 355-420 μm size glass particles and 1.0, 1.4 and 2.0 cm diameter
copper spheres were used in various combinations.

The transient temperature of the hot copper sphere as it is cooled in the flu­
idized bed was recorded by a computer controlled data acquisition system using
a thermocouple embedded at the center of the sphere. A sufficiently small Biot
number observed for all the cases enabled the use of lumped heat capacity theory
for cooling of the sphere. A transient temperature record was used to obtained heat
transfer coefficient. The heat transfer coefficients obtained for the case of a station­
ary sphere immersed in the glass bead fluidized bed were correlated to glass particle
size, superficial air velocity and copper sphere diameter. However, the primary goal
of this study was to correlate the enhanced heat transfer coefficient with the bed
parameters as described previously for linear and oscillating motion of the copper
sphere.

A theoretical model proposed by Gelperin and Einstein [1971] was used to pre­
dict the heat transfer coefficient in the case of linearly moving sphere in a fluidized
bed at minimum fluidization condition.
1.6 Significance of Present Investigation

The superior heat transfer characteristics as well as safer, cleaner and cheaper operational performance have made fluidized beds an alternative to conventional salt and lead baths that are used for heat treatment purposes. The use of fluidized beds for heat treatment of metal parts such as: hardening, annealing, carburizing, nitrocarburizing, bright tempering, etc. is a relatively new but well established industry in the United States, Canada and Europe. For example, for ferrous materials, neutral hardening involves first heating the metal to its austenitic state in an inert atmosphere, then rapidly cooling it to produce a martensitic structure. Conventional quenching is accomplished with oil, water, brine, air, salt or lead baths. Salt and lead baths contain cyanide poison in the molten salt, which must be properly handled and disposed off. Whereas, the fluidized bed bath uses inert sand like bed material, such as aluminum oxide along with inert gases such as nitrogen or argon gas. Furthermore, the fluidized bed can be turned off during shutdown periods to reduce running and maintenance cost. Quenching and heat transfer rates are an important part of these heat treatment processes. The present study provides new information into the heat transfer rate and the enhancement through controlled motion of a submerged sphere motion in a fluidized bed. Furthermore, the knowledge of heat transfer from such a moving object in a fluidized bed could be beneficial to controlling processes such as, freezing of food grains, physical operation involving particle coating, drying and gasification, and burning of a single particle in a fluidized bed combustor. The effect of controlled motion of a heat transfer surface immersed into a fluidized bed on heat transfer rates has been reported in only a
limited number of studies [Reed and Fenske, 1955].

1.7 Literature Survey

Heat transfer from an immersed object to a fluidized bed greatly depends on the hydrodynamics of solid particles in the vicinity of the object. Glass and Harrison [1964] visually observed the flow pattern near a horizontal tube in a two dimensional bed. The vertical component of gas velocity increases in the region close to the immersed object and exceeds the minimum fluidization velocity $U_{mf}$, even though the superficial gas velocity is less than $U_{mf}$. This phenomenon produces local fluidization near the immersed object. At higher superficial velocities, unstable gas voids forms underneath the obstacle, which acts as a source of small bubbles which are periodically detached [Ginoux et al. 1974 and Colver 1985]. However, there exists a defluidized region above the immersed object.

Several industrial processes utilize the high surface-to-bed heat transfer coefficient which have been found to be 20 to 40 times that of gases alone [Kunii and Levenspiel 1969]. The mechanism of heat transfer between a surface and a fluidized bed is complicated because of the large number of variables such as those of gas properties (density, viscosity, specific heat, thermal conductivity), properties of solid particles (diameter, density, sphericity, specific heat, thermal conductivity), conditions at minimum fluidization (superficial velocity, void fraction), geometrical properties (diameter of bed, geometry and orientation of heat transfer surface, spacing, static bed height, position of heat transfer surface in the fluidized bed) and mechanical action (vibration of fluidized bed, motion of the heat transfer surface).
The influence of various parameters on heat transfer coefficient as determined from the various observations can be summarized as follows:

1.7.1 Superficial gas velocity

The common findings from heat transfer in fluidized bed studies suggests that the average heat transfer increases rapidly near fluidization and either levels off [Yamazaki and Jimbo 1970], or drops as the average size of the bubbles increased at higher superficial gas velocities [Richardson and Shakiri 1979, Kharchenko and Makhorin 1964]. As early as 1949, Cibrowski [Zabrodsky 1966] studied heat transfer characteristics between walls of a deep air fluidized bed of alumina particles (44-55, 74-149 and 149-210 μm). The tube wall was electrically heated over a length of 1 meter. He found that the heat transfer coefficient increased with material concentration in the tube which depended on the superficial gas velocity. Mickley and Trilling [1949] studied the effect of solid concentration, glass particle diameter and superficial air velocity on heat transfer coefficient for an externally heated and an internally heated fluidized bed. Glass particles ranging in size from 50 to 5000 μm were suspended in an upward flowing air stream at superficial velocities varying between 24.4 cm/s to 457.2 cm/s. The empirical correlation for an externally heated fluidized bed was given by,

$$ h = 0.0118 \left( \frac{\gamma_{bed}}{d_p^3} \right)^{0.263} $$  \hspace{1cm} (1.1)

and for an internally heated fluidized bed,

$$ h = 0.0433 \left( \frac{\gamma_{bed}}{d_p^3} \right)^{0.233} $$  \hspace{1cm} (1.2)
They noted that the air velocity exerts a direct influence on the heat transfer coefficient in the case of an externally heated fluidized bed. While in case of the internally heated system, the gas velocity does not exert any such effect on heat transfer coefficient and is independent of the solid concentration. Shirai et al. [1966] experimentally investigated heat and mass transfer from the surface of solid spheres fixed within a fluidized bed. They used active carbon, alumina, silica gel and sand particles and 8, 15, 28 cm diameter fluidized beds. The minimum fluidizing velocities ranged from 0.8 to 10.5 cm/s. Five different sizes of brass and stell spheres were used ranging in size from 0.45 to 5 cm in diameter, with and without an internal heater. The heat transfer coefficient was determined from the transient cooling curve of a sphere in case of a sphere without the internal heater. When the sphere had an inside heater, the electrical power consumption and the surface temperature of the sphere and the bed temperature where measured and used to calculate the heat transfer coefficient. Their results indicated that heat transfer coefficient was proportional to $(U/U_{mf})^{0.36}$ considering different materials of fluidizing particles. Whereas, Dow and Jacob's data [1951] showed that $h \propto U^{0.8}$ and Van Heerden et al. [1953] obtained $h \propto U^{0.45}$.

1.7.2 Particle diameter

The common results of most of the experimental studies suggests that the heat transfer coefficient decreases with an increase in particle diameter. In 1950 Baerg et al. [Zabrodsky 1966] carried out detailed experimental work on the maximum heat transfer coefficient between a heated wall and fluidized beds of iron filings, round
sand, molding sand, crushed silica gel, glass beads, cracking catalyst and alumina with particle sizes ranging from 60 to 878 μm, over a wide range of fluidizing air velocities. They proposed the following empirical correlation for $h_{\max}$,

$$h_{\max} = \frac{239.5 \log_e(7.05 \times 10^{-6} \rho_b)}{d_p} \quad (1.3)$$

Yamazaki and Jimbo [1970] investigated the heat transfer between a hot 1.2 cm steel sphere immersed in a fluidized bed of 8.1 cm inner diameter and 80 cm height. They used limestone, molding sand and fused alumina of 68 to 310 μm size. Heat transfer coefficients were calculated using the Heissler chart. It was shown that the heat transfer coefficient increased more rapidly for larger particles and maximum heat transfer coefficient was approximately proportional to the $d_p^{-0.5}$, without considering the effect of porosity. Richardson and Shakiri [1979] found that the maximum heat transfer coefficient was proportional to the $d_p^{-0.4}$. Gelperin and Einstein [1971] reported an initial decrease in the heat transfer coefficient with increasing particle diameter, then leveled off and increased slightly once again for very large particle diameter of 4.8 mm.

1.7.3 Size of the heat transfer surface

The heat transfer coefficient increases with a decrease in the immersed heater diameter while the heat transfer rate diminishes as the diameter of the heater is increased. Shirai et al. [1966] found that an increase in sphere diameter resulted in a decrease in the heat transfer coefficient at minimum fluidizing condition and heat transfer coefficient was found to be proportional only to the $-0.06$ power of the sphere diameter. They correlated their data in the following equation where $h$ is in
kcal/m²hr°C and \(d_p\) and \(D_t\) are in cm, within 30% deviation,

\[ h = 74\left(\frac{U}{U_{mf}}\right)^{0.36}(D_{sph})^{-0.06}(D_t)^{0.27} \]  

Vreedenberg [1952] found no regular relationship between heat transfer coefficient and the height of the heat transfer surface because of the considerable nonuniformity of the bed which was investigated.

Wicke and Fetting [Zabrodsky 1966] published their data on heat transfer between a wall and fluidized beds of carborundum, sand, aluminum, lead and glass particles. The particle size was varied from 65 to 3000 \(\mu\)m. They found that the heat transfer coefficient decreased with increase in height of the heating surface due to the heating of the stream of particles moving along its surface.

More recently, Botterill et al. [1984] measured heat transfer coefficient between spherical probes of different sizes and gas fluidized bed of alumina and sand in the size range 250 to 400 \(\mu\)m with operating condition extended up to 980°C. In the case of low Biot number probes (copper probe, 0.64, 0.87 cm diameter; phosphor bronze 0.89 and 0.95 cm diameter and gold 1.11 cm diameter) the lumped heat capacity method was used to determine the heat transfer coefficient. They also used large spherical probes of 3.16 cm in diameter the heat transfer coefficient was determined by numerically solving the unsteady conduction equation in spherical coordinates. They noted that for small spherical calorimeter probes, at lower values of the ratio of probe to particle heat capacity, enhancement in heat transfer was observed compared to the larger probes.
1.7.4 Static pressure of the bed

Vreedenberg [1952] investigated the effect of air pressure, position and height of the water cooled tubes on heat transfer coefficient in a large fluidized bed of 0.565 m diameter and 1.2 to 1.7 m in height. Sand was used as the fluidized material. the heat transfer coefficient was measured between the fluidized bed and water cooled horizontal and vertical tubes of 1.18 m in height and 34 mm diameter. He found that the heat transfer coefficient was independent of air pressure at a constant superficial air velocity.

Richardson and Shakiri [1979] carried out an experimental investigation of heat transfer from a small electrically heated element of 2.5 cm square teflon sheet wound with 78 turns of tungsten wire to a gas fluidized bed. Glass, diakon, synclyst and aluminum particles were used in the size range of 25 to 450 μm. It was observed that in most cases that the heat transfer coefficient improved as the static pressure of the bed increased in the range of 0.03 to 0.148 MN/m². The gas properties other than density varied only slightly over the range of pressure used. The increase in heat transfer coefficient with pressure was attributed to an improvement in the quality of fluidization [Botterill and Desai 1972] resulting from the formulation of smaller bubbles. This was also observed experimentally by the instantaneous temperature traces. They correlated the maximum heat transfer coefficient by the following equation;

\[ N_{max} = 0.30 A_r^{0.20} P_r^{0.40} \]  

Zabrodsky et al. [1976] cited an experiment which showed that a 7.5 fold reduction of pressure below atmospheric pressure diminished the maximum heat transfer
coefficient by 16% for fluidized beds of 200μm sized particles. This experimental finding was attributed to the gas film between the wall and contacting particle is important for bed-to-wall heat transfer and at low pressure the property of this thin gas film changes in such a way that reduction in heat transfer coefficient is resulted.

Recently, Saxena and Ghanza [1985] studied the effect of pressure on heat transfer coefficient for immersed surfaces in fluidized beds by examining the various correlations and theories developed. They proposed a new classification scheme utilizing the concept of Archimedes number. They summarized their conclusions as follows:

1. For particle system characterized by $Re_{mf} < 40$ and $Ar < 130,000$, minimum fluidizing velocity and heat transfer coefficient are almost independent of pressure.

2. For particle system characterized by $40 < Re_{mf} < 200$ and $130,000 < Ar < 1.6 \times 10^6$, $U_{mf}$ and $h$ are dependent on pressure.

3. For particle system characterized by $Re_{mf} > 200$ and $Ar > 1.6 \times 10^6$, $U_{mf}$ and $h$ becomes proportional to the static pressure of the gas to the 0.5 and $h$ is dependent on pressure.

1.7.5 Particle sphericity

The heat transfer characteristics for irregular shaped aluminum dust particles fluidized by air were studied by Richardson and Shakiri [1979]. They found that the heat transfer coefficient was very low near and above $U/U_{mf} = 2$ because of
channelling. Kharchenko and Makhorin [1964] noted that sharp edged clay particles gave lower values of maximum heat transfer coefficients compared to sand particles under similar conditions.

1.7.6 Position of the heat transfer surface

Shirai et al. [1966] showed that heat transfer coefficient remained almost constant in the bed except in the part where the influence of the perforated plate predominates. Whereas above the settled height of the bed, the heat transfer coefficient reduces abruptly as the bed becomes dilute. In the radial direction, there was no change in heat transfer coefficient except very near to the wall. A short critical review of many experimental studies was reported by Zabrodsky et al. [1976]. They commented that heat transfer coefficient for an internal surface is higher than the external wall at low gas flow rates. They attributed this behavior to the amount of gas collected at the bottom of a body and deflected along its sides causing additional local aeration. This aeration promotes solid motion which in turn increases the heat transfer coefficient. Vreedenberg [1952] found no effect of the position of the heat transfer surface in the bed because of the considerable nonuniformity of the bed which was investigated.

1.7.7 Gas conductivity

The effect of gas thermal conductivity on the heat transfer coefficient was studied by Jackob and Osberg [1957]. They investigated heat transfer between a submerged electrically heated horizontal wire and a fluidized bed of glass particles
ranging in size from 30 to 290\(\mu\)m. The wire diameter (0.132 mm) was of the same order of magnitude as the particle diameter. Eight different gases were used ranging in thermal conductivities from 0.00715 to 0.15 Kcal/m hr\(^{2}\)C. The results showed a direct proportionality between heat transfer coefficient and the gas thermal conductivity. Richardson and Shakiri [1979] conducted heat transfer experiments using air, hydrogen, carbon dioxide and freon-12. The results showed that, the heat transfer coefficients for hydrogen in both fixed and fluidized beds were considerably higher than for the other gases because of its high thermal conductivity.

1.7.8 Baffled fluidized bed

An interesting study of heat transfer in free beds and beds divided into ten sections thermally insulated from each other by horizontal screens was carried out by Massimilla and Bracale [Zabrodsky 1966]. They observed that at lower superficial air velocities the heat transfer coefficient in the free bed was higher than that of a restricted bed, while at higher superficial air velocities, the opposite was true. On the basis of their results, they concluded that, in the region of high flow rates where the particle velocities are quite high, the screen prevented an increase in the bed voidage and that in turn improved the heat transfer.

1.7.9 Bed diameter

Heat transfer coefficients were measured between a hot silver sphere of 2 cm diameter and a fluidized beds of 82.5 and 157 mm diameter by Varygin and Martyushin [Zabrodsky 1966]. Various fluidized particles were used such as: ferrosilicon,
hermatite, carborundum, quartz sand and glass particles ranging in size from 82.5 μm to 1160μm. They found no observable effect of bed diameter on the heat transfer coefficient. Shirai et al. [1966] conducted heat transfer experiments using 8, 15 and 28 cm diameter beds and found that heat transfer coefficient increased with increase in fluidized bed diameter. The heat transfer coefficient was approximately proportional to the $D^{0.27}$.

1.7.10 Bed temperature

Dow and Jacob [1951] carried out an experimental investigation of heat transfer between a fluidized bed and a steam jacketed wall. They found the presence of a core in which the radial temperature remained almost constant and a thermal boundary layer of gas and particles along the wall which was responsible for the main thermal resistance to heat transfer between the wall and the bed. An experimental investigation of heat transfer between a spherical copper alpha calorimeter 6 cm in diameter and a high temperature (300 to 1050°C) beds of quartz sand and fire clay fluidized by flue gases was carried out by Kharchenko and Makhorin [1964]. The bed diameter was 22 cm and unexpanded height of the fluidized bed was kept at 10 cm. They determined the heat transfer coefficient from the transient heating curve of the copper sphere. Their results showed that the heat transfer coefficient increased with the bed temperature and the maximum heat transfer coefficient was approximately linear with the bed temperature. This was caused by the change in physical properties of the gas. They concluded that since the heat transfer coefficient varied linearly with the bed temperature, the radiant component of heat
transfer was unimportant. Botterill et al. [1984] found that a higher heat transfer coefficient was obtained when a hot probe at $800^\circ C$ was immersed in the bed at $350^\circ C$, compared with the case of a cool probe at room temperature.

1.7.11 Heat transfer surface temperature

Botterill et al. [1984] showed that increasing the surface temperature of the heat transfer probe with a constant bed temperature increased the bed-to-surface heat transfer coefficient because of the change in the gas thermal conductivity adjacent to the surface. Whereas, Abuaf and Gutfinger [1974] concluded that, within experimental uncertainty, the heat transfer coefficient was independent of the initial temperature of the object over a initial temperature range of $175^\circ C$ and $50^\circ C$.

1.7.12 Particle thermophysical properties

Ziegler et al. [1964] measured the heat transfer coefficient from a sphere and a cylinder to a bed of solids fluidized in an air stream. Copper, nickel and solder (50% Pb and 50% Sn), whose densities were nearly the same but the thermal properties that varied over a reasonable range were used as fluidizing materials. The celite sphere was electrically heated with a thermocouple soldered to the surface of the sphere. From the heat input and the temperature difference between the heat transfer surface and the bed, the heat transfer coefficients were obtained. The heat transfer coefficient was found to increase with solid heat capacity and was independent of solid thermal conductivity. Copper and nickel bed particles whose thermal conductivities vary widely, but has similar heat capacities, produced almost
the same heat transfer coefficient. However, increasing heat capacity of the particles, the heat transfer coefficient increased by 50 to 100%.

Regarding the effect of particle thermal conductivity, Zabrodsky et al. [1976] noted that for short residence time of the packet of particles on the heat transfer surface that the heat wave penetrates only the particles next to the wall and a change in the thermal conductivity of particle could be important for heat transfer. However if the first row of the particles are separated by a gap of 0.1 \( dp \) from the wall, there was no influence of the particle conductivity.

1.7.13 Local heat transfer coefficient on a tube or sphere perimeter

Samson [1974] experimentally studied the variation in the local surface heat transfer coefficient on a large horizontal cylinder, a vertical cylinder and a sphere immersed in a correspondingly large fluidized bed using small heat transfer plugs. The cylinder measured 11.5 cm in diameter by 11.5 cm long and the sphere 11.5 cm in diameter and were made up of bronze. The fluidized bed 46 cm in diameter by 65 cm length was used with sand particles of 141\( \mu \)m mean diameter fluidized by air. He plotted heat transfer coefficient on polar coordinate graphs, showed the detrimental effects of a defluidized zone formed above the objects and a gas film adjacent to the underside of the objects. In the case of the horizontal cylinder the maximum local surface heat transfer coefficient was observed to occur at 72° from the underside or bottom for both low and high gas flow rates. However, this angle was found to be 90° in a separate study conducted by Gelperin et al. [1963]. This discrepancy can be explained by the presence of a considerably wider defluidized zone in the latter
studies since the diameter of the cylinder used was six times smaller in size. The variation in local radial heat transfer coefficient was observed to be significant along the radial angle at the low fluidizing rate for a vertical cylinder. This was attributed to the lack of uniformity in the fluidization over the bed cross section. However at higher superficial air velocities, the variation in the radial heat transfer coefficient became less marked. In the case of the immersed sphere, the general pattern of the variation in the heat transfer coefficient was found to be similar to that for the horizontal cylinder. The maximum local heat transfer coefficient occurred at 70° from the base. The heat transfer coefficient at the base of the sphere was found to be lower than for the cylinder owing to the presence of gas pockets collected at the base of the object before moving upwards. Since the collecting surface is smaller for a sphere than a cylinder, resulted in thickening of the gas film. The photographs obtained by him of the defluidizing and gas void zones are shown in Figure 1.2.

Baskakov et al. [1972] measured the temperature field on the surface of a sphere submerged in a fluidized bed. This temperature distribution was used for calculating the local heat transfer coefficient. They used 4.0 cm and 8.2 cm diameter spheres and 60 and 320 μm corundum particles. They observed the maximum local heat transfer coefficient at the equatorial zones of the sphere and the heat transfer coefficient at the lower portion of the sphere was found to be higher than for the upper portion.
Figure 1.2: Defluidized and gas film zone, Gelperin et al. [1963]
1.7.14 Orientation of heat transfer surface

Baskakov et al. [1973] carried out experiments to measure the heat transfer coefficient for an inclined plate with the heated surface up and heated surface down immersed in the same fluidized bed. The heat transfer coefficients for the top surface of an inclined plate were lower from those obtained with a vertical plate because of a defluidized layer formed above the plate. Furthermore, heat transfer coefficients to vertical tubes are usually 5 to 15% higher than to horizontal tubes as the latter have inferior conditions of contact with the fluidized bed [Gelperin and Einstein 1971].

1.7.15 Moisture

The effect of particle moisture content was studied by Baskakov et al. [1973]. In the case of a bed of nonabsorbent corundum particles, heat transfer coefficient first increased with increased moisture content because of the reduction in inter particle electrostatic effects within the bed. With a further increase in the moisture, the air became saturated and the flow properties of the bed changed sharply and the heat transfer coefficient was reduced. For porous charcoal particles, particle heat capacity increased as moisture was absorbed and for 13% absorption of moisture by weight, the heat transfer coefficient increased by 26%.

1.7.16 Mixed particle size

The effect of mixed particle size distribution on heat transfer in a gas fluidized bed was studied by Figliola et al. [1986]. They used an electrically heated cylinder
(5 cm diameter by 25.4 cm length) immersed in a fluidized bed of 200 cm height and 30.5 cm x 30.5 cm cross section of glass particles. They found that the heat transfer coefficient was augmented by 25% at G=0.23 Kg/m\(^2\)s to 40% at G=0.30 Kg/m\(^2\)s with the addition of fine particles to 453\(\mu\)m particles reduced the mean particle size by about 11%. However beyond 340\(\mu\)m mixture heat transfer coefficient practically remained constant even though the mixture mean particle size was decrease by 33% further.

1.7.17 Packed bed height

Pillai [1976] investigated heat transfer to a sphere immersed in a shallow fluidized bed. He recorded the transient temperature of a sphere of low Biot number when immersed into heated air fluidized beds. Heat transfer coefficient were measured at bed temperatures of up to 1100°C. Particles of silica sand, zircon sand and silicon carbide ranging in size from 200\(\mu\)m to 800\(\mu\)m were investigated. The heat transfer coefficient was found to be somewhat higher than those reported for deeper beds. He noted major features of shallow beds as the violent coalescence of bubbles, thermal entry region, and a jetting region next to the distributor plate which occupied most of the shallow bed. He suggested that solids motion within a shallow bed might be significantly greater than in a deep bed reducing the residence time of the particles at the immersed surface. He also concluded without any physical explanation that the bed temperature dependence of the maximum heat transfer coefficient was very much stronger in a shallow bed which he expressed in
an empirical correlation;

\[ N_{u_{\text{max}}} = 0.365A_r^{0.22} \theta^{0.82} \]  \hspace{1cm} (1.6)

Kharchenko and Makhorin [1964] carried out experiments to establish the relation between the height of the unexpanded bed and the heat transfer coefficient for quartz sand particles of 340\(\mu\)m mean diameter and superficial air velocities ranging from 65 to 70 cm/s at a bed temperature of 800 \(^\circ\)C. It was found that as the bed height was changed, the heat transfer coefficient reached an optimum value at an initial bed height approximately equal to the bed diameter. While at other values of bed height, the heat transfer coefficients were lower than the optimum values.

1.7.18 Gas convective component

Botterill et al. [1981] performed experiments with spherical probes of low Biot number immersed in a hot 18.8 cm diameter stainless steel fluidized bed operating at temperatures ranging from 250\(^\circ\)C to 700\(^\circ\)C. Sand, coal and alumina particles of mean diameter size range from 380\(\mu\)m to 2320\(\mu\)m were used. They observed that bed-to-surface heat transfer coefficients were sensitive to the superficial velocity with group 'B' materials. Whereas for group 'D' materials, gas convective heat transfer was more important and there was less influence of gas velocity. They also observed a transition in behavior between the characteristics of a bed of group 'D' to 'B' type material as the operating temperature increased.

Shirai et al. [1966] estimated the amount of heat transferred to the particles, by subtracting the amount of heat transferred by the gas phase from the experimental heat transfer coefficient. The heat transfer coefficient for the gas phase was
estimated using the Ranz and Marshall [1952] correlation. From this estimation, they concluded that about 88% of heat is transferred by the particles. Similar results were obtained by Ziegler and Brazelton [1964]. They performed experiments in which simultaneous heat and mass transfer from a 1.27 cm diameter sphere of clay-like material was studied in a gas fluidized beds of copper, glass and alundum particles. The systems were chosen so that the fluidized particles had a capacity for heat transport but not mass transport. O'Brien et al. [1985] measured mass and heat transfer coefficients for a horizontal tube submerged in a fluidized bed of glass particles ranging in size from 215\(\mu\)m to 3400\(\mu\)m. The gas convection component was determined by measurements of the rate of mass loss from a submerged naphthalene cylinder and utilizing the heat and mass transfer analogy. Total heat transfer coefficients were measured under identical conditions using an instrumented and electrically heated copper cylinder. They found that the gas convective coefficient increased significantly with particle size and demonstrated an increased dependence on superficial gas velocity for the large particles. The relative contribution of the gas convection to total heat transfer ranged from 6.8% for the smallest particles at optimum heat transfer to 100% for the largest particles.

1.7.19 Vibrated bed

In recent years, the vibrated fluidized bed has gained significant attention for its ability to overcome certain drawbacks associated with the conventional fluidized bed such as, gas bypassing, attrition and entrainment of particles. It has been found that vibrated fluidized beds offer higher heat and mass transfer rates than
conventional fluidized beds at lower operating pressure drops. Moreover, suspending the equipment on a well-designed resonating spring assembly, the electrical energy for vibration can be maintained to a minimum [Mujumdar 1983]. A vibrated fluidized bed is particularly useful for heat transfer of immersed objects in a bed of hard to fluidized and sticky granular solids. Some of the vibrated fluidized bed applications include the drying of granulated fertilizers, polymer chips, inorganic and organic chemicals, pharmaceuticals, food products (e.g., sugar, cocoa, coffee), asbestos fibers, clay, etc.

Recently Malhotra and Mujumdar [1987] presented results of an experimental study of heat transfer from an immersed electrically heated copper cylinder to a vertically vibrated fluidized bed of glass particles ranging in size from 325 μm to 2360 μm. A two-dimensional fluidized bed of rectangular cross section was vibrated by means of an eccentric mechanism at frequencies varying from 0 to 105 rad/s and amplitudes up to 0.425 cm. An interesting result was reported, the heat transfer coefficient increased by more than 20 times for a packed bed vibrating at accelerations, $\frac{4\omega^2}{g}$ near 2. This enhancement was attributed to vigorous particle mixing. Similar heat transfer enhancement was obtained for aluminum oxide, kaoline and zinc particles when vibrated moderately in nonaerated beds. For fluidized beds, as the superficial velocity increased, the maximum heat transfer coefficient occurred at lower values of vibrational acceleration due to improved particle mixing induced by air flow rate. Malhotra and Mujumdar [1987] also studied the effect of surface stickiness by adding controlled amount of glycerine into the glass particles. The heat transfer coefficients were found to be about 40 to 50% higher for those sticky
particle beds as compared to the dry beds for vibrational accelerations below 4. However for a vibrational acceleration above 4, the difference between heat transfer coefficient for sticky bed and dry bed vanished. In fact, for glass particles of 1017 μm, the dry bed produced a slightly higher heat transfer coefficient than the sticky bed at vibrational acceleration above 2 and $U/U_{mf} = 0.9$.

Lu et al. [1976] conducted an experimental study of heat transfer between a stationary sphere and a vibrated fluidized bed. Also the particle motion by using tracer technique was investigated. Polystyrene beads of 1840 μm size were used as fluidizing particles. The test spheres were made up of aluminum of 1.3 and 3.1 cm diameters, which were inserted into the hot bed and the transient temperature of the center of the sphere was recorded. The heat transfer coefficient was determined using the lumped heat capacity method. The results of the pressure drop across the vibrated fluidized bed suggested that the bed could be fluidized at a lower gas flow rate than that of a unvibrated system. Heat transfer experiments were carried out in two sets. In one set, the amplitude was kept constant at 1.0 cm and frequencies were varied as: 65, 120, 210, 260, 480, 840 rpm. In an another set of experiments, the frequency was kept constant at 65 rpm and the amplitude was varied from 1 to 0.3 cm. The heat transfer coefficient was found to increase with amplitude as well as frequency compared to the stationary bed at lower superficial gas velocities. However, for Reynolds number based on fluidized bed diameter and superficial velocity greater than 1000, the effect of vibration of the heat transfer coefficient was negligible. Their results also showed that additional gas flow rate did not increase the heat transfer coefficient of the system if its vibrational acceleration was greater
than 3. Furthermore, they estimated the residence time of the packet of particles by measuring the particle velocities. Mickley and Fairbanks' [1955] packet model was used to predict the heat transfer coefficients from these residence time data. The agreement between the theoretical and experimental heat transfer coefficients were very reasonable in the higher vibrational acceleration region and was poor in the lower vibrational acceleration region.

Abuaf and Gutfinger [1974] performed an experimental study of the heat transfer and coating in a vibrated fluidized bed. Nylon particles of mean diameter of 125μm size were used in a 11.4 cm diameter and 25.6 cm high (packed bed height=15.5 cm) fluidized bed. The heat transfer experiments were performed with copper plate of dimensions 6.4 ×5.4×0.3 cm³ with attached thermocouples and heat transfer coefficients were determined by the cooling temperature-time history record. This work showed the spread about 15 to 22% in heat transfer coefficient at a constant superficial gas velocity. This result confirmed an important aspect of fluidized bed heat transfer regarding the difficulties in obtaining reproducible heat transfer data. They attributed this spread to the dynamic conditions of the fluidized bed.

1.7.20 Vibrating plate

Reed and Fenske [1955] in an effort to attain higher heat transfer rate conducted an experimental study of measuring the heat transfer rate from a vibrating plate immersed in a gas fluidized bed. Nickel powder, lead powder and carbon granules of 26, 38, 880 and 550 μm average particle diameters were used. They found that
below 1 ft/s superficial air velocity and at an amplitude of 0.313 in. with vibrational frequencies up to 2000 cycles/min, the vibrating motion of the plate was effective in enhancing the heat transfer rate. This enhancement was more pronounced for the nickel particles, somewhat less for the lead particles, and still less for the large carbon granules. At the high gas flow rates of 3.8 and 7.5 ft/s used to fluidize the steel shot, the vibrating motion of the plate, had practically no effect on heat transfer coefficient. The results of the horizontal vibrating plate immersed in a fluidized bed of 100 μm granular material showed that as the amplitude of the vibration increased from 0.158 cm to 0.795 cm, the heat transfer coefficient increased rapidly with the frequency of oscillation. They plotted the product of frequency and amplitude vs heat transfer coefficient between a vibrated plate and bed of aerated nickel powder and showed that at 0.2 ft/s superficial velocities, the data for 0.0625 and 0.125 inch stroke length coincided. However, the 0.313 inch stroke line fell below the curve obtained at the lower strokes. This was attributed to bed density at the plate region being lower than the average in the bed and at such a large stroke and high frequencies, the plate probably sweeps out a volume sufficiently large and rapidly so that the particles do not immediately fill the space left by the motion of the plate.

1.7.21 Freely moving sphere

In recent years, studies of heat transfer from a freely moving object in a fluidized bed has gained special attention. Knowledge of such heat transfer data is essential in some physical operations such as; drying, gasification and combustion of fossil fuels.
Prins et al. [1986] evaluated heat transfer coefficient for a freely moving sphere in a gas fluidized beds of glass particles of 1000μm size and alumina particles of 631μm size at elevated temperatures of 300°C, 600°C and 900°C. Silver and graphite spheres of 0.4 to 2.0 cm in diameter were used as heat transfer probes, which were connected to very thin long and flexible thermocouples so that they could move almost freely through the entire bed. The fluidized bed used had 10.3 cm internal diameter and 12 cm of packed bed height. The bed was maintained at 300°C using hot air for glass particles bed and at 600°C and 900°C for fused alumina particles bed. The cold sphere was immersed in the hot fluidized bed and the transient temperature of the center of the sphere was recorded to determine the heat transfer coefficient utilizing the low Biot number aspect of the system. It was shown that the difference in the heat transfer coefficient between a fixed silver sphere and a freely moving graphite sphere was relatively small and tended to become less for beds of large sized particles. For smaller sized particles the heat transfer coefficient for a freely moving graphite sphere was smaller than that of a fixed silver sphere for the fluidized bed of glass particles. They correlated their final results within 15% accuracy, within the range; 10 < Ar < 2000 and 3 < D_{sph/d_p} < 200. For a fixed sphere;

\[ Nu_{max} Ar^{-m} = 4.175 \left( \frac{D_{sph}}{d_p} \right)^{-0.278} \]  

(1.7)

where,

\[ m = 0.087 \left( \frac{D_{sph}}{d_p} \right)^{0.128} \]  

(1.8)

For a freely moving sphere;

\[ Nu_{max} Ar^{-n} = 3.539 \left( \frac{D_{sph}}{d_p} \right)^{-0.257} \]  

(1.9)
Where,
\[ n = 0.105 \left( \frac{D_{sph}}{d_p} \right)^{0.082} \]  

(1.10)

Rios and Gibert [1984] studied heat transfer from a fixed object, a light object freely floating and circulating in the entire layer, and a light object only allowed to move in a restricted zone in an air fluidized bed of glass particles ranging in size from 315-400 \( \mu \)m. Light plastic spheres of different sizes 0.8 to 3.5 cm in diameter were used as large freely moving objects. In some cases they were allowed to circulate throughout the bed in others they were permitted to move in the upper part of the bed delimited by inserting a large mesh wire lattice 4 cm below the upper surface. Heat transfer coefficients were determined from the heat stored from several cold spheres that were plunged into the hot fluidized bed during a given time. Experimental determination of this stored heat was carried out using a calorimeter. Their results showed that for 0.8 and 2 cm diameter plastic spheres, without a wire lattice within the layer, the heat transfer coefficient increased continually over the gas velocity range from the minimum fluidizing velocity \( U_{mf} \) to \( 2U_{mf} \) and then become independent of superficial air velocity. They also compared heat transfer coefficients for stationary, freely moving and hindered moving spheres at \( U = 2.3U_{mf} \) for various sphere sizes. One unexpected result obtained was that the heat transfer coefficient for a stationary sphere were independent of sphere diameter at \( U_{mf} \) to \( 2U_{mf} \). The heat transfer coefficients for fixed and freely moving 0.8 cm diameter spheres were quite different. The heat transfer was enhanced by 68% for a freely moving 0.8 cm diameter plastic sphere and 32% for hindered moving plastic sphere. This result was quite contrary to that of Prins et al. [1986]. However, for larger
plastic spheres, e.g., 3.5 cm diameter the heat transfer coefficient for stationary, freely moving and hindered moving plastic spheres were found to be very similar.

1.8 Theoretical Models

Three basic trends has been observed in modeling the heat transfer phenomena in fluidized beds.

1.8.1 Conductive heat transfer through the gas boundary layer

In this model heat is conducted through the gas boundary layer near the heat transfer surface. This gas layer was assumed to be scoured by solids moving along the heat transfer surface decreasing the boundary layer thickness near the wall and increasing the heat transfer coefficient. Such a model has been developed by Leva [1959], Dow and Jacob [1951] and Levenspiel and Walton [1954].

Levenspiel and Walton [1954] developed a model where the resistance to the heat flow in due to a laminar gas layer which is destroyed by the scouring action of solid particles passing through it. The average thickness of the laminar gas layer is much thinner than that of a tube without solid particles. They assumed stationary particles of uniform diameter $d_p$ arranged in equally spaced horizontal layers. As shown in the Figure 3, the boundary layer is destroyed at the points of contact with the stationary solid and develops again. The distance between two successive layers of solid particles is given as:

$$D_T = \frac{\pi d_p}{6(1 - \epsilon)}$$ (1.11)
Figure 1.3: Laminar boundary layer thickness between two layers of particles close to the heat transfer wall [Levenspiel and Walton 1954]
The heat transfer coefficient for an average boundary layer thickness;

\[ h = \frac{K_g}{\delta} \]  \hspace{1cm} (1.12)

The boundary layer thickness can be given by;

\[ \delta = 3.333 \left( \frac{\mu g D_r}{U \rho g} \right)^{0.5} \gamma \]  \hspace{1cm} (1.13)

were

\[ \gamma = \left[ \left(1 + \beta^2\right)^{1.5} - \beta^2 \right] \]  \hspace{1cm} (1.14)

and

\[ \beta = 0.041(1 - \epsilon) \left( \frac{D_r U \rho g}{\mu g} \right)^{0.5} \]  \hspace{1cm} (1.15)

Though this model agreed well with the experimental data for air fluidized beds of coal, glass and silica catalyst. This model does not take into account the influence of the solid particles on the heat transfer and hence the mechanism can not be considered complete.

1.8.2 Unsteady heat conduction by particles

Photographic observations show that solid particles tend to associate in groups and remain in contact with the heat transfer surface for a certain amount of time [Botterill et al. 1962]. They considered an isolated solid particle in contact with the heat transfer surface. By neglecting conduction through the points of contacts, radiation effects, and thermal convection through the surrounding fluid, the unsteady heat conduction between a plane surface and an insulated spherical particle surrounded by a static gas was addressed. The system of Laplace equations for
heat transfer within the two homogeneous phases, namely fluid and solid particles with residence times up to $70\mu s$ was solved numerically. Because of the low thermal conductivity of the gas, the rise in the temperature of the particles was very small throughout the residence time of the particle. The calculated spacing of the isotherms showed that significant heat transfer is confined to the region around the point of contact between the particle and surface, where the paths are very short. However, heat entering the particle is rapidly conducted because of high thermal conductivity of the solid phase. The average rate of heat transfer to a single particle during a given residence time was calculated by integration of the instantaneous rate. The overall heat transfer rate was estimated from the average rates taking into account the number of particles in contact per unit area of the heat exchange surface and scaling to unit temperature difference. The experimental results, obtained by varying the residence time of the particles at the surface by stirring the bed compared well with this model for short residence times. In the case of long residence time, the heat penetrates further into the bulk of the bed and the second row of particles also play an important role in the heat transfer mechanism. Botterill [1975] extended this model to the case of heat transfer into two adjacent particles touching the surface in order to compare the experimental results obtained at large residence times.

Gabor [1970b] proposed two unsteady state heat transfer models relating residence time to the rate of the heat transfer based on a string of particles normal to the heater surface and characteristic length functionally dependent on the depth of heat penetration from the heater surface. The first model was based on a string
of spheres of indefinite length normal to the heater wall when viewed from the side
and in an hexagonal pattern when viewed from the heater surface. He made the
following assumptions in this model:

1. An orthorhombic array is a reasonable approximation of the array of the
   particulate phase of both fluidized beds and packed beds. The void fraction
   for such an array is 0.395.

2. As the volumetric heat capacity of solid particles is many orders of magnitude
greater than the volumetric heat capacity of gases, the heat removal by gas
convection can be neglected.

3. The radial heat transfer between particles is neglected.

The system of heat conduction equations was solved numerically for air-glass and
air-copper systems. A grid of five divisions per particle radius and incremental time
interval of $t/d_p^2$ of $1.455 \times 10^{-3}$ and $3.89 \times 10^{-4} \, hr/ft^2$ was used. The heat
transfer flux was determined by weight averaging the temperature of each particle
and then multiplying this temperature by the particle heat capacity and dividing by
the total time. The heat transfer coefficients calculated from this model agreed well
with the experimental data for glass and copper particles in air with the assumption
that the gas gap between particles is $0.015d_p$ and $0.0075d_p$ between the wall and
the first layer of particles. The second model was based on heat transfer through a
series of alternating gas and solid slabs. The thickness of the solid slab was taken as
$0.66d_p$ and the thickness of the gas phase between the particles was taken as $0.1d_p$
and $0.05d_p$ respectively as suggested by Kunii and Smith [1960]. These values of
gas slab thicknesses gave a good agreement with published experimental data. The main drawback of this model is that it cannot be applied to a fluidized bed with strong agitation and bubbles.

More recently, Ganzha et al. [1982] developed a mechanistic theory for the heat transfer between an immersed surface and a fluidized bed of large particles in the absence of radiation. The conduction heat transfer coefficient was computed by considering a composite of infinite layers of gas and solid. For a large particle system, it was assumed in this model that all of the resistance to heat transfer is confined to the first row of particles near the heat transfer surface only. The heat is transferred by conduction through the gas film between the surface and the particle with a uniform thickness of \( \delta \). Particles were replaced by equivalent cylinders having the same volume as the actual particles and of equal height and diameter. The transient one dimensional heat conduction equations for each phase were solved using appropriate initial and boundary conditions. The approximate (within 30 \%) solution valid for all the gases except helium and hydrogen in the temperature range of 273 to 1100K and pressure range 0.1-10 MPa was given as:

\[
h_{\text{cond}} = 1.06 \frac{K_g}{\delta} \tag{1.16}
\]

They modified this equation to account for the bed voidage and developed an equation for the thickness of the gas-film for a curved heat transfer surface as the gas film thickness is larger for the case of a curved surface. The final equation was given as:

\[
Nu_{\text{cond}} = 1.02 \frac{(1 - \epsilon)0.666}{(0.114 + \frac{h_c}{d_p})} \tag{1.17}
\]

The convection heat transfer coefficient was calculated recognizing the fact that for
beds of large particles, the Reynolds number is invariably larger than 100 and the flow around the particle is turbulent [Leva 1959]. The turbulent boundary layer formed on the immersed surface is continuously disturbed by the front half of the bed of particles and formed again in their wake [Levenspiel and Walton 1954]. The heat transfer surface was regarded as being covered with a continuous arrangement of unit orthorhombic cells which in time keeps reforming as new particles arrive at the surface. The heat transfer from the surface through the turbulent boundary layer was considered similar to that of a flat plate immersed in a turbulent gas stream. The heat transfer from a plate placed in a turbulent fluid flow is given by:

\[(Nu_l)_{conv} = CRe_l^{0.8}Pr^{0.43}\]  

(1.18)

For a unit orthorhombic particle arrangement they [Levenspiel and Walton 1954] obtained an expression for \(l\) as;

\[l = 0.451d_p(1.0 - \epsilon)^{2/3}\]  

(1.19)

\[(Nu_l)_{conv} = CoRe^{0.8}Pr^{0.43}\left[\frac{(1 - \epsilon)^{0.133}}{\epsilon^{0.8}}\right]\]  

(1.20)

The constant \(Co\) was obtained by simultaneous measurement of total heat transfer coefficient and void fraction. Ganzha et al. [1982], performed an experiment with a single 1.3 cm diameter vertical tube and staggered bundles in a restricted bed at superficial gas velocities. This experiment gave a mean value of \(Co\) as 0.12.

**1.8.3 Packet theory**

Mickley and Fairbanks [1955] developed this model which is generally known as packet theory. They introduced a picture of a packet of particles at the bulk bed
temperature is swept into close proximity with the heat transfer surface under the bubble induced circulation pattern occurring within the bed. The packet of particles remain in contact with the heat transfer surface for a certain period of time called the residence time of the packet only to be swept away from the surface and replaced by a fresh packet of emulsion. When this packet first arrives close to heat transfer surface, there is a high local temperature difference between the packet and the surface, consequently heat transfer is rapid. The longer the residence time, the more closely the surface and local bed temperatures approach and this decreases the heat transfer rates. Their reasoning for this packet theory was based upon the experimental evidence that the dense phase bubbling beds retained a constant void fraction independent of superficial gas velocity and that this void fraction was essentially that of quiescent beds. In a dense fluidized bed each particle may be expected to be in contact with several neighbors most of the time. The packets are not permanent; they have finite persistence in time. Their void fraction, density, heat conductivity and heat capacity are assumed to be the same as those of the quiescent bed. The main features of the packet theory as shown in Figure 1.4.

Consider a packet of particles at bed temperature $T_b$, swept into contact with a flat surface of temperature $T_w$. Unsteady conduction will commence on contact. If $A_c$ is the contact area between the packet and the wall and considering a homogeneous packet, the instantaneous heat flux after time $\tau$ is given by:

$$q_{wi} = A_c \left( \frac{K_{eff} \rho_f m_f C_p}{\pi \tau} \right)^{1/2} (T_w - T_b)$$

(1.21)
Surface
Fresh element sweeps away emulsion at the surface

Unsteady state conduction into emulsion element at surface

Heated element leaves the surface

Figure 1.4: Packet theory model
The local instantaneous heat transfer coefficient is given by,

\[ h_{wil} = \left( \frac{K_{eff} \rho_m C_p}{\tau} \right)^{1/2} \] (1.22)

If \( \psi(\tau) \) is the fraction of the surface occupied by packets of age between \( \tau \) and \( \tau + d\tau \), then the local time averaged heat transfer coefficient;

\[ h_{wl} = \int_0^\infty h_{wil}(\tau)\psi(\tau)d\tau \] (1.23)

\[ h_{wl} = \left( K_{eff} \rho_m C_p S_L \right)^{1/2} \] (1.24)

where,

\[ S_L = \pi^{-1} \left[ \int_0^{\infty} \psi(\tau) \tau^{-1/2} d\tau \right]^2 \] (1.25)

and the average heat transfer coefficient over an entire isothermal area is,

\[ h = \frac{1}{A} \int_A h_{wl} dA \]

\[ (K_{eff} \rho_m C_p S)^{1/2} \] (1.26)

\[ S^{1/2} = \frac{1}{A} \int_A S^{1/2} dA \] (1.27)

Here \( S \) is a stirring factor representative of the frequency of particle replacement at the surface. Mickley and Fairbanks (1955) applied this derivation to two idealized bed models.

1. Slug flow of solids past the surface In a bed which is operating at low superficial gas velocities, there is not much turbulence, and solid particles are observed to move upward at the center and downward along the outside walls.

If the solids move downward with a constant velocity \( u_p \), the residence time
of the packet $r$ is given by:

$$
\tau = \frac{L_H}{u_p}
$$

(1.28)

where, $L_H$ is the length of the heater. For this condition, the heat transfer coefficient is:

$$
h = \left[ \frac{4K_{eff} \rho_m f C_p}{\pi} \left( \frac{u_p}{L_H} \right) \right]^{1/2}
$$

(1.29)

2. **Side mixing** With a large surface and highly turbulent beds, side wise transfer of the packet of particles occurs. The packet is flowing down while some of the surface is exchanged with solid brought in sideways from the core of the fluidized bed. When the side mixing is predominant, the heat transfer coefficient is given by:

$$
h_{wl} = \left( K_{eff} \rho_m f C_p \right)^{1/2}
$$

(1.30)

where $s$ is the average replacement of packets at the wall per unit time by means of side mixing, and $S_L = s$.

Mickley et al. [1961] experimentally measured instantaneous and time averaged heat transfer coefficient in a fluidized bed. The low values of instantaneous heat transfer coefficient were attributed to gas bubbles, and high values to the sudden contact with a fresh packet of emulsion. One essential deficiency of this model is that the instantaneous heat transfer coefficient is inversely proportional to the square root of the residence time of the packet. Hence at very low values of residence time the heat transfer coefficient becomes extremely high.
1.8.4 Packet model including contact resistance

Baskakov [1964] extended the packet model with an added contact resistance to heat transfer located at the wall. This prevented the heat transfer coefficient for small residence times from becoming infinite. The expression for an added contact resistance was given as:

$$\frac{1}{h_c} = \frac{d_p}{\pi K_g \left[ \ln \left( \frac{K_p}{0.1 K_g} - 1 \right) \right]}$$

(B.31)

Baskakov [Gelperin and Einstein 1971] assumed this contact resistance to be independent of time and in series with the thermal resistance of the packet itself $R_a$. The wall resistance $R_w$ was given by;

$$R_w = \frac{\delta_w}{K_{ew}}$$

(1.32)

$\delta_w$ is the extent of the zone near the wall and it is of the order of $dp/2$. Experimental results showed that the surface-to-bed heat transfer is influenced by gas thermal conductivity and the particle diameter. This dependence can be accounted for by expressing the contact resistance as,

$$R_w = \frac{d_p}{m K_g}$$

(1.33)

For a cubic packing of the particles at the surface, Koppel et al. [1970] obtained $m=2\pi$. Zabrodsky et al. [1981] predicted from their work $m$ to be of the order of 7.2. Baskakov and Vitt [1973] obtained from their data $m=8$, and Botterill et al. [1962] used $m=10$ to make their two particle model agree with experimental predictions. Catipovic et al. [1980] also used $m=6$, and Xavier and Davidson [1985]
found that $m=4$ was more appropriate for a horizontal surface. Thus in general $4 \leq m \leq 10$ has been suggested. The contact resistance has been assumed to be in series with the actual thermal resistance in the packet of particles, hence,

$$h_i = \frac{1}{R_w + R_a}$$  \hspace{1cm} (1.34)

where,

$$R_a = \left( \frac{\pi r}{K_{eff} \rho_f m C_p} \right)^{1/2}$$  \hspace{1cm} (1.35)

The average heat transfer coefficient can be given by,

$$h = \frac{2}{R_a} \left[ 1 - \frac{R_w}{R_a} \ln(1 + \frac{R_a}{R_w}) \right]$$  \hspace{1cm} (1.36)

Patel [1967] presented two surface renewal models, which included both contact resistance and finite characteristic length.

1. **Model I** A packet with the same properties as the bed at the minimum fluidization condition, initially at $T_b$ comes in contact with the heat transfer surface to receive heat from the wall through a contact resistance. Heat penetrates up to a certain distance during the residence time.

2. **Model II** In this model it is assumed that heat from the heat transfer surface is transferred first to the single spherical particle near the wall and the particle near the wall also loses heat by conduction to a packet of particles of certain thickness situated between the wall particle and the bulk of the bed. Patel[1967] obtained asymptotic values for the heat transfer coefficients for zero and infinite residence time.
These various penetration models can be categorized as: having no contact resistance but a finite characteristic length [Toor and Marchellow 1958]; having no contact resistance and infinite characteristic length [Mickley and Fairbanks 1955]; having finite contact resistance but infinite characteristic length [Baskakov 1964]; with contact resistance and a finite characteristic length [Gutfinger and Abuaf 1974]. Furthermore in all these penetration type models, the residence time of the packet is to be estimated by experimentation. This can be achieved by: measuring the temperature fluctuations in the heater [Baskakov et al. 1973]; using a stirrer as heat transfer surface [Botterill et al. 1962]; by tracing the particle movements [Wei-Ming Lu et al. 1976] or by moving the packed bed at various speeds. Moreover, the measurement of the fraction of time bubbles shroud the heat transfer surface has been used to correct the heat transfer coefficient.

The resistance near the wall region has been assumed as a first approximation to be independent of time. More recently, this resistance has been associated with the surface roughness [Decker and Glicksman 1981]. They assumed a representative model of surface roughness of the particle to properly deal with the conduction resistance. Furthermore, the heat transfer surface was assumed to be smooth as compared to the particle roughness. The region of possible solid contacts extended to the point on the particle where the average distance between the particle and the surface is equal to the surface roughness. With this assumption, the overall particle-to-surface heat transfer was found to be only modestly influenced by the magnitude of the roughness. Due to the limited number of contact points due to microscopic roughness elements, small size and limited heat capacity of the roughness asperities,
very high heat transfer rates occur for only the first 10 to 20 ms of the residence time. During this time the contact zone of the asperities approaches a mean temperature between that of the surface and the bed. This is followed by a longer period of slower heat transfer rate until the average temperature of the entire particle changes appreciably. This initial period of high heat transfer rate occurs so rapidly that its influence on the time averaged heat transfer is negligible.

Gloski et al. [1984] developed an experimental apparatus to measure the heat transfer rate for periods as short as 10 to 20 ms. A thin tin foil heater, which was rapidly heated to initiate the transient heat transfer from the heater to a fluidized bed of glass particles ranged in size from 1000 μm to 645 μm. From the transient record of the heater temperature, heat transfer coefficients were determined for discrete times (10, 20, 40, 60, 80, 100, 120 and 140 ms). For gas flow conditions slightly above and below the minimum fluidization, large heat transfer coefficients were obtained for glass particles with diameter between 650 to 1000 μm in the initial 10 ms, followed by a rapid decrease in the heat transfer coefficient during the first 10 to 20 ms. At longer times, the instantaneous heat transfer coefficient remained constant within the level of experimental uncertainty. From these results, they concluded that the elevated heat transfer coefficients at short times were due to the conduction heat transfer between the surface asperities on the particle which were in contact with the heat transfer surface, and that no gas layer existed separating the particles from the surface in fluidized beds. The constant heat transfer coefficient found after the initial 20 ms was attributed to the greater local resistance due to the greater local voidage.
These penetration models considered a packet of particles having constant voidage and constant thermophysical properties. Kubie and Broughton [1975] extended the packet theory to include the property variations described in terms of the voidage variations in the vicinity of a constraining surface. Most of the assumptions in their model are similar to those introduced by Mickley and Fairbanks [1955] except for the consideration of the variation of voidage normal to the wall and the influence of the surface on the local packing and hence alteration of the local thermophysical properties. They considered the heat transfer to be in the direction normal to the wall in the packet of emulsion and applied the semi infinite layer approximation. The conduction equation in one dimension was given as:

$$\rho(x)C_p(x)\frac{\partial T_e}{\partial t} = \frac{\partial}{\partial x} (K(x)\frac{\partial T_e}{\partial x})$$

(1.37)

With boundary conditions:

$$t \geq 0, \quad x = 0, \quad -K(x)\frac{\partial T_e}{\partial x} = f_s \quad t = 0, \quad x \geq 0, \quad T = 0$$

The voidage variation with distance from the constraining surface takes the form of a damped oscillation curve having a minimum voidage at about one particle radius from the surface. In the case of fluidized beds the oscillations of the voidage appears to be damped much more rapidly and the voidage minimum is shifted further from the constraining surface [Korolev et al. 1971]. Kubie and Broughton [1975] utilized this observation and expressed voidage variation in terms of distance from the constraining surface from simple geometrical considerations. They further assumed that at a distance of one particle diameter away from the constraining surface the void fraction is equal to that of bulk void fraction. The heat capacity of the gas was
neglected and the local effective thermal conductivity of a packet was expressed in terms of the particle conductivity, the local void fraction and the gas conductivity using the method proposed by Kunii and Smith [1960]. They solved the unsteady equation numerically using the voidage distribution and thermophysical property distribution. The results obtained with this model agreed well with controlled residence time data [Kubie and Broughton 1975]. This property boundary layer model required no physically unjustified concepts in order to produce general agreement with experimental data.

Recently, Chen and Pei [1985] developed a theoretical model of heat transfer between a fluidized bed and an immersed sphere and a cylindrical surface based on two-phase boundary layer and surface renewal theory. The following assumptions were made in this model:

1. A spherical or cylindrical heating surface has an average constant temperature and has radius R.

2. There is a concentric two-phase boundary layer of constant average thickness \( \delta_T \) surrounding the heat transfer surface.

3. The radial temperature derivative is zero at the outer boundary of the layer.

4. The thermophysical properties are constant.

5. Radiant heat transfer is negligible.

6. The two-phase thermal boundary layer is renewed from time to time by the moving particles and fluidizing gas.
The transient heat transfer problem described by the differential equation and boundary conditions in cylindrical and spherical coordinates, assuming azimuthal symmetry were solved by taking laplace transform. The solutions were simplified utilizing the fact that in a fully developed fluidized bed the particle velocity is large and \( R > d_p \) as,

\[
Nu = \frac{K_M d_p}{K_g} \left( \frac{S_1}{\alpha_M} \right)^{1/2}
\]

Based on their model, a correlation for predicting the maximum heat transfer coefficient was obtained which correlated experimental data well.

Yoshida et al. [1969] proposed a model for mechanisms of heat transfer between a fluidized bed and the wall surface, following the film penetration theory. They considered a finite characteristics length of emulsion packet \( l_e \), which suddenly contacts a heat transfer surface and after certain time it is replaced by a fresh packet of emulsion from the core portion of the bed. They represented this phenomenon by,

\[
\frac{\partial T_e}{\partial t} = \alpha_{eff} \frac{\partial T^2_e}{\partial x^2}
\]

with the boundary conditions;

\[
T_e = T_b \text{ at } t = 0 \quad T = T_w \text{ at } x = 0 \quad T = T_b \text{ at } x = l_e
\]

From the solution of above equation, the instantaneous local heat transfer coefficient \( h_i \) was found to be,

\[
h_i = \left[ \frac{K_\text{eff} \rho \text{eff} C_p}{\pi \tau} \right]^{1/2} \left[ 1 + 2 \sum_{i=1}^{\infty} \exp \left( -\frac{i^2 l_e^2}{\alpha_{eff} \tau} \right) \right]
\]
which converges rapidly and is useful for short residence times. Whereas for long residence time, the following equation is useful;

\[ h_i = \frac{K_{eff}}{l_e} \left[ 1 + 2 \sum_{i=1}^{\infty} \exp \left( \frac{2 \pi^2 \alpha_{eff} \tau}{l_e^2} \right) \right] \]  

(1.41)

The time averaged value of heat transfer coefficient was obtained from;

\[ h_w = \int_0^\infty h_i I(\tau) d\tau \]  

(1.42)

where \( I(\tau) \) is the age distribution function of emulsion packet on the surface. They used two types of age distribution functions.

1. Random surface renewal This is most representative of a surface in the main body of the bed continually contacted by rising bubbles. The age distribution of emulsion packet was is given by;

\[ I(t) = \frac{1}{\tau} e^{-\frac{t}{\tau}} \]  

(1.43)

Using this, for rapid replacement, \( \alpha_{eff} \tau/l_e < 0.8 \) within 20% error,

\[ h_w = \left[ \frac{K_{eff} \rho_{eff} C_p}{\tau} \right]^{\frac{1}{2}} \]  

(1.44)

for slow replacement, \( \alpha_{eff} \tau/l_e > 1.2 \), within 20% error,

\[ h_w = \frac{K_{eff}}{l_e} \]  

(1.45)

2. Uniform surface renewal This model is most representative of an emulsion flowing smoothly past the heating surface and the age distribution function was given by, for \( 0 < t < \tau \)

\[ I(t) = \frac{1}{\tau} \]  

(1.46)
For $t > \tau$

$$I(t) = 0 \quad (1.47)$$

For short contact times,

$$h_w = \frac{2}{\pi^{0.5}} \left( \frac{K_{eff} \rho_{eff} C_p}{\tau} \right)^{1/2} \quad (1.48)$$

For long contact times or steady state heat transfer,

$$h_w = \frac{K_{eff}}{l_e} \quad (1.49)$$

Gabor [1970a] developed a model predicting the rate of heat transfer from a flat plate and a cylinder immersed in a packed and fluidized bed at incipient fluidization. He solved steady state conduction equation for both the cases. The heat transfer coefficient for the flat plate,

$$h = \left( \frac{4K_{eff} C_p G}{\pi l} \right)^{0.5} \quad (1.50)$$

and for cylindrical heater,

$$h = \left( \frac{4K_{eff} C_p G}{\pi l} \right)^{0.5} + \frac{K_{eff}}{2al} \quad (1.51)$$

For $K_{eff}$, the following expression was used,

$$K_{eff} = K_e^{o} + 0.1(C_p d_p G) \quad (1.52)$$

where $K^{o}$ is the conductivity term without gas flow, predicted for an orthorhombic packing. The theoretical predictions for both the cylinder and flat plate heaters were in good agreement with his experimental data.
1.8.5 Residence time study

Baskakov et al. [1973] conducted an experimental study of heat transfer from a cylindrical probe (made up of a 5 micron platinum strip glued to the side of 1.5 and 3 cm diameter vertical rubber cylinder) immersed in a fluidized bed of corundum particles of mean diameter of 120, 320 and 500 \( \mu \text{m} \). The beds used were 9.8 and 9.2 cm diameter and the fluidizing gases were carbon dioxide, helium and air at temperatures from 20°C to 550°C. They determined the bubble contact time, packet contact time, bubble frequency and the fraction of the time that bubbles shrouded the heat transfer surface \( f_0 \) from the foil temperature fluctuations of the platinum strips mounted on the vertical tubes.

The contact resistance \( R_w \), taken as an inverse of the values of the instantaneous heat transfer coefficient was determined from the platinum foil probe. The contact resistance decreased almost linearly with the particle diameter. The gas resistance was found to be independent of the fluidizing velocity and height at which the probe was located within the bed. They obtained the empirical correlations for \( \tau \) and \( f_0 \) as follows:

\[
\tau = 0.44\left[\frac{d_{pg}}{U_{mf}(U/U_{mf} - A1)^2}\right]^{0.14}\left(\frac{d_p}{D_{probe}}\right)^{0.225} (1.53)
\]

\[
f_0 = 0.33\left[\frac{U^2_{mf}(U/U_{mf} - A1)^2}{d_{pg}}\right]^{0.14} (1.54)
\]

The empirical factor \( A1 \), made allowance for the effects of probe diameter and particle shape. From the frequency and amplitude of the temperature fluctuation of the heating element, Richardson and Shakiri [1979] estimated the mean residence
time of the solids at the surface and applied Mickley's packet theory to estimate heat transfer coefficients. These predicted heat transfer coefficients compared well with their experimental values.
2 EXPERIMENTAL SET UP AND PROCEDURE

2.1 Introduction

The aim of this investigation was to evaluate the heat transfer coefficients for a stationary and a moving (linearly and oscillating) sphere immersed in an air fluidized bed of glass particles for various superficial air velocities. The following measurements were required as part of the experiment: air flow rate, temperature history of the center of the copper sphere, air pressure at the inlet of the rotameter, pressure drop across the bed, linear velocity of the sphere, and frequency and amplitude of the oscillating sphere. A experimental rig was designed in order to acquire these parameters.

2.2 Experimental Setup

The schematic diagram and the actual photograph of the experimental set up are shown in Figures 2.1 and 2.2. The main components of the system are: the fluidized bed; the rotameters for air flow measurements; the linear and oscillating mechanisms for controlling the motion of the sphere; various manometers for measuring rotameter pressure and pressure drop across the bed; the copper sphere; a heating cup used to preheat the copper sphere and a computer controlled data
acquisition system.

2.2.1 Fluidized bed

The fluidized bed is shown in Figure 2.3. This bed was designed by the Chemical Engineering Department at Iowa State University to study the fluidization characteristics of fly ash. The bed consisted of 0.3175 cm thick plexiglas column with an inside diameter of 9.5 cm which served as a container for the bed of glass particles. There were three separate column sections of the same diameter which could be fastened together tightly by bolts and sealed with o-rings. The bottom two sections were both 20.3 cm high to allow convenient loading or unloading of the bed materials. The longest column section available was 3 feet high which could be use for deep bed fluidization experiments which serve as a transport disengagement section to keep particle elutriation to a minimum. However, this section was not used for the heat transfer study as the required length of the hollow steel pipe (surrounding the copper sphere and thermocouple) was too long for a sturdy system. The fluidization column was fitted with a porous disc at the base to act as an air distributor. The porous gas distributor plate employed in this work was half inch thick Alundum porous disc (mixture P236, 240 μm porosity), manufactured by ceramic division of the Norton Company. After the plexiglas column was set up with rubber gaskets in place, an attempt was made to check the uniformity of the air distribution within the column. Using glass particles in a very shallow bed, it was observed that the flow was initiated first around the inside wall at very low flow rates while particle movement was also observed simultaneously in the center region.
Figure 2.1: Schematic of the experimental setup
Figure 2.2: Photograph of the experimental set up
of the column. This was considered satisfactory fluidization in terms of a uniform air distribution. In order to check the leakage of air through the gasket and other joints, the fluidized bed was immersed in a large water tank and air was allowed to pass through the fluidized bed. In this way leaks were found near the bottom flange. A closer inspection of the rubber gasket revealed that it was not effectively sealing off the air bypass above the porous plate. To remedy this problem, a high vacuum grease manufactured by Dow Corning was used to seal the gaps at the gasket joints. Pressure taps were located along the column sections to measure the pressure drop across various sections of the fluidized bed. Fiberglass plugs were used to prevent particles from entering the manometer lines.

2.2.2 Flowmeters

Six rotameters of different ranges were used to measure the air flow rate. Appendix A describes the rotamer calibration equations and lists all the rotameters along with their range. All rotameters except rotamer number 6 were calibrated using wet test meters of appropriate range. The higher capacity rotamer number 6 was calibrated using the flow meter calibration rig located in the Chemical Engineering Department at Iowa State University.

2.2.3 Heater Cup

A metal heating cup connected to an A. C. rheostat was used to heat the copper spheres. This heater cup was placed on a swinging platform for easy removal of the heater cup after heating the sphere to a required temperature.
Figure 2.3: Schematic of the fluidized bed
Figure 2.4: Schematic of a motor driven gear mechanism for controlling sphere motion in linear direction inside the fluidized bed
2.2.4 Sphere Driving Mechanisms

The schematic of the motor and gear drive assembly for controlling the linear vertical motion of the sphere is shown in Figure 2.4. This mechanism was designed to move the copper sphere in a vertically linear direction inside the fluidized bed. A sliding aluminum block, holding hollow steel rod and the thermocouple and copper sphere assembly firmly and was allowed to slide freely over a set of two parallel polished shafts with the help of a nylon thread attached to the sliding block. This nylon thread was wounded around two driving shafts and an idle shaft as shown in Figure 2.4. This driving shafts were driven by a reversible D.C. variable speed motor connected through a gear train. The motor was driven by a dual D.C. power supply (Hewlett-Packard, 0-24 V). Various linear speeds of the sphere were obtained by varying the speed of the D. C. motor or changing the gear ratio so as to obtain the required torque. With this arrangement it was possible to move a sphere in the vertical downward direction for about 22 cm in the fluidized bed. Two stop and reverse switches were installed at the top and bottom end of the steel shafts to prevent any sudden impact of the sliding block with the brackets housing the steel shafts. Moreover a stop and reverse manual switch was also connected in between the D. C. power supply and the D. C. motor.

The schematic diagram of the mechanism used to oscillate the copper spheres is shown in Figures 2.5. As in the case of the linear motion in this arrangement, a sliding block made of aluminum was allowed to slide freely over a set of two polished steel shafts. This sliding block carried the hollow steel rod, thermocouple and sphere assembly, and was also connected to an aluminum wheel 8 cm in diameter.
Figure 2.5: Schematic of a sliding mechanism for controlling oscillatory sphere motion within the fluidized bed
and 3 mm in thickness. The aluminum wheel was mounted on a shaft of a gear train driven by a 0-24 volts D.C. motor (Electro Craft Corporation, Hopkins, MN). The wheel and the sliding block were connected by an adjustable connecting rod. This arrangement was similar in principle to that of a piston oscillating inside a cylinder with a connecting rod. The amplitude of the oscillating motion was varied by placing the connecting rod at various positions inside the wheel groove. The frequency of oscillation was changed by varying the speed of the D. C. motor with an appropriate gear ratio was selected for required torque.

2.2.5 Copper sphere

Three copper spheres 1.0, 1.4 and 2.0 cm in diameters were used for the heat transfer studies. Precision copper spheres (99.9 % purity) used in this study were obtained from Industrial Tectonics Inc., Dextor, MI. Copper was selected because of its high thermal conductivity, availability and its low cost. A small thermocouple wire (copper-constantan, 0.254 mm wire diameter) was passed through a long hollow steel tube having a 1.7 mm outside diameter and were soldered into a 0.6 mm hole extending all the way to the center of the copper sphere. The soldering procedure adopted here was one in which, an extremely small amount of silver solder was inserted into the hole of the copper sphere. Then the sphere was heated from the outside with a propane torch and as soon as the soldered appeared to have melted, the clean thermocouple bead was inserted inside the hole and very quickly the copper sphere along with the thermocouple wire was quenched in a water bath. The continuity of the thermocouple connection was checked by a multimeter. During
the experiments, frequent breakage of the thermocouple wire would occur because of fatigue. In this case, the soldering process was repeated using a clean hole and new clean thermocouple. The copper sphere was insulated but held rigidly to the end of the steel tube using epoxy.

2.2.6 Fluidized bed material

Glass microbeads (density=2.5 gms/cc, conductivity=0.0025 cal/sec/sq. cm, specific heat= 0.27 cal/gm/°C) were used as the fluidized bed material obtained from the Microbead Division of Cataphote Corporation, Jackson, Mississippi. These glass particles were sized by sieving with a mechanical shaker in three size groups; 5-44, 126-147 and 355-420 µm. This size range was also examined by taking the photographs of these glass particles as shown in Figures 2.6, 2.7, 2.8. From the measurement of the size of the glass particle in the photograph and knowing the magnification factor, the actual size of the glass particles were determined. The size range obtained this way was very close to the size range determined from sieving.

2.2.7 Data acquisition system

A computer controlled data acquisition system was designed and developed for this experimental work. It consisted of PRO 380 computer, an LA50 printer, an analog to digital module (Digital Corporation), a digital multimeter (Hewlett-Packard) and a reference junction compensator (Omega) as shown in Figure 2.9. In a somewhat different setting thermocouple readings were recorded directly by the ADM unit alone without a digital voltmeter using a modified interactive analog
Figure 2.6: Photograph of 5-44 µm glass particles, magnification factor=320
Figure 2.7: Photograph of 126-147 µm glass particles, magnification factor=64
Figure 2.8: Photograph of 355-420 μm glass particles, magnification factor=64
data module (IADM) program written in FORTRAN 77. This setup allowed high rates of data acquisition (up to 500 readings/second). However, at this rate and without using any integrating circuit for filtering the signal, the main signal was affected by noise. Though this noise did not seem to affect the final result in terms of the heat transfer coefficient, a digital voltmeter was connected to the IEEE bus bar of the ADM unit and new software was developed to link the communication network between the PRO 380 computer and the HP digital voltmeter. The HP voltmeter was set up in such a way that it integrated the signal for about ten power cycles. In this way, the noise in the original signal was eliminated. However, the data recording rate was reduced to 3 to 4 temperature readings per second. This rate was deemed to be adequate for this study.

2.3 Experimental Procedure

In a typical experimental run, the height of the packed bed was kept at about 24 cm. The bed was fluidized by passing air at a specified superficial air velocity and the rotameter reading for the pressure of the air entering the rotameter, the pressure drop across the bed and the height of the fluidized bed were measured and were entered into the interactive computer program developed for data acquisition. Next the copper sphere was preheated to approximately 180-200 °C in the electrically heated heater cup filled with glass particles. The heating cup was manually removed and the D.C. motor driving the linear mechanism was started. For a stationary sphere experiment, the motor was shut down when the sphere reached a predetermined height of about 10 cm above the distributor plate. This height
Figure 2.9: Computer controlled data acquisition system
corresponded to about half the bed depth of 24 cm. At this point, the data acquisition system was started quickly. The software program was set to stop the data acquisition as soon as the sphere temperature reaches to about 40 °C. The initial temperature of the sphere was always about 150 °C. This was achieved in some cases by heating the sphere to a higher temperature. In the case of the initial temperature at the time of the start of a test the data acquisition system was higher than 150 °C. The computer program was set to analyze data only at a temperature close to the specified value. All raw data in terms of thermocouple output voltage, including the rotameter reading, rotameter number, the pressure drop across the bed, the pressure at the outlet of the rotameter, the fluidized bed height, the packed bed height and the fluidized bed temperature were stored in a data file.

A computer program was written to reduce these data into an average heat transfer coefficient. The heat transfer experiment was repeated for various superficial velocities, three copper sphere diameters, and three glass particle sizes. Typically three to four runs were carried out for a constant superficial air velocity, to study the repeatability of the experiment. In some cases if the difference in average heat transfer coefficient was found to be greater than 15 to 20 % then five to six runs were made to improve the average value of the heat transfer coefficient obtained. In all of the heat transfer to a stationary sphere experiment about 600 runs were made. The average heat transfer coefficients for each set of runs were stored subsequently in the computer.
2.3.1 Linearly downward moving sphere

The mechanism shown in Figure 2.4 was used to move a sphere in a linearly downward direction. The data acquisition program was started and all required information such as: the sphere size, linear velocity, superficial air velocity, pressure drop across the bed, pressure at the rotameter outlet, bed temperature, and glass particle size were entered into the computer. As in the case of the stationary sphere experiments, the copper sphere was preheated. The supporting steel rod was adjusted in such a way that the copper sphere was placed just above the fluidized bed surface. At this point the electric motor was turned on and as soon as the sphere was immersed fully into the fluidized bed the data acquisition system was turned on by simply pressing the return key on the key pad of PRO 380 computer. The temperature at the center of the copper sphere was recorded at the rate of three to four readings per second automatically over a preset time fed into the computer depending upon the linear velocity of the moving copper sphere. This time was predetermined in such a way that the sphere could traverse a constant distance of about 22 cm inside the fluidized bed. The time over which the sphere was in motion (similar to the data acquisition time) varied from about 3 seconds for the sphere moving at 7.5 cm/s to about 35 seconds for the sphere moving at 0.4 cm/s. Table 2.1 lists all the variables used in heat transfer experiments for the linearly downward moving sphere. For each linear velocity and superficial air velocity, typically three to four runs were made for a particular sphere and glass particle size. In all about 3000 runs for the linearly downward moving sphere case were made. The average heat transfer coefficients for each set of runs was recorded.
Table 2.1: List of variables used in the study of heat transfer from a linearly downward moving sphere

<p>| | | | | |</p>
<table>
<thead>
<tr>
<th></th>
<th></th>
<th></th>
<th></th>
<th></th>
</tr>
</thead>
<tbody>
<tr>
<td>sphere diameter cm</td>
<td>1.0</td>
<td>1.4</td>
<td>2.0</td>
<td></td>
</tr>
<tr>
<td>Glass particle size μm</td>
<td>5-44</td>
<td>126-147</td>
<td>355-420</td>
<td></td>
</tr>
<tr>
<td>Range of superficial</td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>air velocity</td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>for 5-44 μm glass</td>
<td>0.04</td>
<td>to</td>
<td>2.2</td>
<td></td>
</tr>
<tr>
<td>particles, cm/s</td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Range of superficial</td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>air velocity</td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>for 126-147 μm glass</td>
<td>1.58</td>
<td>to</td>
<td>18.5</td>
<td></td>
</tr>
<tr>
<td>particles, cm/s</td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Range of superficial</td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>air velocity</td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>for 355-420 μm glass</td>
<td>11.5</td>
<td>to</td>
<td>35.8</td>
<td></td>
</tr>
<tr>
<td>particles, cm/s</td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Linearly downward</td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>velocity of sphere</td>
<td>0.4</td>
<td>1.1</td>
<td>1.9</td>
<td>3.0</td>
</tr>
<tr>
<td>cm/s</td>
<td></td>
<td></td>
<td></td>
<td>4.6</td>
</tr>
<tr>
<td></td>
<td></td>
<td></td>
<td></td>
<td>7.5</td>
</tr>
</tbody>
</table>
2.3.2 Oscillating sphere

A mechanism shown in Figure 2.6 was used to oscillate the copper sphere inside the fluidized bed of glass particles. The preheated sphere was dropped quickly into the fluidized bed and adjusted (by tightening a set screw on the steel tube) to oscillate at about 10 cm above the distributor plate. This corresponded to about half the bed height. At this time the electric motor and the data acquisition system was started simultaneously. The sphere was adjusted to oscillate at a particular frequency and amplitude. The temperature history of the copper sphere was recorded until the temperature dropped to about 40 °C from its initial temperature of about 150 °C. As explained previously the initial temperature of the copper sphere was adjusted to this value by trial and error. The frequency of the oscillation was varied from 1.1 to 2.85 Hz by varying the speed of the D.C motor. The amplitudes of the oscillation were varied from 6.9 cm to 1.8 cm by adjusting the position of the connecting rod inside a radial slot in the aluminum wheel. As in all the previous cases, about three to four runs were taken for the same conditions in order to study the repeatability of the experiment. Table 2.2 lists all of the variables used. About 3200 runs of heat transfer from an oscillating sphere were made. The average heat transfer coefficient for each set of runs was recorded.

2.4 Method of Evaluation of Average Heat Transfer Coefficient

In all the experiments, the Biot number of the heated spheres were kept sufficiently small so that the lumped heat capacity method was valid [Incropera and DeWitt, 1981]. Biot number can be defined as,
Table 2.2: List of variables used in the study of heat transfer from an oscillating sphere

<table>
<thead>
<tr>
<th>Sphere diameter cm</th>
<th>1.0</th>
<th>1.4</th>
<th>2.0</th>
</tr>
</thead>
<tbody>
<tr>
<td>Glass particle size $\mu$m</td>
<td>5-44</td>
<td>126-147</td>
<td>355-420</td>
</tr>
<tr>
<td>Range of superficial air velocity for 5-44 $\mu$m glass particles, cm/s</td>
<td>0.08 to 2.2</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Range of superficial air velocity for 126-147 $\mu$m glass particles, cm/s</td>
<td>1.58 to 14.8</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Range of superficial air velocity for 355-420 $\mu$m glass particles, cm/s</td>
<td>11.5 to 35.8</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Peak-to-peak amplitude, cm</td>
<td>1.8</td>
<td>4.0</td>
<td>6.9</td>
</tr>
<tr>
<td>Frequency Hz</td>
<td>1.1</td>
<td>2.0</td>
<td>2.85</td>
</tr>
</tbody>
</table>
Table 2.3: Maximum Biot number and response time for immersed copper sphere-fluidized bed system

<table>
<thead>
<tr>
<th>Diameter, cm</th>
<th>$h_{\text{max}}$, w/$m^2 K$</th>
<th>Biot no.</th>
<th>$\tau_c$</th>
</tr>
</thead>
<tbody>
<tr>
<td>Copper sphere</td>
<td>~ 900</td>
<td>0.0039</td>
<td>6.4</td>
</tr>
<tr>
<td>Copper sphere</td>
<td>~ 800</td>
<td>0.0048</td>
<td>10.0</td>
</tr>
<tr>
<td>Copper sphere</td>
<td>~ 800</td>
<td>0.0069</td>
<td>14.3</td>
</tr>
</tbody>
</table>
\[ Bi = \frac{hL_{ch}}{K_{cu}} \]  

(2.1)

where for spheres,

\[ L_{ch} = \frac{D_{sph}}{6} \]  

(2.2)

As summarized in Table 2.3 the worst case Biot number corresponded to the maximum heat transfer coefficient and the largest size copper sphere tested was found to be less than 0.007. Table 2.3 also shows the calculated values of characteristic response time of the sphere-bed system.

\[ \tau_c = \frac{\rho_{sph}C_{sph}V}{h_{sph}A_{sph}} \]  

(2.3)

This time constant is an indicator of the time of cooling required in a given run. For example, during the time corresponding to one time constant the copper sphere cools down to about 63.2%. Since the Biot number is much less the 0.1, the internal resistance within the copper sphere can be safely neglected. An energy balance on the hot copper sphere immersed into the fluidized bed can be given as:

\[ \rho_{sph}C_{sph}V \frac{dT}{dt} = h_tA_{sph}(T_b - T) \]  

(2.4)

where, for the sphere,

\[ \frac{V}{A_{sph}} = \frac{D_{sph}}{6} \]  

(2.5)

A linear relationship was used for the heat capacity of the copper sphere.

\[ C_{sph} = a + bT \]  

(2.6)
where, for copper,

\[ a = \frac{351.2 J}{KgK} \]

\[ b = \frac{0.1085 J}{KgK^2} (Valid \ for \ 27 - 327^\circ C) \]

Solution of equation 2.4 with the linear relationship for the heat capacity of the copper (equation 2.6) may be written as,

\[(a + bT_b) \left( \ln \left( \frac{T_b - T}{T_b - T_0} \right) \right) + b(T - T_0) = - \left( \frac{h_t A_{sph}}{\rho_{sph} V} \right) t \]  \hspace{1cm} (2.7)

Equation 2.7 enables the determination of the total heat transfer coefficient \( h_t \), averaged over time and sphere surface area from the slope of a plot of the left hand side of the above equation versus time. A regression fit was used in the computer program for the best slope. Correlation coefficients for the regression analysis were found typically to be 0.99 or greater.

The total heat transfer coefficient \( h_t \) was interpreted here to be that due to fluidized bed convection heat transfer coefficient \( h \) (particle convection and gas convection) and radiation heat transfer coefficient, \( h_r \).

\[ h = h_t - h_r \]  \hspace{1cm} (2.8)

where,

\[ h_r = \sigma \varepsilon_{sph} (T_0^2 - T_b^2) (T_0 + T_b) \]  \hspace{1cm} (2.9)

Example values for total heat transfer coefficient \( h_t \) and radiation heat transfer coefficient \( h_r \) are 245 W/m² K and 0.869 W/m² K, respectively. Here \( T_0 \) is the initial temperature of the sphere. It is assumed that the sphere is radiating to a
large black cavity. Thus the radiation heat transfer coefficient \( h_r \) considers the maximum possible radiation. In all experimental conditions the initial temperature range was low enough (about 160 to 140 °C) so that radiation was negligible.

A typical thermocouple output voltage versus time is shown in Figure 2.10 for a stationary 2.0 cm diameter sphere during cooling in a bed of 355-420 μm glass particles at a superficial velocity of 15 cm/s. The superficial air velocity is just above the incipient fluidization. Figure 2.11 shows the cooling trace for the same sphere at the same conditions moving linearly downward through the bed at a speed of 3.0 cm/s.
Figure 2.10: Typical thermocouple output versus time for 2 cm diameter stationary sphere.
Figure 2.11: Typical thermocouple output versus time for 2 cm diameter linearly downward moving sphere at 3.0 cm/s
3 RESULTS AND DISCUSSION

The experimental results for heat transfer from stationary, linearly downward moving and oscillating spheres are presented in this Chapter. The pressure drop across the bed was used to calculate the minimum fluidization velocity for each glass particle size. A detailed tabulation of these experimental data is presented in the Appendix.

3.1 Minimum Fluidization Velocity and Classification of Glass Powder

The pressure drop across the fluidized bed was measured as a function of the superficial air velocity for each glass particle size. The minimum fluidization velocity was determined experimentally using the pressure drop data as shown in Figures 3.1, 3.2 and 3.3. At the minimum fluidization condition, as stated in Chapter 1, the force balance on the bed is;

\[ (\text{Drag force by upward moving gas}) = (\text{Weight of particles}) \]  \hspace{1cm} (3.1)

The pressure drop across the bed at minimum fluidization condition can be given as:

\[ \frac{\Delta p}{L_m f} = (1 - \epsilon_m f)(\rho_p - \rho_g)g \]  \hspace{1cm} (3.2)
Figure 3.1: Bed-pressure drop vs. superficial air velocity for 5-44 μm glass particles
Particle Size=126-147 micron
Umf=1.6 cm/s

Figure 3.2: Bed-pressure drop vs. superficial air velocity for 126-147 μm glass particles
Figure 3.3: Bed-pressure drop vs. superficial air velocity for 355-420 μm glass particles
The experimental correlation for the pressure drop through packed beds of uniformly sized solids was obtained by Ergun [1952] as,

\[
\frac{\Delta p}{L} = 150 \frac{(1 - \varepsilon)^2}{\varepsilon^3} \frac{\mu g U}{(\phi_s d_p)^2} + 1.75 \frac{1 - \varepsilon}{\varepsilon^3} \rho g U^2 \phi_s d_p \tag{3.3}
\]

The first term in the above equation represents the pressure drop due to viscous energy loss while the second term is the loss from kinetic energy. The expression for the minimum fluidization velocity can be obtained by combining equations 3.2 and 3.3. Furthermore, it was found experimentally [Kunii and Levenspiel, 1969] that for a wide variety of systems \(1/\phi_s e_m^3 \approx 14\) and \((1 - \varepsilon_m)/\phi_s^2 e_m^3 \approx 11\). Using this result, the expression for minimum fluidization velocity reduces to,

\[
U_{mf} = \frac{d_p^2 (\rho_p - \rho_g) g}{1650 \mu g} \quad \text{for} \quad \text{Re}_p < 20 \tag{3.4}
\]

and

\[
U_{mf}^2 = \frac{d_p (\rho_p - \rho_g) g}{24.5 \rho_g} \quad \text{for} \quad \text{Re}_p > 1000 \tag{3.5}
\]

The Reynolds number based on the particle diameter was found to be much less than 20 for all three glass particle sizes used in the present study. Hence equation 3.5 was used to calculate the minimum fluidization velocities. The terminal velocity of the glass particles can be estimated by force balance on a single particle under steady state condition,

\[
U_t = \left(\frac{4d_p (\rho_p - \rho_g) g}{3C_D \rho_g}\right)^{\frac{1}{2}} \tag{3.6}
\]

Using the values of the drag coefficients according to the Reynolds number range, the expression for the terminal velocity can be given as,

\[
C_D = \frac{24}{\text{Re}_p} \quad \text{Re}_p < 0.4, \quad U_t = \frac{g(\rho_p - \rho_g)d_p^2}{18 \mu g} \tag{3.7}
\]
Table 3.1 lists the maximum particle Reynolds number, the minimum fluidization velocity as obtained from equation 3.5 and from the experiment, and the terminal velocity for each glass particle size.

As shown in Table 3.1, the difference between the minimum fluidization velocity as obtained from the equation 3.5 and by experiment is about 20%. Considering the fact that equation 3.5 utilizes two empirical correlations namely, from Ergun [1952] and Wen and Yu [1966], this discrepancy in the results is acceptable.

The hydrodynamic behavior of a fluidized bed depends on the properties of the powder used. The powders have been classified in four groups: C, A, B and D in the order of increasing particle size [Geldart, 1986]. Group C powders are very cohesive and normal fluidization is very difficult. Channeling occurs because the inter-particle forces are greater than the drag force of the fluid exerted on the particles. Generally particles less than 20 μm exhibits such behavior. The pressure drop across the bed is lower than the theoretical value (bed-weight per unit cross-sectional area). The heat transfer between a surface to fluidized bed is much poorer than with group A and group B powders because of poor particle mixing. Group A powders are characterized by considerable expansion of the bed between minimum fluidizing velocity $U_{mf}$, and minimum bubbling velocity $U_{mb}$. These powders are slightly cohesive. A further increase in the superficial velocity beyond $U_{mb}$ increases the size and the number of bubbles. Small number of bubbles produce rapid particle mixing and the surface of the bed resembles a boiling liquid. In
Table 3.1: Maximum particle Reynolds number, minimum fluidization and terminal velocity for each glass particle-air system

<table>
<thead>
<tr>
<th>Glass Particle size, μm</th>
<th>Maximum Reynolds number based on maximum U</th>
<th>Terminal Velocity cm/s</th>
<th>Reynolds number based on terminal velocity cm/s</th>
<th>Minimum Fluidization Velocity from experiment cm/s</th>
<th>Minimum Fluidization Velocity from equation 3.5 cm/s</th>
</tr>
</thead>
<tbody>
<tr>
<td>5-44</td>
<td>0.029</td>
<td>3.96</td>
<td>0.061</td>
<td>0.052</td>
<td>0.043</td>
</tr>
<tr>
<td>126-147</td>
<td>1.103</td>
<td>105.2</td>
<td>9.02</td>
<td>1.6</td>
<td>1.34</td>
</tr>
<tr>
<td>355-420</td>
<td>7.52</td>
<td>300.0</td>
<td>7.30</td>
<td>13.2</td>
<td>10.85</td>
</tr>
</tbody>
</table>
contrast to group A powders, group B powders have negligible inter-particle forces and formation of bubbles starts at or slightly above minimum fluidization velocity. Bed expansion is small and the bed collapses very rapidly when the gas supply is cut off. There is little powder circulation in the absence of bubbles and bubbles burst at the surface of the bed as discrete entities. Also most bubbles rise more quickly than the interstitial gas velocity and bubble size increases with both bed height and excess gas velocity $U - U_{mf}$. Group D powders are coarse and/or dense. All but the largest bubbles rise more slowly than the interstitial fluidizing gas so that the gas flows into the base of the bubble and out of the top, providing the mode of gas-exchange and by-passing which is different from the group A or group B powders. A quantitative description of these properties of the powders belonging to these groups is given in Table 3.2 [Geldart, 1986]. Recently a new class of powder belonging to group named AC has been characterized [Geldart, 1986] by the absence of a meaningful incipient fluidization point and the absence of a contraction of the bed when bubbles first appear.

The three glass powders used in this experiment have been classified in these four groups. Using the Figure 3.4 [Geldart, 1986] for $\rho_p - \rho_g = 2500 \text{ kg/m}^3$, it is found that the glass particles in size range of 5-44 $\mu$m with a mean diameter of 24.5$\mu$m belongs to group of A powders. The glass particles ranging in size 126-147 $\mu$m and 355-420 $\mu$m with mean diameters of 136.5$\mu$m and 387.5$\mu$m respectively, belong to group B powders.

Although the 5-44 $\mu$m glass particles with a mean diameter of 24.5 $\mu$m falls in the range of group A powder. As shown in Figure 3.4, it is very close to the boundary
## Table 3.2: Summary of group property [Geldart, 1986]

### Increasing size and density

<table>
<thead>
<tr>
<th>Group</th>
<th>C</th>
<th>A</th>
<th>B</th>
<th>D</th>
</tr>
</thead>
<tbody>
<tr>
<td>Most Obvious Characteristic</td>
<td>Cohesive, difficult to fluidize.</td>
<td>Bubble-free range of fluidization</td>
<td>Starting bubbles at 12 cm</td>
<td>Coarse Solids</td>
</tr>
<tr>
<td>Typical solids</td>
<td>Flour</td>
<td>Caking catalyst</td>
<td>Building sand</td>
<td>Crushed limestone</td>
</tr>
<tr>
<td>Typical Property</td>
<td>Cement</td>
<td></td>
<td>Table salt</td>
<td>Coffee beans</td>
</tr>
<tr>
<td>1. Bed expansion</td>
<td>Low when bed channels, can be high when fluidized</td>
<td>High</td>
<td>Moderate</td>
<td>Low</td>
</tr>
<tr>
<td>2. Deaeration rate</td>
<td>Initially fast, exponential</td>
<td>Slow, linear</td>
<td>Fast</td>
<td>Fast</td>
</tr>
<tr>
<td>3. Bubble properties</td>
<td>No bubbles, channels, and cracks</td>
<td>Splitting, recoalescence predominates; maximum size exists; large wake</td>
<td>No limit on size</td>
<td>No known upper size; small wake</td>
</tr>
<tr>
<td>4. Solids mixing</td>
<td>Very low</td>
<td>High</td>
<td>Moderate</td>
<td>Low</td>
</tr>
<tr>
<td>5. Gas backmixing</td>
<td>Very low</td>
<td>High</td>
<td>Moderate</td>
<td>Low</td>
</tr>
<tr>
<td>6. Slug properties</td>
<td>Solid slugs</td>
<td>Asymmetric</td>
<td>Asymmetric</td>
<td>Horizontal voids, solids slugs, wall slugs</td>
</tr>
<tr>
<td>7. Spouting</td>
<td>No</td>
<td>No except in very shallow beds</td>
<td>Shallow beds only</td>
<td>Yes, even in deep beds</td>
</tr>
<tr>
<td>Effect on properties</td>
<td>Mean Particle size within group</td>
<td>Cohesiveness increases as dp decreases</td>
<td>Properties improve as size decreases.</td>
<td>Properties improve as size decreases.</td>
</tr>
<tr>
<td>Particle size distribution</td>
<td>Not known</td>
<td>Increasing less than 45 micron fraction improves properties</td>
<td>None</td>
<td>Increases segregation</td>
</tr>
<tr>
<td>Increasing pressure, temperature, viscosity, density of gas</td>
<td>Probably improves</td>
<td>Definitely improves</td>
<td>Uncertain, some possibly</td>
<td>Uncertain, some possibly</td>
</tr>
</tbody>
</table>
Figure 3.4: Diagram for powder classification into groups (Geldart, 1986)
of powder A and powder C. Moreover, a study of the fluidization characteristics of this powder revealed the presence of extensive channeling in the bed up to a superficial air velocity of 0.1 cm/s, then resembling more like C powder. At higher superficial velocities, this powder showed reasonably good fluidization behavior as a group A powder.

The remaining two powders sizes of 126-147 and 355-420 μm, clearly behaves as group B powders. As expected for this group, bubbling starts very close to the minimum fluidization condition.

3.2 Stationary Sphere

Experimental investigation was carried out to study the effect of sphere size, superficial air velocity and glass particle size on the average heat transfer coefficient for a stationary sphere immersed in an air fluidized bed. All experiments were carried out with the copper sphere positioned at about 10 cm from the bottom of the bed. The experimental values of time averaged heat transfer coefficient h are plotted (Figures 3.5, 3.6 and 3.7) against the superficial air velocity for each sphere size and glass particle size. Most of the systems showed similar trends and their behavior could be characterized by the three zones, referring to Figures 3.6 and 3.7 for the fluidized bed of 126-147 μm and 355-420 μm glass particles.

1. In the first region where the superficial air velocity is less than that of minimum fluidization velocity, the bed behaves almost like a packed bed. The glass particles are not in motion with respect to the other particles or the immersed sphere. In this region, the heat transfer coefficient was found to
Figure 3.5: Heat transfer coefficient vs. superficial air velocity for 5-44 μm glass particles and various size copper spheres
Figure 3.6: Heat transfer coefficient vs. superficial air velocity for 126-147 μm size glass particles and various size copper spheres
Figure 3.7: Heat transfer coefficient vs. superficial air velocity for 355-420 μm size glass particles and various size copper spheres
be very low as there is no circulation of the particulate phase. Although it seems that the heat transfer coefficient is almost constant, close examination of the numerical values revealed that the heat transfer coefficient decreased slightly as the superficial velocity increased until it reached the minimum fluidization velocity. This slight decrease in the heat transfer coefficient might be explained as follows: the expansion of the bed caused by the upward flowing air increases the void fraction and, thus, decreases the effective thermal conductivity of the bed in the region close to the heated copper sphere.

2. In the second region, where the ratio of superficial air velocity to the minimum fluidization velocity varies from 1 to 6 as in the case of 126-147 µm glass particles and from 1 to 2 for the case of 355-420 µm particles, the heat transfer coefficient rises very rapidly. The particle movement increases as small bubbles form within the bed. These small bubbles increase the circulation of the particulate phase near the sphere, decreasing the residence time of the emulsion packet.

3. At a higher superficial velocity, the heat transfer coefficient begins to level off. In this region, large bubbles appear in the bed and the bed behaves very close to a slugging bed as the bubble diameter grows to the same size as the column diameter of the fluidized bed. As the size of the bubble increases, the copper sphere is engulfed by more and more bubbles. During the time when the sphere is surrounded by a bubble, only gas phase heat transfer takes place. The resulting low thermal conductivity and heat capacity of the air phase, compared to the glass particles, reduces the overall heat transfer coefficient
even though the increased frequency of bubbles decreases the residence time of the emulsion packet.

The above trends explained in terms of three regions applies, more or less to all three sizes of the copper spheres and the glass particle sizes of 126-147 \( \mu m \) and 355-420 \( \mu m \). For the smallest size glass particles tested (5-44 \( \mu m \)) the observed behavior is slightly different. For the ratio of \( U/U_{mf} \) up to about 2, the heat transfer coefficient is very low and remains almost constant as the inter-particle forces produce severe channeling while prevents smooth fluidization. This is followed by a gradual increase in the heat transfer coefficient up to the point when the superficial air velocity reaches about 0.7 cm/s. In this region small bubbles start to form in the bed giving rise to a local particulate circulation. As the superficial air velocity increases, the frequency and the size of the bubbles increases and the surface movement of the bed starts. At this point the heat transfer coefficient rises rapidly without a maximum or minimum. At about a superficial air velocity of 1.5 to 2.0 cm/s, attrition of small size glass particle starts and bursting of large bubbles near the upper surface causes vigorous bed movement. This increases the heat transfer coefficient continuously with superficial velocity as circulation of the particulate phase becomes very effective for this size of glass particles. At a superficial velocity close to or greater than that of the terminal velocity of the particles, leveling off or even a decrease in the heat transfer coefficient could be expected as the particulate phase becomes leaner.

Kharchenko and Makhorin [1964] reported maxima in the heat transfer coefficients when plotted versus superficial velocity, an effect that was more pronounced
with an increase in the bed temperature for a spherical alpha-calorimeter of 6 cm in diameter. The existence of a maximum heat transfer coefficient is well established trend particularly for internal objects such as plates, cylinders and heat transfer probes [Davidson and Harrison, 1971]. However, Yamazaki and Jimbo [1970] showed only a leveling off of the heat transfer curve with an increase in the superficial velocity for a 1.2 cm steel sphere immersed in a fluidized bed of limestone or molding sand tested over a variety of particle sizes and values of $U/U_{mf}$ up to 45. Ziegler and Brazelton [1964] obtained almost constant heat transfer coefficients for the range of $U/U_{mf}$ of 3 to 7 for a half inch clay-like sphere immersed in a fluidized bed of glass particles. Shirai et al. [1966] also showed that the heat transfer coefficient leveled off above the value of $U/U_{mf}=10$, for a 3 cm diameter sphere immersed in a fluidized bed of active carbon, alumina, silica gel and sand particles, having minimum fluidization velocities from 1.6 cm/s to 10.5 cm/s. More recently, similar trend was obtained by Pillai [1976] for spheres ranging in size from 5 to 15 mm in diameter and a fluidized bed of silica sand, zircon sand and silicon carbide at 550 °C. The heat transfer coefficients were found to be almost constant above $U/U_{mf}=3$. This absence of a maximum heat transfer coefficient and subsequent decrease in heat transfer coefficient seems to be yet another trend as can be inferred from the four previous research works cited and the present study, particularly for the case of a spherical object immersed in an air fluidized bed.

The effect of diameter of the copper sphere on the average heat transfer coefficient is shown in Figures 3.5, 3.6 and 3.7 for various glass particle sizes. Figures 3.6 and 3.7 show clearly a decrease in the average heat transfer coefficient with increase
in the sphere diameter. This increase in heat transfer coefficient with decrease in sphere diameter is more predominant at superficial air velocities higher than that of minimum fluidization velocity. This decrease in heat transfer coefficient will be explained qualitatively. The residence time of an emulsion packet on any surface immersed in a fluidized bed can be defined as;

\[ \tau = \frac{L}{u_p} \]  \hspace{1cm} (3.9)

where \( L \) is the projected or equivalent length of heat transfer surface and \( u_p \) is the relative velocity of the emulsion packet with respect to the heat transfer surface. For a smaller copper sphere, the emulsion packet has to travel less distance to cover the projected length of the sphere and hence the residence time of the packet is less. This increases the heat transfer coefficient for a small sphere as compared to a larger sphere. Shirai et al. [1966] showed that heat transfer coefficient decreases with an increase in sphere diameter varying from 0.45 to 5.0 cm in an air fluidized bed. Moreover, Prins et al. [1985] found that the mass transfer coefficient decreased with increasing test sample diameter up to a certain limit which was roughly characterized by a gas contact time of 25 microseconds. This gas contact time was defined as;

\[ t_{cg} = \frac{\epsilon_m f D_{sp}}{U_{mf}} \] \hspace{1cm} (3.10)

At superficial velocities below that of minimum fluidization, the particulate phase circulation within the bed is almost absent, making the residence time of the emulsion packet very large. Hence, in this region the effect of sphere diameter is not prominent as can be seen from the Figures 3.5, 3.6 and 3.7. Furthermore, from Figure 3.5, it is evident that this normal trend of decreasing the heat transfer co-
efficient with increasing sphere diameter is not observed in the lower superficial air velocity region.

The effect of the mean powder size of the bed particles can be seen in Figures 3.8, 3.9 and 3.10 for various sizes of the copper spheres. In general the average heat transfer coefficient is higher for smaller size particles. The increase is a result of the higher mobility of the smaller size particles which in turn reduces the residence time of the emulsion packet. Furthermore, decreasing the particle diameter causes an increase in the heat transfer coefficient due to the reduction in the gas film between the body and the aggregates of particles. In a fluidized bed the heat transfer is viewed in terms of two resistances in series, a) a contact resistance $R_c$ due to an increase in voidage near the heat transfer surface and b) the resistance inside the emulsion packet itself. For large particles the contact resistance becomes the controlling factor and heat transfer mainly occurs through a gas film. Hence, the heat transfer coefficient decreases with an increase in the glass particle diameter. This effect has also been observed by Kharchenko and Makhorin [1964], Richardson and Shakiri [1979], Yamazaki and Jimbo [1970] and Kunii and Levenspiel [1969].

The scatter in the heat transfer coefficients are shown in Figure 3.11. The spread is comparatively larger near the minimum fluidization condition. This plot clearly emphasizes the difficulties in obtaining reproducible heat transfer data in fluidized bed studies. This scatter in the heat transfer coefficients which is rarely reported in the literature is mainly due to dynamic conditions of the fluidized bed.
Figure 3.8: Heat transfer coefficient vs. superficial air velocity for a copper sphere diameter of 1.0 cm and various size glass particles
Figure 3.9: Heat transfer coefficient vs. superficial air velocity for a copper sphere diameter of 1.4 cm and various size glass particles
Figure 3.10: Heat transfer coefficient vs. superficial air velocity for a copper sphere of 2.0 cm diameter and various size glass particles.
Figure 3.11: Scatter in heat transfer coefficient as a function of the superficial air velocity for a copper sphere of 1.4 cm diameter and 126-147 μm glass particles.
3.2.1 Role of Particle and Gas Convective Heat Transfer

The average heat transfer coefficient $h$, may be considered as the sum of two terms; namely, particle convective transport $h_{pc}$, and gas convective transport $h_{gc}$ in the absence of the radiation heat transfer.

$$h = h_{gc} + h_{pc}$$ (3.11)

The gas convective heat transfer coefficient can be estimated from the Ranz and Marshal [1952] correlation for an isolated sphere.

$$NU = 2 + 0.6Re^{1/2}Pr^{1/3}$$ (3.12)

The ratio of the fraction of the total average heat transfer coefficient for the sphere due to gas convective transport, $h_{gc}/h$ is shown in Figures 3.12, 3.13 and 3.14 for 1.4 cm copper sphere immersed in the fluidized bed of various sizes of glass particles. A comparison of these three powder sizes shows that the percentage heat transfer carried by the particle convection is larger for the smaller glass particles. The particle heat transport observed to be about 98 % for the 5-44 $\mu$m glass particles, about 94 % for the 126-147 $\mu$m glass particles and about 92 % for the 355-420 $\mu$m glass particles. At lower superficial air velocities, the gas convection heat transport is as high as 37 % for the 355-420 $\mu$m glass particles. In all the cases the particle convective component is higher for the small particle sizes as compare to the large particle system.
Figure 3.12: Contribution of gas convection heat transfer to particle convection heat transfer for a 1.0 cm copper sphere immersed in a fluidized bed of 5-44 μm glass particles.
Figure 3.13: Contribution of gas convection heat transfer to particle convection heat transfer for a 1.0 cm copper sphere immersed in a fluidized bed of 126-147 μm glass particles
Figure 3.14: Contribution of gas convection heat transfer to particle convection heat transfer for a 1.0 cm copper sphere immersed in a fluidized bed of 355-420 μm glass particles.
3.2.2 Heat transfer coefficient as a function of average sphere temperature

As indicated in Chapter 2, the average heat transfer coefficient is calculated from a cooling curve obtained for a copper sphere-air fluidized bed system. A typical cooling curve is shown in Figure 2.12. The average heat transfer coefficient is evaluated for the whole curve where the initial sphere temperature is about 420 to 400 K and the final sphere temperature is about 300 K. Dividing one cooling curve into a number of smaller curves, each representing a sphere temperature drop of 5 K, the heat transfer coefficient for each individual temperature step was calculated. As for example one complete cooling curve from 400 K to 300 can be divided into 20 temperature steps such that during the first step the sphere temperature would drop from 400 K to 395 K etc. For each 5 K temperature drop, the heat transfer coefficient was evaluated and plotted against the average sphere temperature as shown in Figure 3.15 for 1.0 cm copper sphere immersed in a fluidized bed of 126-147 μm glass particles. For this study, a high speed data acquisition system (ADM module) was used with the frequency range of 10 to 50 readings per second. The average sphere temperature is the arithmetic mean between the initial and the final sphere temperatures during each temperature step. The time required for a temperature step of 5 K varies from 0.6 second for the higher average sphere temperatures to 0.9 second for the lower average sphere temperatures. Considering this, the heat transfer coefficient obtained for this temperature step can be considered as almost an instantaneous heat transfer coefficient. High values of the instantaneous heat transfer coefficient suggests that during that period a fresh (cold) packet of
emulsion might have come in the contact with the copper sphere. While the low values of the instantaneous heat transfer coefficient implies that the copper sphere might have been engulfed by a large bubble during that period. Although the heat transfer coefficients are very much scattered, the general trend as shown in Figure 3.15 is that of a decreasing heat transfer coefficient with a decreasing average sphere temperature. The same cooling curve was also divided into the various other temperature steps such as; 10 K, 15 K, 20 K, and 30 K. The heat transfer coefficients for each of the temperature steps are plotted against the average sphere temperature as shown in Figures 3.15 to 3.19.

The spread in the heat transfer coefficient is almost negligible for larger temperature steps as shown in Figures 3.15 to 3.19. However as noted earlier, the heat transfer coefficient decreases with the average sphere temperature for all temperature steps. This decrease is probably a result of the changing thermophysical properties of the packet of emulsion with the average film temperature. However the heat transfer coefficient decreases by 12 to 20 % as the average sphere temperature varies from 400 K to 310 K. Considering the range of the average sphere temperatures, it is not likely that the change in the thermophysical properties of the emulsion packet will contribute to this considerable reduction in the heat transfer coefficient. Hence this reduction in the heat transfer coefficient could be attributed partly to change in the thermophysical properties of the emulsion packet and partly to the dynamic conditions of the fluidized bed during the respective temperature steps. Furthermore it can be inferred from the scatter in the heat transfer coefficients that in order to obtain the average value of heat transfer coefficient, the
Figure 3.15: Heat transfer coefficient vs. average sphere temperature for a temperature step of 5 K and for a 1.0 cm copper sphere immersed in a fluidized bed of 126-147 μm glass particles.
Figure 3.16: Heat transfer coefficient vs. average sphere temperature for a temperature step of 10 K and for a 1.0 cm copper sphere immersed in a fluidized bed of 126-147 μm glass particles.
Figure 3.17: Heat transfer coefficient vs. average sphere temperature for a temperature step of 15 K and for a 1.0 cm copper sphere immersed in a fluidized bed of 126-147 μm glass particles
Figure 3.18: Heat transfer coefficient vs. average sphere temperature for a temperature step of 20 K and for a 1.0 cm copper sphere immersed in a fluidized bed of 126-147 μm glass particles.
Figure 3.19: Heat transfer coefficient vs. average sphere temperature for a temperature step of 30 K and for a 1.0 cm copper sphere immersed in a fluidized bed of 126-147 μm glass particles.
temperature step should be more than 20 K.

The study of the temperature dependence characteristics of the heat transfer coefficient was also extended to the packed beds. Figure 3.20 shows the heat transfer coefficient as a function of the average sphere temperature obtained from a single cooling curve for a system of hot copper sphere immersed in a packed bed of 126-147 \( \mu \text{m} \) glass particles. In a packed bed due to the absence of the particulate phase circulation, the residence time of the emulsion packet surrounding the sphere is essentially infinite. The heat transfer coefficient is very high for the initial time as the emulsion packet is cold, but slowly the temperature of the emulsion packet rises, decreasing the temperature potential and also the heat transfer coefficient. This temperature dependence property of the heat transfer coefficient can be verified by a better suited experimental method of steady state heat transfer from an immersed surface to the bed. This can be achieved by inserting a heater in side the object or by electrically heating the object and monitoring the power required to maintain the constant temperature of the immersed object.

### 3.3 Heat Transfer from Linearly Downward Moving Sphere

Heat transfer coefficients were evaluated for spheres moving linearly downward at various velocities in an air fluidized bed of various size glass particles. Figures 3.21 to 3.29 shows such plots. The copper sphere was moved in a downwardly direction by the gear train mechanism as shown in Figure 2.4. The time required for a sphere to traverse the fluidized bed from the top to the bottom of the fluidized bed varied from 60 second for a sphere moving at 0.4 cm/s velocity to 3.2 second
Figure 3.20: Heat transfer coefficient vs. average sphere temperature for a 1.0 cm copper sphere immersed in a packed bed of 126-147 μm glass particles.
for a sphere moving at 7.5 cm/s. Even though it takes only 3.2 second for a fast moving sphere to traverse the fluidized bed from the top to the bottom, the high heat transfer coefficient associated with the motion of the sphere ensured that the sphere temperature dropped at least by 30° to 40 ° C during this time. As established earlier, this much temperature drop is necessary to get the average heat transfer coefficient.

It is reasonable to expect that the heat transfer between an immersed object and a fluidized bed to closely depend on the flow configuration around the object. Figures 3.21, 3.22 and 3.23 shows plots of the heat transfer coefficient versus superficial air velocity at various average sphere velocity in downward direction for a fluidized bed of 5-44 μm glass particles and various size copper spheres. Near minimum fluidization the heat transfer coefficient for the case of the sphere moving at 7.5 cm/s is observed to increase about 10 to 13 times as compared to that of a stationary sphere. Even at a sphere velocity of 0.4 cm/s, the increase in the heat transfer coefficient is about 25 % for the 1.4 cm sphere and about 280% for the 1.0 cm sphere. As seen in Figures 3.21, 3.22 and 3.23, each plot can be divided in to two regions. In region one where the superficial air velocity is less than 0.5 cm/s the heat transfer coefficient increases continuously with the linear velocity of the sphere. Moreover, the heat transfer coefficient decreases with an increase in the superficial air velocity at higher sphere velocities while it increases with superficial air velocity at lower sphere velocities. Even though there is a considerable amount of scatter in the data, this general trend is fairly well established. It is interesting to note that the heat transfer coefficient remains almost constant with the superficial
Figure 3.21: Heat transfer coefficient for a 1.0 cm copper sphere in linearly downward motion (various speeds) versus superficial air velocity for a fluidized bed of 5-44 \( \mu \text{m} \) glass particles.
Figure 3.22: Heat transfer coefficient of a 1.4 cm copper sphere in linearly downward motion (various speeds) versus superficial air velocity for a fluidized bed of 5-44 μm glass particles.
Figure 3.23: Heat transfer coefficient of a 2.0 cm copper sphere in linearly downward motion (various speeds) versus superficial air velocity for a fluidized bed of 5-44 μm glass particles.
air velocity for 1.0 cm diameter sphere moving at 1.1 and 1.9 cm/s. In the region two, where the superficial velocity is greater than 0.5 cm/s, the increase in the heat transfer coefficient with the sphere velocity is very small and scatter in the data is larger compared to the region one.

In Figures 3.24 to 3.26 and 3.27 to 3.29, the average heat transfer coefficients are plotted against the superficial air velocity for various sphere velocities and fluidized beds of 126-147 and 355-420 μm glass particles respectively. The general trend of increasing heat transfer coefficient with sphere velocity in region one where the superficial air velocity is lower than 5 cm/s for the 126-147 μm glass particles and 20 cm/s for the 355-420 μm glass particles, remains the same as mentioned previously. As shown in Figures 3.24 to 3.26, in region two, where the superficial air velocity is greater than 5 cm/s, the increase in the heat transfer coefficient with sphere velocity is much lower than that for the fluidized bed of 5-44 μm glass particles. Moreover for 355-420 μm glass particles, the heat transfer coefficient remains almost constant for all sphere velocities in region two where the superficial air velocity is greater than 22 cm/s. The increase in the heat transfer coefficient for the 1.0 cm sphere at 7.5 cm/s in the fluidized bed of 126-147 μm is about 10 to 12 times as compared to the stationary sphere near minimum fluidization condition.

As shown in the Figure 3.29, the heat transfer coefficient remains almost constant for 2.0 cm diameter sphere moving at 3 cm/s in the fluidized bed of 355-420 μm glass particles. Furthermore, the scatter in the data is less compare to that of the 5-44 μm glass particles. In this case the heat transfer coefficient increases by 4 to 5 times near the minimum fluidization condition for the sphere moving at 7.5
Figure 3.24: Heat transfer coefficient of a 1.0 cm copper sphere in linearly downward motion (various speeds) versus superficial air velocity for a fluidized bed of 126-147 μm glass particles.
MOVING SPHERE
(Linearily Downward)

SPHERE VELOCITY

- □ STATIONARY
- △ 0.4 cm/s
- ◇ 1.1 cm/s
- ▽ 1.9 cm/s
- □ 3.0 cm/s
- ◆ 4.6 cm/s
- ○ 7.5 cm/s

SPHERE DIAMETER = 1.4 cm
PARTICLE SIZE = 126-147 micron
Um = 1.6 cm/s

Figure 3.25: Heat transfer coefficient of a 1.4 cm copper sphere in linearly downward motion (various speeds) versus superficial air velocity for a fluidized bed of 126-147 µm glass particles
Figure 3.26: Heat transfer coefficient of a 2.0 cm copper sphere in linearly downward motion (various speeds) versus superficial air velocity for a fluidized bed of 126-147 μm glass particles.
Figure 3.27: Heat transfer coefficient of a 1.0 cm copper sphere in linearly downward motion (various speeds) versus superficial air velocity for a fluidized bed of 355-420 μm glass particles
Figure 3.28: Heat transfer coefficient of a 1.4 cm copper sphere in linearly downward motion (various speeds) versus superficial air velocity for a fluidized bed of 355-420 μm glass particles.
Figure 3.29: Heat transfer coefficient of a 2.0 cm copper sphere in linearly downward motion (various speeds) versus superficial air velocity for a fluidized bed of 355-420 μm glass particles
As mentioned above in all the cases the heat transfer coefficient for a sphere moving at 7.5 cm/s is observed to increase by about 4 to 13 times with respect to the stationary sphere. However at higher superficial air velocities, the difference between the heat transfer coefficient for a moving and a stationary sphere is diminished. This might be explained as follows: at lower superficial air velocities, the absence of appreciable circulation of the particulate phase makes the residence time of the emulsion packet very large for a stationary sphere. Thus, the motion of the sphere at the low superficial air velocities decreases the residence time considerably and improving the heat transfer coefficient. At higher superficial air velocities, the circulation of the particles around the sphere is already high and the motion of the sphere does not significantly change the residence time of the emulsion packet, thus, the influence of the forced motion of the sphere on the heat transfer coefficient is reduced.

The effect of the glass particle size on the heat transfer coefficient for a moving sphere is the same as that of the stationary sphere: the heat transfer coefficient increases with a decrease in the glass particle size. However the well established trend of increasing heat transfer coefficient for a stationary sphere with a decrease in the sphere diameter is not very clearly established for a moving sphere.

The effect of the forced linear motion of the heat transfer surface on heat transfer coefficient is not recorded in the literature. However, a limited amount of work has been done on the heat transfer between a moving bed of solid particles and an immersed surface. Heat transfer between a surface and a fluidized or moving
packed bed is determined by the time of contact of the solid particles with the surface and the porosity of the bed. At the same contact time a dense moving bed may be considered as a hypothetical fluidized bed whose porosity is equal to that of a dense bed. Colakyan and Levenspiel [1984] observed that the heat transfer coefficient increases with linear solid velocity up to 10.0 cm/s and after that it levels off for a packed bed of 0.27 and 0.8 mm silica sand moving over a heated cylinder. At the minimum fluidization condition, the sphere moving within a fluidized bed is analogous to a moving bed of particles passing through a heat transfer surface. The existence of the maximum heat transfer coefficient with the solid velocity suggests that the continuous increase in the heat transfer coefficient at minimum fluidization condition observed in the present work may levelled off at higher sphere velocities.

Botterill and Denloye [1978a] used a flowing packed bed to model the heat transfer behavior of a freely fluidized bed of solid particles. The residence time of the bed material adjacent to the transfer surface was regulated by controlling the downward rate of the flow of the bed past a small exposed heat transfer surface. They found that the flowing packed bed to surface heat transfer coefficient increases with increase in solid velocity. Dunsky et al. [1966] carried out an experiment in which the residence time of the solids on a heat transfer surface was controlled by the spining action of an annular packed bed column. They showed that the heat transfer coefficient increases with an increase in the speed of the moving bed and gradually reached the maximum value which was different for each size of particles. In recent studies, the sphere was allowed to move freely which tends to follow the natural motion of the bulk circulation. Prins et al. [1986] measured the heat
transfer coefficient from a gas fluidized bed to a fixed and nearly freely moving single sphere over a wide range of bed particles and heat transfer sphere diameters. The heat transfer coefficient for a fixed sphere was found to be higher; however, the difference was relatively small and decreased for larger bed particles. Rios and Gibert [1984] showed that the heat transfer coefficient increases by 170% for a freely moving sphere when compared to a stationary sphere at \( U = 2.3U_{mf} \). This increase follows the same trend as established in the present research.

3.3.1 Industrial figure of merit

The cost of nitrogen fluidizing gas is a significant operational cost in industrial heat treatment processes. A trade-off now seems possible between nitrogen gas throughput and object motion such that the same or improved heat transfer rate is maintained. To demonstrate this Figure 3.30 plots an industrial figure of merit. This is defined here as the ratio of the heat transfer coefficient to the superficial gas velocity plotted versus superficial air velocity based on the data of Figure 3.26 and normalized to the maximum value of figure of merit for the stationary sphere \( [i.e \ (h_m/U)/(h_{s_{max}})] \). A high figure of merit (i.e., greater than 1) compared to the stationary sphere indicates overall improvement in heat transfer at reduced operating cost of the bed. In fact, not only is the figure of merit improved at reduced superficial air velocity of the bed (reduced cost of inert gas), but the absolute value of the heat transfer coefficient is also increases above the maximum value attained for the stationary sphere when the sphere is moved linearly downward at 7.5 cm/s.
Figure 3.30: Industrial figure of merit, heat transfer to operating cost ratio, normalized to the best stationary sphere operation of the bed.
3.4 Heat Transfer from Oscillating Spheres

The fluidizing action of the mechanical stirrer has been utilized to aid in handling particulate solids in various processes. Oscillating, rotating and impeller type agitators have been used to blend various solid components in cement manufacturing. Rotating stirrers have been used in laboratory cracking catalyst activity evaluation. Vibrating objects immersed in beds of solids have been used as an aid in fluidization at low superficial gas velocities. Relative motion between the immersed object and solid particle is the major factor contributing to the enhanced heat transfer coefficient in the fluidized beds. In this study, hot spheres of various diameters were allowed to oscillate at various frequencies and amplitudes in order to study the effect of oscillating motion on the heat transfer coefficient.

Figures 3.31, 3.32 and 3.33 shows the heat transfer coefficient as a function of superficial air velocity for various frequencies of a 1.0 cm oscillating sphere and peak-to-peak amplitudes of 1.8 cm, 4.0 cm and 6.9 cm respectively. The heat transfer coefficient increases by 7.5 to 14 times for a sphere oscillating when compared to the stationary sphere at 2.85 hz as the peak-to-peak amplitude increases from 1.8 to 6.9 cm. As shown in Figure 3.31, the heat transfer coefficient increases continuously with the superficial air velocity at lower peak-to-peak amplitude of 1.8 cm. Whereas at higher peak-to-peak amplitudes, the heat transfer coefficient increases rapidly at the lower superficial air velocities followed by a gradual increase or leveling off at the higher superficial air velocities. At the peak-to-peak amplitude of 4.0 cm and 6.9 cm and the oscillating frequency of 2.85 hz, the heat transfer coefficient remains almost constant for a superficial air velocity greater than 0.2 cm/s.
Figures 3.34-3.36 and Figures 3.37-3.39 show the plots of heat transfer coefficient versus the superficial air velocity for 1.4 and 2.0 cm sphere oscillating at various frequencies and peak-to-peak amplitudes within a fluidized bed of 5-44 μm glass particles. The general trend remains the same as in the previous case; for the lower peak-to-peak amplitude, the heat transfer coefficient increased continuously with the superficial air velocity, while for higher amplitudes the heat transfer coefficient increases rapidly at lower superficial air velocities and remains almost constant at higher superficial air velocities.

Figures 3.40 to 3.48 shows plots of the heat transfer coefficient for various peak-to-peak amplitudes and sphere diameters versus the superficial air velocity for the fluidized bed of 126-147 μm glass particles. Figures 3.49 to 3.57 shows plots of the heat transfer coefficient for various peak-to-peak amplitudes and sphere diameters versus the superficial air velocity for the fluidized bed of 355-420 μm glass particles.

As shown in these figures, the general trend remains the same as the heat transfer coefficient increases with increasing frequency and peak-to-peak amplitude.

The oscillating motion of the sphere can be converted into an average velocity assuming a sinusoidal motion of the sphere.

\[ V_{avg} = 2 A_s F_s \]  \hspace{1cm} (3.13)

Table 3.2 shows the equivalent average velocity of the sphere oscillating at a particular frequency and peak-to-peak amplitude.

The heat transfer coefficients are plotted as a function of superficial air velocities for different equivalent average sphere velocities in Figure 3.58. The heat transfer coefficient increases gradually as the average sphere velocity increases. However,
Table 3.3: Average velocity of the sphere oscillating at various frequencies and amplitudes

<table>
<thead>
<tr>
<th>Peak-to-peak amplitude, cm</th>
<th>Frequency Hz</th>
<th>Average Velocity cm/s</th>
</tr>
</thead>
<tbody>
<tr>
<td>1.8</td>
<td>1.1</td>
<td>3.96</td>
</tr>
<tr>
<td>4.0</td>
<td>2.0</td>
<td>7.20</td>
</tr>
<tr>
<td>4.0</td>
<td>1.1</td>
<td>8.80</td>
</tr>
<tr>
<td>1.8</td>
<td>2.85</td>
<td>10.26</td>
</tr>
<tr>
<td>6.9</td>
<td>1.1</td>
<td>15.18</td>
</tr>
<tr>
<td>4.0</td>
<td>2.00</td>
<td>16.00</td>
</tr>
<tr>
<td>4.0</td>
<td>2.85</td>
<td>22.8</td>
</tr>
<tr>
<td>6.9</td>
<td>2.00</td>
<td>27.6</td>
</tr>
<tr>
<td>6.9</td>
<td>2.85</td>
<td>39.33</td>
</tr>
</tbody>
</table>
at the average sphere velocities of 8.8 cm/s and 16.0 cm/s the heat transfer coefficient decreased unexpectedly. The increase in heat transfer coefficient was found to be more at the lower superficial air velocities as compared to the higher superficial air velocities. The variation in the heat transfer coefficient with the average velocity for four different superficial air velocities is shown in Figure 3.59. The heat transfer coefficient increases rapidly with increasing average sphere velocity at lower superficial velocity, e.g., at superficial velocity of 0.087 cm/s, the heat transfer coefficient increases by more than three times as the average sphere velocity increases from 3.96 cm/s to 39.33 cm/s. Whereas, at superficial air velocity of 1.95 cm/s, the heat transfer coefficient increases only by 20 % in the same range of the average sphere velocity. Figure 3.60 shows the heat transfer coefficient versus the superficial air velocity for various average velocity for a 1.0 cm sphere oscillating in a fluidized bed of 126-147 μm glass particles. The heat transfer coefficient increases rapidly with the average sphere velocity at lower superficial air velocity and increases gradually at higher superficial air velocity. At superficial air velocities of 7.6 cm/s and 14.9 cm/s the heat transfer coefficient remains almost constant as shown in Figure 3.61.

The heat transfer coefficient does not increase rapidly with the average sphere velocity at lower superficial air velocities in case of the 355-420 μm glass particles. As shown in Figure 3.63 an increase in the heat transfer coefficient with the average sphere velocity is very gradual and the scatter in data is much greater as compared to the other two powders. It is interesting to note that the heat transfer coefficient decreases with an increasing average sphere velocity for the fluidized bed of 355-420 μm glass particles. As shown in Figure 3.63, the heat transfer coefficient is about
7% lower for the average sphere velocity of 27.6 cm/s as compared to the average sphere velocity of 3.96 cm/s at the superficial air velocity of 13.2 cm/s. At higher superficial velocities, the heat transfer coefficient decreases with increasing average sphere velocity. In fact for the superficial air velocities of 16.7, 18.5, 24.4 and 33.2 cm/s the heat transfer coefficient is lowest for the highest average sphere velocity of 39.33 cm/s. This decrease in the heat transfer coefficient with increase in the average sphere velocity is very peculiar and it is found only for the 355-420 μm glass particles.

3.4.1 Comparison with linearly downward moving sphere

The oscillating motion of the sphere can be converted into an equivalent average linear velocity. The average velocities obtained by the equation 3.13 is listed in Table 3.3. Using this average sphere velocity for an oscillating sphere, the heat transfer results obtained for an oscillating sphere can be compared with that of the linearly downward moving sphere. In case of the linearly downward moving sphere, the sphere was allowed to travel in six different velocities; 0.4, 1.1, 1.9, 3.0, 4.6, 7.5 cm/s. Whereas, the average sphere velocity varies from 3.96 to 39.33 cm/s in case of the oscillating sphere. In Figures 3.64, 3.65 and 3.66, the heat transfer coefficient for 1.0 cm sphere moving in a linearly downward motion at 4.6 cm/s and 7.5 cm/s in a fluidized beds of various size glass particles is compared to the heat transfer coefficient for a sphere oscillating at an average velocity of 3.96 cm/s and 7.20 cm/s respectively under the same conditions. At the minimum fluidization condition, the heat transfer coefficient for an oscillating sphere with an average sphere velocity
of 7.2 cm/s is 250 W/M² K and for a sphere moving in linearly downward direction at 7.5 cm/s is 650 W/M² K. The heat transfer coefficient was observed to be about 260% higher for the linearly moving sphere when compared to the oscillating sphere with almost the same average velocity at lower superficial air velocity. At higher superficial air velocities, the difference in the heat transfer coefficients for the oscillating and linearly moving spheres is negligible.

Figure 3.65 shows the heat transfer coefficient as a function of the superficial air velocity and for various average sphere velocities for an oscillating and linearly moving sphere inside the fluidized bed of 126-147 μm glass particles. The heat transfer coefficient at minimum fluidization condition for the case of an oscillating sphere is about 1.5 times lower than that of a linearly downward moving sphere which is lower than that of 5-44 μm glass particle system. However the general trend is the same as that of the 5-44 μm glass particle system. At higher superficial air velocities, the heat transfer coefficient remains almost constant for the range of the average sphere velocities shown in the Figure 3.65.

For 355-420 μm glass particles, the heat transfer coefficient is only about 40% higher for a linearly downward moving sphere at 7.5 cm/s as compared to an oscillating sphere with an average sphere velocity of 7.2 cm/s near minimum fluidization condition. The two curves, for the oscillating sphere and the linearly downward moving sphere cross at a superficial air velocity of 17 cm/s as shown in Figure 3.66. For superficial air velocities, greater than 17 cm/s, the heat transfer coefficient is consistently higher for the case of an oscillating sphere than a linearly downward moving sphere. Thus, the oscillating sphere is the more efficient for 355-
420 μm glass particles system as it gives a higher heat transfer coefficients at higher superficial air velocities. The high heat transfer coefficient obtained for the case of a linearly downward moving sphere as compare to the oscillating sphere at lower superficial air velocities can be physically explained as follow: at low superficial air velocity, the particulate phase circulation is absent, and hence, the oscillating sphere, comes in contact with the same emulsion packet again and again increasing the temperature of the emulsion packet and reducing the overall heat transfer coefficient. However, in case of a linearly downward moving sphere, although there is no particulate circulation at low superficial air velocities, the sphere moving through the bed comes in contact with a fresh (cold) emulsion packet as it moves down the fluidized bed and hence the heat transfer coefficient is very high.

As can be seen from the table 3.3, for certain combinations of peak-to-peak amplitudes and frequencies, the average sphere velocity remains the same, e.g., at peak-to-peak amplitude of 6.9 cm and frequency of 1.0 Hz, the average sphere velocity is 15.18 cm/s while, for a peak-to-peak amplitude of 4.0 cm and frequency of 2.0 Hz, the average sphere velocity is 16.0 cm/s. Hence it is reasonable to expect that the heat transfer coefficients obtained for these two combination should be close. Figures 3.67, 3.68 and 3.68 compares two such cases for each glass powder and each sphere. The heat transfer coefficients remains the same for the two combinations of the amplitudes and frequencies for which the average sphere velocities are almost the same for each sphere size. Furthermore, as in case of a stationary sphere, the heat transfer coefficient increases with decreasing sphere diameter for an oscillating sphere.
It follows in general from the aforementioned results that the heat transfer coefficient increases with increasing frequency and amplitude. The only exception found in this study was for the 355-420 μm glass particle system, in which the heat transfer coefficient at the higher frequency was observed to decrease. The motion of the heat transfer surface relative to the particulate phase and to the gas phase generates a greater disturbance, enhancing the heat transfer. Reed and Fenske [1955] conducted an experiment to study the effect vibration of an hollow element with extended surface on heat transfer coefficient in a fluidized beds of nickel powder, lead powder, steel shot and carbon granules. They showed that at low superficial air velocity of 6.1 cm/s, the heat transfer coefficient increased by 66 % for a plate vibrating at 2000 cycles/min and 0.795 cm stroke when compared to a stationary plate. For a superficial air velocity at and below 30.5 cm/s the vibrating motion of the plate was effective in enhancing the heat transfer coefficient. At higher superficial air velocities of 115 cm/s and 228.6 cm/s used to fluidized steel shot, the vibrating motion of the plate had essentially no effect on the heat transfer coefficient. Moreover they showed that the heat transfer coefficient increased with the vibration frequency up to 1000 cycles/min and a stroke of 0.795 cm for nickel powder fluidized by air at 6.1 cm/s superficial velocity. Beyond this vibration frequency, the heat transfer coefficient was observed to decrease with increasing frequency.

Recently, some experiments have been carried out in which instead of vibrating the heat transfer surface, the fluidized bed itself is mounted on a vibrating table, in order to study its effect on heat transfer coefficient. Lu Wei-Ming et al. [1976] and Malhotra and Majumdar [1987] studied such effect. They found that
the heat transfer coefficient to a stationary sphere increases as the amplitude and frequency of the fluidized bed vibration increased in lower superficial gas velocity region. Wei-Ming Lu et al. found that for values of Reynold number based on the sphere diameter and gas properties greater than 1000, the effect of the fluidized bed vibration on the heat transfer was negligible. Their results also showed that the higher gas flow rate did not increase the heat transfer coefficient of the system if the vibration intensity, $\omega^2/g$ is greater than 3. These findings are similar in nature with the present investigation.
Figure 3.31: Heat transfer coefficient of a 1.0 cm copper sphere in oscillating motion (various frequencies) with the peak-to-peak amplitude of 1.8 cm versus superficial air velocity for a fluidized bed of 5-44 μm glass particles
Figure 3.32: Heat transfer coefficient of a 1.0 cm copper sphere in oscillating motion (various frequencies) with the peak-to-peak amplitude of 4.0 cm versus superficial air velocity for a fluidized bed of 5-44 μm glass particles.
Figure 3.33: Heat transfer coefficient of a 1.0 cm copper sphere in oscillating motion (various frequencies) with the peak-to-peak amplitude of 6.9 cm versus superficial air velocity for a fluidized bed of 5-44 μm glass particles.
Figure 3.34: Heat transfer coefficient of a 1.4 cm copper sphere in oscillating motion (various frequencies) with the peak-to-peak amplitude of 1.8 cm versus superficial air velocity for a fluidized bed of 5-44 µm glass particles.
Figure 3.35: Heat transfer coefficient of a 1.4 cm copper sphere in oscillating motion (various frequencies) with the peak-to-peak amplitude of 4.0 cm versus superficial air velocity for a fluidized bed of 5-44 μm glass particles.
Figure 3.36: Heat transfer coefficient of a 1.4 cm copper sphere in oscillating motion (various frequencies) with the peak-to-peak amplitude of 6.9 cm versus superficial air velocity for a fluidized bed of 5-44 μm glass particles.
Figure 3.37: Heat transfer coefficient of a 2.0 cm copper sphere in oscillating motion (various frequencies) with the peak-to-peak amplitude of 1.8 cm versus superficial air velocity for a fluidized bed of 5-44 μm glass particles.
Figure 3.38: Heat transfer coefficient of a 2.0 cm copper sphere in oscillating motion (various frequencies) with the peak-to-peak amplitude of 4.0 cm versus superficial air velocity for a fluidized bed of 5-44 μm glass particles
Figure 3.39: Heat transfer coefficient of a 2.0 cm copper sphere in oscillating motion (various frequencies) with the peak-to-peak amplitude of 6.9 cm versus superficial air velocity for a fluidized bed of 5-44 μm glass particles
Figure 3.40: Heat transfer coefficient of a 1.0 cm copper sphere in oscillating motion (various frequencies) with the peak-to-peak amplitude of 1.8 cm versus superficial air velocity for a fluidized bed of 126-147 \( \mu \text{m} \) glass particles.
Figure 3.41: Heat transfer coefficient of a 1.0 cm copper sphere in oscillating motion (various frequencies) with the peak-to-peak amplitude of 4.0 cm versus superficial air velocity for a fluidized bed of 126-147 μm glass particles.
Figure 3.42: Heat transfer coefficient of a 1.0 cm copper sphere in oscillating motion (various frequencies) with the peak-to-peak amplitude of 6.9 cm versus superficial air velocity for a fluidized bed of 126-147 μm glass particles.
Figure 3.43: Heat transfer coefficient of a 1.4 cm copper sphere in oscillating motion (various frequencies) with the peak-to-peak amplitude of 1.8 cm versus superficial air velocity for a fluidized bed of 126-147 μm glass particles.
Figure 3.44: Heat transfer coefficient of a 1.4 cm copper sphere in oscillating motion (various frequencies) with the peak-to-peak amplitude of 4.0 cm versus superficial air velocity for a fluidized bed of 126-147 µm glass particles.
Figure 3.45: Heat transfer coefficient of a 1.4 cm copper sphere in oscillating motion (various frequencies) with the peak-to-peak amplitude of 6.9 cm versus superficial air velocity for a fluidized bed of 126-147 μm glass particles.
Figure 3.46: Heat transfer coefficient of a 2.0 cm copper sphere in oscillating motion (various frequencies) with the peak-to-peak amplitude of 1.8 cm versus superficial air velocity for a fluidized bed of 126-147 μm glass particles.
Figure 3.47: Heat transfer coefficient of a 2.0 cm copper sphere in oscillating motion (various frequencies) with the peak-to-peak amplitude of 4.0 cm versus superficial air velocity for a fluidized bed of 126-147 \( \mu \)m glass particles.
Figure 3.48: Heat transfer coefficient of a 2.0 cm copper sphere in oscillating motion (various frequencies) with the peak-to-peak amplitude of 6.9 cm versus superficial air velocity for a fluidized bed of 126-147 μm glass particles.
Figure 3.49: Heat transfer coefficient of a 1.0 cm copper sphere in oscillating motion (various frequencies) with the peak-to-peak amplitude of 1.8 cm versus superficial air velocity for a fluidized bed of 355-420 μm glass particles.
Figure 3.50: Heat transfer coefficient of a 1.0 cm copper sphere in oscillating motion (various frequencies) with the peak-to-peak amplitude of 4.0 cm versus superficial air velocity for a fluidized bed of 355-420 μm glass particles.
Figure 3.51: Heat transfer coefficient of a 1.0 cm copper sphere in oscillating motion (various frequencies) with the peak-to-peak amplitude of 6.9 cm versus superficial air velocity for a fluidized bed of 355-420 μm glass particles.
Figure 3.52: Heat transfer coefficient of a 1.4 cm copper sphere in oscillating motion (various frequencies) with the peak-to-peak amplitude of 1.8 cm versus superficial air velocity for a fluidized bed of 355-420 μm glass particles
Figure 3.53: Heat transfer coefficient of a 1.4 cm copper sphere in oscillating motion (various frequencies) with the peak-to-peak amplitude of 4.0 cm versus superficial air velocity for a fluidized bed of 355-420 μm glass particles.
Figure 3.54: Heat transfer coefficient of a 1.4 cm copper sphere in oscillating motion (various frequencies) with the peak-to-peak amplitude of 6.9 cm versus superficial air velocity for a fluidized bed of 355-420 μm glass particles.
Figure 3.55: Heat transfer coefficient of a 2.0 cm copper sphere in oscillating motion (various frequencies) with the peak-to-peak amplitude of 1.8 cm versus superficial air velocity for a fluidized bed of 355-420 μm glass particles.
Figure 3.56: Heat transfer coefficient of a 2.0 cm copper sphere in oscillating motion (various frequencies) with the peak-to-peak amplitude of 4.0 cm versus superficial air velocity for a fluidized bed of 355-420 μm glass particles
Figure 3.57: Heat transfer coefficient of a 2.0 cm copper sphere in oscillating motion (various frequencies) with the peak-to-peak amplitude of 6.9 cm versus superficial air velocity for a fluidized bed of 355-420 μm glass particles.
Figure 3.58: Heat transfer coefficient of a 1.0 cm oscillating copper sphere moving at various equivalent average sphere velocities versus superficial air velocity for a fluidized bed of 5-44 μm glass particles.
Figure 3.59: Heat transfer coefficient of a 1.0 cm oscillating copper sphere versus average sphere velocity for various superficial air velocities.
Figure 3.60: Heat transfer coefficient of a 1.0 cm oscillating copper sphere moving at various equivalent average sphere velocities versus superficial air velocity for a fluidized bed of 126-147 μm glass particles.
Figure 3.61: Heat transfer coefficient of a 1.0 cm oscillating copper sphere versus average sphere velocity for various superficial air velocities.
Figure 3.62: Heat transfer coefficient of a 1.0 cm oscillating copper sphere moving at various equivalent average sphere velocities versus superficial air velocity for a fluidized bed of 355-420 μm glass particles.
Figure 3.63: Heat transfer coefficient of a 1.0 cm oscillating copper sphere versus average sphere velocity for various superficial air velocities
Figure 3.64: Heat transfer coefficient of a 1.0 cm oscillating and linearly downward moving copper sphere versus superficial air velocities for various average velocities and 5-44 μm glass particles.
Figure 3.65: Heat transfer coefficient of a 1.0 cm oscillating and linearly downward moving copper sphere versus superficial air velocities for various average velocities and 126-147 μm glass particles.
Figure 3.66: Heat transfer coefficient of a 1.0 cm oscillating and linearly downward moving copper sphere versus superficial air velocities for various average velocities and 355-420 μm glass particles.
Figure 3.67: Heat transfer coefficient for various sphere diameters versus superficial air velocity for the same linear average velocity and 5-44 μm glass particles.
Figure 3.68: Heat transfer coefficient for various sphere diameters versus superficial air velocity for the same linear average velocity and 126-147 μm glass particles.
Figure 3.69: Heat transfer coefficient for various sphere diameters versus superficial air velocity for the same linear average velocity and 355-420 μm glass particles.
4 HEAT TRANSFER CORRELATION

Heat transfer correlations were obtained for each of three cases, namely, for the stationary sphere, the linearly downward moving sphere, and the oscillating sphere. SAS and RS/1 statistical packages were used to correlate the experimental data. Instead of relying on the statistical parameters such as, correlation coefficient, residuals, etc., the correlations were directly compared with the experimental data in its original form for each case.

4.1 Heat Transfer Correlation for a Stationary Sphere

The heat transfer coefficient for a stationary sphere system incorporated the following independent variables: glass particle diameter, sphere diameter and superficial air velocity. The empirical correlation was proposed in the following form,

$$ Nu_s = C_{s1} \left( \frac{U}{U_{mf}} \right)^{C_{s2}} \left( \frac{gd^3}{\mu^2} \frac{n_p (\rho_p - \rho_g)}{\rho_g} \right)^{C_{s3}} \left( \frac{D_{sph}}{d_p} \right)^{C_{s4}} $$

In the above formulation Nusselt number is based on the sphere diameter and the properties of air. The constants, $C_{s1}$, $C_{s2}$, $C_{s3}$ and $C_{s4}$ were evaluated from a multiple log-linear regression analysis. The range of this equation was confined to the fluidized region only (i.e., not including packed bed) as shown in Figure 4.1. Most practical application apply to the fluidized bed region and, moreover, because
of the shape of the plot of the heat transfer coefficient versus the superficial air velocity, it is not possible to correlate all the data with the log-linear correlation model assumed for this case. The result of the log-linear multi-variable regression obtained using SAS program for a stationary sphere with ±25 % is given as follow;

\[
Nu_s = 0.176 \left( \frac{U}{U_{mf}} \right)^{0.664} \left( \frac{gd_p \rho_g (\rho_p - \rho_g)}{\mu_g^2} \right)^{0.44} \left( \frac{D_{sph}}{d_p} \right)^{0.798}
\]

(4.2)

The range of the application of this correlation and the agreement between the predicted and the experimental heat transfer coefficients are shown in Figures 4.1, 4.2 and 4.3. For the glass particle size of 5-44 \( \mu \)m, the correlated values of the heat transfer coefficient agrees within ±25 % in the fluidized bed region as shown in Figure 4.1. However for the glass particle size of 126-147 \( \mu \)m (Figure 4.2), where the heat transfer coefficients are underpredicted by as much as 20 %. The heat transfer coefficient levels off above the superficial velocity of 25 cm/s for the glass particle system of 355-420 \( \mu \)m where the change in the slope of the curve in which the proposed correlation is not adequately flexible to adopt. The empirical correlation (equation 4.2) obtained implies that, the heat transfer coefficient \( h \) is proportional to \( D_{sph}^{-0.2} \). Shirai et al. [1966] found that \( h \propto D_{sph}^{-0.06} \). The direct relation of the heat transfer coefficient to the glass particle diameter is \( d_p^{0.52} \). This result is contradictory to the experimental findings of an increasing heat transfer coefficient with a decrease in the glass particle diameter (Figures 3.5, 3.6 and, 3.7). However, considering the dependence of the minimum fluidizing velocity \( U_{mf} \) on the glass particle diameter, expressed in the equation 3.4, the heat transfer coefficient
Figure 4.1: Experimental and predicted heat transfer coefficients as a function of superficial air velocity for the fluidized bed of 5-44 μm glass particles and stationary sphere.
Figure 4.2: Experimental and predicted heat transfer coefficients as a function of superficial air velocity for the fluidized bed of 126-147 μm glass particles and stationary sphere.
Figure 4.3: Experimental and predicted heat transfer coefficients as a function of superficial air velocity for the fluidized bed of 355-420 μm glass particles and stationary sphere.
h is proportional to \( d_p^{-0.79} \). This result can be compared to the following results documented by various researchers.

\[
h \propto d_p^{-0.23} \quad (Dow \text{ and } Jacob[1951])
\]

\[
h \propto d_p^{-0.50} \quad (Yamazaki \text{ and } Jimbo[1970])
\]

Figure 4.4 shows the plot of the heat transfer coefficient versus the left hand side of the equation 4.2 without the coefficient 0.176 (XSCORR) and the experimental values of the heat transfer coefficient. The correlation obtained agrees reasonably well with the experimental data.

### 4.2 Heat Transfer Correlation for Linearly Downward Moving Sphere

The heat transfer results for the linearly downward moving sphere were correlated in the form suggested by the equation 4.1 with an added term to include the linear velocity of the sphere \( V_{sp} \). The first attempt for a moving sphere was to correlate all the experimental data in one single correlation as given below (within about \( \pm 30 \% \)):

**Correlation I**

\[
N_{um} = 0.498 \left( \frac{U}{U_{mf}} \right)^{0.251} \left( \frac{g d_p \rho g (\rho_p - \rho_g)}{\mu g^2} \right)^{0.385} \left( \frac{D_{sp}}{d_p} \right)^{0.793} \times \left( \frac{V_{sp}}{U_{mf}} \right)^{0.132}
\]

The above correlation is compared with the experimental values of the heat transfer coefficient given in Figures 4.5, 4.6 and 4.7. As seen from these Figures,
Figure 4.4: Experimental and predicted heat transfer coefficients as a function of XSCORR.
Figure 4.5: Experimental and predicted heat transfer coefficients (correlation I. equation 4.3) as a function of superficial air velocity for the linearly downward moving sphere in a fluidized bed of 5-44 μm glass particles.
Figure 4.6: Experimental and predicted heat transfer coefficients (correlation I, equation 4.3) as a function of superficial air velocity for the linearly downward moving sphere in a fluidized bed of 126-147 μm glass particles.
Figure 4.7: Experimental and predicted heat transfer coefficients (correlation I. equation 4.3) as a function of superficial air velocity for the linearly downward moving sphere in a fluidized bed of 355-420 μm glass particles.
the correlation does not represent the experimental results entirely correctly. One obvious problem occurs in the region of lower superficial air velocities and higher sphere velocities, where the heat transfer coefficient decreases with increase in the superficial air velocity up to certain value of the superficial air velocity, as in Figure 4.5 above \( V_{sph} = 0.4 \text{ cm/s} \) and than either increases or remains nearly constant, changing the slope of the curve from negative to positive. However, the proposed form of the correlation does not permit such a change of the slope.

In order to obtain a better correlation, the heat transfer results were divided into two regions according to the observed trend of the plot of the heat transfer coefficient versus the superficial air velocity as shown in Figure 4.5. In the region A, the heat transfer coefficient is either decreasing or increasing with increase in the superficial air velocity. In the region B, the heat transfer coefficient remains nearly constant or increases with increase in the superficial air velocity. The transition from the region A to the region B occurs at different superficial air velocity for each glass particle size such as; 0.325 cm/s, 3.744 cm/s and 20.0 cm/s for 24.5, 136.5 and 387.5 \( \mu \text{m} \) mean diameter glass particles respectively. This transition point can be defined by one single parameter for all the glass particle sizes and sphere diameters as;

\[
\text{Parameter1} = \left( \frac{U}{U_{mf}} \frac{d_p}{(1-\epsilon_{mf})^3} \right) \left( \frac{\epsilon_{mf}}{\epsilon_{mf}} \right) = 191 \mu \text{m}
\]
Figure 4.8: Experimental and predicted heat transfer coefficients (correlation II. equation 4.5) as a function of superficial air velocity for the linearly downward moving sphere in a fluidized bed of 5-44 μm glass particles
Figure 4.9: Experimental and predicted heat transfer coefficients (correlation II, equation 4.5) as a function of superficial air velocity for the linearly downward moving sphere in a fluidized bed of 126-147 μm glass particles.
Figure 4.10: Experimental and predicted heat transfer coefficients (correlation II, equation 4.5) as a function of superficial air velocity for the linearly downward moving sphere in a fluidized bed of 355-420 μm glass particles.
When the value of the parameter \( l \) is less than 191 \( \mu m \), the observed trend in the heat transfer coefficient is that of the region A, i.e., either increasing or decreasing with an increase in the superficial air velocity. Whereas, if the value of the parameter \( l \) is greater than 191 \( \mu m \), the trend of the region B is observed. Furthermore, the region A can also be divided into two subdivisions namely A1 and A2, representing the decreasing and increasing trend of the heat transfer coefficient with increasing superficial air velocity respectively. The general form of the correlation is given by (within ± 25 %);

**Correlation II**

\[
N_u \mu m = C_{m1} \left( \frac{U}{U_{mf}} \right)^{C_{m2}} \left( \frac{g d^3 \rho_p (\rho_p - \rho_g)}{\mu_g^2} \right)^{C_{m3}} \left( \frac{D_{sph}}{d_p} \right)^{C_{m4}}
\]

\[
\times \left( \frac{V_{sph}}{U_{mf}} \right)^{C_{m5}}
\]

The coefficients \( C_{m1} \) to \( C_{m4} \) for each region and the corresponding condition for the application of the correlation is listed in Table 4.1. The sphere velocities are defined as,

\[
V_1 = 0.4 \text{ cm/s} \quad V_2 = 1.1 \text{ cm/s} \quad V_3 = 1.9 \text{ cm/s} \quad V_4 = 3.0 \text{ cm/s} \quad V_5 = 4.6 \text{ cm/s} \quad V_6 = 7.5 \text{ cm/s}
\]

As shown in Figures 4.8 to 4.10, the agreement between the predicted and the observed values of the heat transfer coefficient is better than the previous correlation. However for the case of the 355-420 \( \mu m \) glass particles, the correlation does
Table 4.1: Coefficients for the equation 4.5 and corresponding range of application for all glass particle size

<table>
<thead>
<tr>
<th>Region</th>
<th>( C_{m1} )</th>
<th>( C_{m2} )</th>
<th>( C_{m3} )</th>
<th>( C_{m4} )</th>
<th>( C_{m5} )</th>
<th>Range of Application</th>
</tr>
</thead>
<tbody>
<tr>
<td>A1</td>
<td>0.404</td>
<td>-0.0406</td>
<td>0.481</td>
<td>0.719</td>
<td>0.389</td>
<td>Parameter ( l \leq 191 ) ( \mu m ) For 5-44 ( \mu m ) ( D_{sph} = 1.0 cm ), ( V_1 ) to ( V_6 ) ( D_{sph} = 1.4 cm ), ( V_3 ) to ( V_6 ) ( D_{sph} = 2.0 cm ), ( V_2 ) to ( V_6 ) For 126-147 ( \mu m ) ( D_{sph} = 1.0 cm ), ( V_3 ) to ( V_6 ) ( D_{sph} = 1.4 cm ), ( V_4 ) to ( V_6 ) ( D_{sph} = 2.0 cm ), ( V_5 ) to ( V_6 ) For 355-420 ( \mu m ) ( D_{sph} = 1.0 cm ), ( V_3 ) to ( V_6 ) ( D_{sph} = 1.4 cm ), ( V_4 ) to ( V_6 ) ( D_{sph} = 2.0 cm ), ( V_4 ) to ( V_6 )</td>
</tr>
<tr>
<td>A2</td>
<td>0.406</td>
<td>0.424</td>
<td>0.456</td>
<td>0.729</td>
<td>0.216</td>
<td>Parameter ( l \leq 191 ) ( \mu m ) For 5-44 ( \mu m ) ( D_{sph} = 1.4 cm ), ( V_1 ) to ( V_2 ) ( D_{sph} = 2.0 cm ), ( V_1 ) For 126-147 ( \mu m ) ( D_{sph} = 1.0 cm ), ( V_1 ) to ( V_2 ) ( D_{sph} = 1.4 cm ), ( V_1 ) to ( V_3 ) ( D_{sph} = 2.0 cm ), ( V_1 ) to ( V_4 ) For 355-420 ( \mu m ) ( D_{sph} = 1.0 cm ), ( V_1 ) to ( V_2 ) ( D_{sph} = 1.4 cm ), ( V_1 ) to ( V_3 ) ( D_{sph} = 2.0 cm ), ( V_4 ) to ( V_6 )</td>
</tr>
<tr>
<td>B</td>
<td>0.334</td>
<td>0.439</td>
<td>0.384</td>
<td>0.814</td>
<td>0.033</td>
<td>Parameter ( l \geq 191 ) ( \mu m ) for all range of glass particles and copper spheres used</td>
</tr>
</tbody>
</table>
not follow the experimental trend in the heat transfer coefficient at low superficial air velocity and high sphere velocity.

In order to improve the correlation further, each powder size was correlated separately in the following form;

\[
Nu_m = C_{mm1} \left( \frac{U}{U_{mf}} \right)^{C_{mm2}} \left( \frac{D_{sph}}{d_p} \right)^{C_{mm3}} \left( \frac{V_{sph}}{U_{mf}} \right)^{C_{mm4}}
\]  

(4.6)

In this proposed correlation the Archimedes' number is not included as an independent variable as the correlation is sought for each particle size range. Tables 4.2, 4.3 and 4.4 lists the coefficients for each glass particle size and the range of application of the correlation.

Figures 4.11 to 4.19 shows the experimental and the predicted heat transfer coefficients as a function of superficial air velocity for each glass particle size and copper sphere diameter. Furthermore, Figures 4.20, 4.21 and 4.22 show the overall goodness of fit of the correlation.

4.3 Heat Transfer Correlation For Oscillating Sphere

Heat transfer results for an oscillating sphere were correlated in the same general format as given by the equation 4.1. Instead of the linear velocity of the sphere, the average translational sphere velocity given by the product of the frequency and the peak-to-peak amplitude of the oscillation was used.
Figure 4.11: Experimental and predicted heat transfer coefficients (correlation III, equation 4.6) as a function of superficial air velocity for the linearly downward moving 1.0 cm diameter sphere in a fluidized bed of 5-44 μm glass particles.
Figure 4.12: Experimental and predicted heat transfer coefficients (correlation III, equation 4.6) as a function of superficial air velocity for the linearly downward moving 1.4 cm diameter sphere in a fluidized bed of 5-44 \( \mu \text{m} \) glass particles.
Figure 4.13: Experimental and predicted heat transfer coefficients (correlation III, equation 4.6) as a function of superficial air velocity for the linearly downward moving 2.0 cm diameter sphere in a fluidized bed of 5-44 μm glass particles.
Figure 4.14: Experimental and predicted heat transfer coefficients (correlation III, equation 4.6) as a function of superficial air velocity for the linearly downward moving 1.0 cm diameter sphere in a fluidized bed of 126-147 μm glass particles
Table 4.2: Coefficients for equation 4.6 and the corresponding range of application for 5-44 μm glass particle system

<table>
<thead>
<tr>
<th>Region</th>
<th>$C_{mm1}$</th>
<th>$C_{mm2}$</th>
<th>$C_{mm3}$</th>
<th>$C_{mm4}$</th>
<th>Range of Application</th>
</tr>
</thead>
<tbody>
<tr>
<td>A1</td>
<td>0.219</td>
<td>-0.0301</td>
<td>0.774</td>
<td>0.443</td>
<td>Parameter $1 \leq 191 \mu m$ $D_{sph} = 1.0 \text{ cm, } V_1 \text{ to } V_6$ $D_{sph} = 1.4cm \text{ } V_3 \text{ to } V_8$ $D_{sph} = 2.0cm \text{ } V_2 \text{ to } V_8$</td>
</tr>
<tr>
<td>A2</td>
<td>0.045</td>
<td>0.322</td>
<td>0.886</td>
<td>0.696</td>
<td>Parameter $1 \leq 191 \mu m$ $D_{sph} = 1.0 \text{ cm, } V_1 \text{ to } V_6$ $D_{sph} = 2.0cm \text{ } V_1$</td>
</tr>
<tr>
<td>B</td>
<td>0.238</td>
<td>0.474</td>
<td>0.823</td>
<td>0.0667</td>
<td>Parameter $1 \geq 191 \mu m$ of glass particles and copper spheres used</td>
</tr>
</tbody>
</table>
Table 4.3: Coefficients for equation 4.6 and the corresponding range of application for 126-147 μm glass particle system

<table>
<thead>
<tr>
<th>Region</th>
<th>$C_{mm1}$</th>
<th>$C_{mm2}$</th>
<th>$C_{mm3}$</th>
<th>$C_{mm4}$</th>
<th>Range of Application</th>
</tr>
</thead>
</table>
| A1     | 5.398     | -0.099    | 0.721     | 0.274     | Parameter1 $\leq 191$ μm  
$D_{sp} = 1.0$ cm, $V_3$ to $V_8$  
$D_{sp} = 1.4$ cm $V_4$ to $V_6$  
$D_{sp} = 2.0$ cm $V_5$ to $V_8$ |
| A2     | 1.85      | 0.456     | 0.903     | 0.132     | Parameter1 $\leq 191$ μm  
$D_{sp} = 1.0$ cm $V_1$ to $V_2$  
$D_{sp} = 1.4$ cm $V_1$ to $V_3$  
$D_{sp} = 2.0$ cm $V_1$ to $V_4$ |
| B      | 3.01      | 0.289     | 0.839     | 0.02      | Parameter1 $\geq 191$ μm  
of glass particles and copper spheres used |
Table 4.4: Coefficients for equation 4.6 and the corresponding range of application for 355-420 μm glass particle system

<table>
<thead>
<tr>
<th>Region</th>
<th>$C_{mm1}$</th>
<th>$C_{mm2}$</th>
<th>$C_{mm3}$</th>
<th>$C_{mm4}$</th>
<th>Range of Application</th>
</tr>
</thead>
<tbody>
<tr>
<td>A1</td>
<td>16.88</td>
<td>-0.239</td>
<td>0.716</td>
<td>0.166</td>
<td>Parameter1 ≤ 191 μm</td>
</tr>
<tr>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td>$D_{ph}=1.0,cm$, $V_3$ to $V_6$</td>
</tr>
<tr>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td>$D_{ph}=1.4,cm$, $V_4$ to $V_6$</td>
</tr>
<tr>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td>$D_{ph}=2.0,cm$, $V_4$ to $V_6$</td>
</tr>
<tr>
<td>A2</td>
<td>12.83</td>
<td>0.549</td>
<td>0.717</td>
<td>0.778</td>
<td>Parameter1 ≤ 191 μm</td>
</tr>
<tr>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td>$D_{ph}=1.0,cm$, $V_1$ to $V_2$</td>
</tr>
<tr>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td>$D_{ph}=1.4,cm$, $V_1$ to $V_3$</td>
</tr>
<tr>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td>$D_{ph}=2.0,cm$, $V_1$ to $V_3$</td>
</tr>
<tr>
<td>B</td>
<td>9.91</td>
<td>0.175</td>
<td>0.778</td>
<td>0.011</td>
<td>Parameter1 ≥ 191 μm</td>
</tr>
<tr>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td>of glass particles and copper spheres used</td>
</tr>
</tbody>
</table>
\[ Nu_0 = 0.794 \left( \frac{U}{U_{mf}} \right)^{0.228} \left( \frac{gd_p^3 \rho g (\rho_p - \rho_g)}{\mu^2 g} \right)^{0.361} \left( \frac{D_{sph}}{d_p} \right)^{0.685} \times \left( \frac{2F_{sph} A_{sph}}{U_{mf}} \right)^{0.159} \]  

(4.7)

Figures 4.23 to 4.25 shows the goodness of fit for the above correlation. The difference in the predicted and the experimental heat transfer coefficient is about ± 30% for the case of the 5-44 μm glass particle at higher frequency and higher peak-to-peak amplitude as seen in Figure 4.23.

The correlations obtained given here can serve as guidelines for a reasonable prediction of the heat transfer coefficient under certain conditions. The validity of these correlations beyond the range of the present experiment is questionable. The heat transfer characteristics of a fluidized bed depends largely on the hydrodynamics of the bed. Many variables affect the hydrodynamic behavior of the bed such as the distributor plate design, the physical dimensions of the bed, the geometry of the immersed object, and the properties of the fluidized particles.
Figure 4.15: Experimental and predicted heat transfer coefficients (correlation III. equation) as a function of superficial air velocity for the linearly downward moving 1.4 cm diameter sphere in a fluidized bed of 126-147 μm glass particles.
Figure 4.16: Experimental and predicted heat transfer coefficients (correlation III, equation 4.6) as a function of superficial air velocity for the linearly downward moving 2.0 cm diameter sphere in a fluidized bed of 126-147 μm glass particles.
Figure 4.17: Experimental and predicted heat transfer coefficients (correlation III, equation 4.6) as a function of superficial air velocity for the linearly downward moving 1.0 cm diameter sphere in a fluidized bed of 355-420 μm glass particles.
Figure 4.18: Experimental and predicted heat transfer coefficients (correlation III, equation 4.6) as a function of superficial air velocity for the linearly downward moving 1.4 cm diameter sphere in a fluidized bed of 355-420 μm glass particles.
Figure 4.19: Experimental and predicted heat transfer coefficients (correlation III, equation 4.6) as a function of superficial air velocity for the linearly downward moving 2.0 cm diameter sphere in a fluidized bed of 355-420 μm glass particles.
Figure 4.20: Predicted heat transfer coefficients (correlation III. equation 4.6) as a function of the experimental heat transfer coefficient for a fluidized bed of 5-44 μm glass particles
Heat transfer correlation
Linearly downward moving sphere

Figure 4.21: Predicted heat transfer coefficients (correlation III, equation 4.6) as a function of the experimental heat transfer coefficient for a fluidized bed of 126-147 μm glass particles
Figure 4.22: Predicted heat transfer coefficients (correlation III. equation 4.6) as a function of the experimental heat transfer coefficient for a fluidized bed of 355-420 μm glass particles
Figure 4.23: Experimental and predicted heat transfer coefficients as a function of superficial air velocity for an oscillating sphere in a fluidized bed of 5-44 μm glass particles.
Figure 4.24: Experimental and predicted heat transfer coefficients as a function of superficial air velocity for an oscillating sphere in a fluidized bed of 126-147 μm glass particles.
Figure 4.25: Experimental and predicted heat transfer coefficients as a function of superficial air velocity for an oscillating sphere in a fluidized bed of 355-420 μm glass particles.
Figure 4.26: Predicted heat transfer coefficients as a function of the experimental heat transfer coefficient for an oscillating sphere in a fluidized bed.
5 THEORETICAL HEAT TRANSFER MODEL

Many fundamental models have been proposed as described in chapter 1 to analyze the particle convective heat transfer process in a fluidized beds. A simple model containing the main features of the fluidized bed condition is the packet theory [Mickley and Fairbanks, 1955]. However, it predicts excessively high coefficients for very short particle residence times of a packet of emulsion adjacent to the heat transfer surface. To avoid this problem, a new model was postulated [Gelperin and Einstein, 1971]. At the instant \( t=0 \), the heat transfer surface, which is at temperature \( T_w \), is approached by a packet of solid particles with the initial temperature equal to that of the bulk bed. In the zone adjacent to the wall, the voidage is different from that of the packet having a thermal resistance is \( R_w \). However, the emulsion packet next to this zone is assumed to be homogeneous with a thermal resistance \( R_a \). At the boundary of this zone the temperature drops from \( T_w \) to \( \tilde{T} \) as shown in Figure 5.1. The packet of the solid particles are being heated from the boundary (dotted line in Figure 5.1) of this zone, having an instantaneous thermal resistance,

\[
R_a = \left( \frac{\pi \tau}{K_{eff} \rho_m f C_p} \right)^{1/2} \tag{5.1}
\]
Figure 5.1: Schematic diagram to illustrate the mechanism of heat transfer from a surface to a fluidized bed
and an instantaneous heat transfer coefficient, assuming resistances in series,

\[ h_i = \frac{1}{R_w + R_a} \]  \hspace{1cm} (5.2)

Defining, the average heat transfer coefficient as;

\[ h = \int_0^t h_i \, dt \]  \hspace{1cm} (5.3)

The average heat transfer coefficient can be given by,

\[ h = \frac{2}{R_a} \left[ 1 - \frac{R_w}{R_a} \ln(1 + \frac{R_a}{R_w}) \right] \]  \hspace{1cm} (5.4)

The wall resistance \( R_w \), was assumed to be \( d_p/10Kg \) [Geldart, 1986]. In a slightly different situation where, instead of the temperature of the boundary of the zone, \( T \), being constant, the temperature of the heat transfer remains constant. This condition is more typical of heat transfer in a fluidized bed. The average heat transfer coefficient obtained for this condition can be given as [Davidson and Harrison, 1971];

\[ h = \frac{2}{R_a} \left\{ 1 - \frac{\pi R_w}{2 R_a} \left[ 1 - \exp \left( \frac{R_a}{R_w \sqrt{\pi}} \right)^2 \text{erfc} \left( \frac{R_a}{R_w \sqrt{\pi}} \right) \right] \right\} \]  \hspace{1cm} (5.5)

Instead of this strict solution, without large errors, an alternate simplified solution can be give as [Gelperin and Einstein, 1971],

\[ h = \frac{1}{R_w + 0.5R_a} \]  \hspace{1cm} (5.6)

the numerical difference between equations 5.3, 5.4 and 5.5 is negligible for most practical problems.
In order to estimate the average heat transfer coefficient using the above equations, for a fluidized bed system, the properties of the packet and the residence time of the emulsion packet should be known. The effective thermal conductivity of the packet $K_{eff}$ can be given as a sum of two components, one responsible for the heat transfer mechanisms independent of fluid flow and the other representing the effect of the fluid flow [Xavier and Davidson, 1985];

$$\frac{K_{eff}}{K_g} = 1 + \frac{(1 - \epsilon_{mf})(1 - \frac{K_g}{K_p})}{0.63(\frac{K_g}{K_p})^{0.18}} + 0.1\rho_gC_gdpU_{mf}$$  \hspace{1cm} (5.7)

The following values of the properties were used,

- $K_p = 1.05 W/m^0 K$
- $C_{ps} = 1133 J/Kg^0 K$
- $\rho_p = 2500 Kg/m^3$
- $\rho_{mf} = \rho_p(1 - \epsilon_{mf})$
- $\epsilon_{mf} = 0.52$ for 5 - 44 µm glass particles
- $\epsilon_{mf} = 0.44$ for 126 - 147 µm glass particles
- $\epsilon_{mf} = 0.42$ for 355 - 420 µm glass particles

Furthermore, to use equation 5.3 or 5.4, the residence time of the emulsion packet needs to be evaluated. Baskakov et al. [1973] determined the residence time from the temperature fluctuation of a heat transfer surface immersed in a fluidized bed. Another method to estimate the residence time is to measure the solid velocity past the heat transfer surface [Colakyan and Levenspiel, 1984]. In the
present study such parameters were not measured. However, the residence time of the packet could be estimated. Ignoring the effect of particulate phase circulation near the minimum fluidizing condition, the residence time of the emulsion packet on the heated sphere can be given as:

\[ t = \frac{L_h}{V_{sph}} \]  

and the projected length of the heated sphere can be given by,

\[ L_h = \frac{\pi D_{sph}}{2} \]  

In Figures 5.2, 5.3 and 5.4 the heat transfer coefficient obtained by using equation 5.6 and the experimental values of the heat transfer coefficient are plotted against the residence time obtained using equations 5.7 and 5.8 for the various sphere diameters and sphere linear velocities. As seen in these Figures the agreement between the theoretical heat transfer coefficient and the experimental heat transfer coefficient for a linearly downward moving sphere at a minimum fluidization condition is good.

Two parameters play an important role in the transfer of heat from an immersed surface to the fluidized bed, namely the thermal time constant of the particles close to the heat transfer surface and the residence time of the emulsion packet. The thermal time constant of a single particle in the layer adjacent to the heat transfer surface, assuming the heat transfer coefficient to be \( 24K_g/d_p \) and neglecting the internal temperature gradients is given by,

\[ TC = \frac{\rho_p C_p d_p^2}{144 K_g} \]  

Figure 5.2: Experimental and theoretical (equation 5.6) heat transfer coefficient versus the residence time of the emulsion packet for 5-44 μm glass particle fluidized bed near minimum fluidization velocity
Figure 5.3: Experimental and theoretical (equation 5.6) heat transfer coefficient versus the residence time of the emulsion packet for 126-147 μm glass particle fluidized bed near minimum fluidization velocity.
Figure 5.4: Experimental and theoretical (equation 5.6) heat transfer coefficient versus the residence time of the emulsion packet for 355-420μm glass particle fluidized bed near minimum fluidization velocity.
The thermal time constant is an indication of how fast a single particle responds to the temperature of an adjacent heat transfer surface. Three regimes of packet thermal response can be classified;

1. For residence time much lower than the thermal time constant, only first layer of particles are affected and the resistance near the heat transfer wall $R_{w}$ dominates.

2. When the residence time and the thermal time constant are of the same order of magnitude, additional layers of particles are affected and wall and packet resistance are significant.

3. If the residence time is much greater than the thermal time constant, first layer of particles approach the temperature of the heat transfer wall and the packet resistance $R_{a}$ dominates.

Furthermore, the order of magnitude of the thermal penetration distance in to the emulsion packet can be estimated from the scale analysis of a one dimensional transient heat conduction equation. The thermal penetration distance can be given as;
Using a nominal residence time of 1.0 second and the properties of the glass particle size of 355-420 μm, the thermal penetration depth was obtained as 175 μm. Thus, for small residence time the assumption of the emulsion packet being homogeneous fails. This may be the reason for very large difference in theoretical and experimental heat transfer coefficient at residence time lower than 0.5 second, in the case of 5-44 μm and 126-147 μm glass particles as shown in Figure 5.2 and 5.3. As at small residence time only the first layer of the particles are actively involved in the heat transfer process. In such case the single particle heat transfer models described in Chapter 1 should be used.

Figure 5.4 shows that the experimental heat transfer coefficients are consistently higher than the theoretical result. This may be explained as follow. For this size of the glass particle, the bubbling velocity and the minimum fluidization velocity are nearly equal. Thus, even at minimum fluidization condition some particulate circulation is present giving rise to a lower residence time and higher heat transfer coefficient.

5.1 Hydrodynamics of Fluidized Bed

Heat transfer in fluidized beds largely depend on the dynamics of gas-solid motion inside the bed. An attempt was made to study the motion of the particulate and the gas phase in the fluidized bed used for the present study. The main goal of this study was to estimate the particulate velocity close to an immersed sphere. This
velocity then, can be used to establish the residence time of the particulate packet for the packet theory to predict the heat transfer coefficients at various superficial air velocities.

Hydrodynamic models of fluidization use the principles of conservation of mass, momentum and energy. The nonlinear coupled partial differential equations, written in a form suggested by Soo [1962] are modified by Gidaspow [1986] to account for the particle-to-particle friction. The continuity and separate phase momentum equations for transient two-phase flow in cylindrical coordinates are:

Gas Continuity

\[
\frac{\partial \rho g \varepsilon}{\partial t} + \frac{1}{r} \frac{\partial \rho g \varepsilon r U_g}{\partial r} + \frac{\partial \rho g \varepsilon V_g}{\partial y} = 0
\]  \hspace{1cm} (5.12)

Particle Continuity

\[
\frac{\partial \rho_p (1 - \varepsilon)}{\partial t} + \frac{1}{r} \frac{\partial \rho_p (1 - \varepsilon) r U_p}{\partial r} + \frac{\partial \rho_p (1 - \varepsilon) V_p}{\partial y} = 0
\]  \hspace{1cm} (5.13)

Gas Momentum Equation in \( r \) Direction

\[
\frac{\partial \rho g \varepsilon U_g}{\partial t} + \frac{1}{r} \frac{\partial \rho g \varepsilon U_g^2}{\partial r} + \frac{\partial \rho g \varepsilon U_g V_g}{\partial y} = -\varepsilon \frac{\partial P}{\partial r} + \beta_r (U_g - U_p)
\]  \hspace{1cm} (5.14)

Particle Momentum Equations in \( r \) Direction

\[
\frac{\partial \rho_p (1 - \varepsilon) U_p}{\partial t} + \frac{1}{r} \frac{\partial \rho_p (1 - \varepsilon) U_p^2}{\partial r} + \frac{\partial \rho_p (1 - \varepsilon) U_p V_p}{\partial y} = -\frac{\partial P}{\partial r} + \beta_r (U_p - U_g) - \frac{\partial \tau_{rr}}{\partial r}
\]  \hspace{1cm} (5.15)

where

\[
\frac{\partial \tau_{rr}}{\partial r} = G(\varepsilon) \frac{\partial \varepsilon}{\partial r}
\]  \hspace{1cm} (5.16)
Gas Momentum Equation in y Direction

\[
\frac{\partial \rho g \epsilon U_g}{\partial t} + \frac{1}{r} \frac{\partial \rho g \epsilon \rho V_g V_g}{\partial \rho} + \frac{\partial \rho g \epsilon V_g^2}{\partial y} = -\epsilon \frac{\partial P}{\partial y} + \beta_y (V_g - V_p) - \rho g \epsilon g
\]  

(5.17)

Particle Momentum Equation in y Direction

\[
\frac{\partial \left[ \rho_p (1 - \epsilon) V_p \right]}{\partial t} + \frac{1}{r} \frac{\partial \left[ \rho_p (1 - \epsilon) V_p U_p \right]}{\partial \rho} + \frac{\partial \rho_p (1 - \epsilon) V_p^2}{\partial y} = -(1 - \epsilon) \frac{\partial \rho_p}{\partial y} + \beta_y (V_p - V_g) - \frac{\partial \tau_{yy}}{\partial y} - \rho_p (1 - \epsilon) g
\]  

(5.18)

where

\[
\frac{\partial \tau_{yy}}{\partial y} = G(\epsilon) \frac{\partial \epsilon}{\partial y}
\]  

(5.19)

In the equations of motion, \( \beta_r \) and \( \beta_y \) are the friction coefficients between the gas and the solid particles [Kunii and Levenspiel, 1969]. The terms \( \tau_{rr} \) and \( \tau_{yy} \) are the normal components of the solid phase stress tensor. Gidaspow [1986] correlated these terms based on the experimental findings of Rietma and Mutsers [1974] employing a particle-to-particle interaction term \( G(\epsilon) \). The K-FIX program [Gidaspow, 1986] was used to solve the set of nonlinear partial differential equations for \( U_p, U_g, V_p, V_g, \epsilon, \) and \( P \). The program employs the following boundary and initial conditions:

1. At \( y=0 \), prescribed gas mass flux through the distributor plate.

2. At \( y=\)bed height, \( P=\)atmospheric pressure.

3. At \( r=0 \), free slip condition for both the phases.
4. At \( r = \) bed radius, no slip condition for the gas phase and slip condition for the particulate phase.

5. At \( t \leq 0 \), the bed is at minimum fluidization condition. At \( t = 0 \), superficial velocity is set to some value greater than that of the minimum fluidization condition.

The K-FIX code employs a staggered finite difference mesh system. Phase velocities are centered on cell boundaries, whereas all other quantities are centered at the center of the mesh. Mass and momentum fluxes across the cell boundaries are full donor cell differenced. The finite differenced equations can be solved semi-implicitly, by a combination of point relaxation, Newton's and secant methods.

Figure 5.5 shows the particulate velocity profile at 10 cm above the distributor plate, for 136-147 \( \mu m \) glass particle-fluidized bed system at 3 second after the superficial velocity was increased from 1.6 to 4.0 cm/s. For this case, 10 grids in the radial direction and 40 grids in the vertical direction were used. As it can be seen from the Figure 5.5, the particulate velocity is higher at the center of the bed. Even though the solution is not stable particulate circulation pattern is predicted in terms of negative particle velocity at the wall of the bed. For a stable solution, K-FIX program should be run on a fast computer till the particulate velocity achieves a steady state condition.
Figure 5.5: Plot of the radial distance in the bed versus the particulate velocity for 126-147 μm glass particle-fluidized bed system at 3 second
6 CONCLUSIONS AND RECOMMENDATIONS FOR FUTURE WORK

6.1 Conclusions

The heat transfer coefficients for a stationary, a linearly downward moving, and an oscillating copper sphere in an air fluidized bed–glass particle system were evaluated experimentally. The various parameters studied in this work were: the superficial air velocity, copper sphere diameter, glass particle diameter, linear downward sphere velocity, and the frequency and the amplitude of oscillation of the heated sphere. In addition the effect of the temperature of the sphere on the heat transfer coefficient was also studied. Important features of the results are summarized in this chapter.

6.1.1 Stationary sphere

The following heat transfer effects were observed for a small sphere held stationary in an air fluidized bed of glass particles.

1. The average heat transfer coefficient decreased slightly in the region where the superficial air velocity was less than that of the minimum fluidization velocity. However, in the fluidized bed region the average heat transfer coefficient
increased very rapidly followed by a levelling off with further increase in the superficial air velocity. This general trend in the heat transfer coefficient was not observed in the case of 5-44 \( \mu \text{m} \) glass particles where the heat transfer coefficients were low even at a superficial air velocity twice that of the minimum fluidization velocity.

2. The average heat transfer coefficient decreased with decreasing sphere diameter at superficial air velocities greater than that of the minimum fluidization velocity. At superficial velocities lower than that of the minimum fluidization velocity, the effect of the sphere diameter was found to be insignificant.

3. Decreasing the glass particle size was found to increase the heat transfer coefficient for all glass particle sizes and sphere diameters.

4. The contribution to sphere-bed heat transfer rate for a stationary submerged sphere was found as expected to be dominated by particle convective heat transfer in comparison to the gas convective part. At higher superficial air velocity the fraction of particle convective heat transport was observed to be about 92 to 98 % for the case of 5-44 and 126-147 \( \mu \text{m} \) glass particles. Whereas, for lower superficial air velocities and glass particle of 355-420 \( \mu \text{m} \), the particle convective portion was found to be only 63 %.

5. It was found that in order to reliably detect an average heat transfer coefficient, using transient heat transfer theory, that the temperature step corresponding to the initial and final sphere temperature from which the average heat transfer coefficient is evaluated should be more than 20 °C. In general
the heat transfer coefficient was found to increase with increasing average sphere temperature. For the packed bed heat transfer, the heat transfer coefficient was also found to increase with increasing average sphere temperature. This was followed by a rapid drop in the heat transfer coefficient at the lower average sphere temperature.

6. The empirical correlation (equation 4.2) obtained agreed well with the experimental values within ± 25%.

6.1.2 Linearly downward moving sphere

The following general conclusions can be drawn for the heat transfer study of a sphere moving linearly downward in an air fluidized bed.

1. At the lower superficial air velocity the average heat transfer coefficient increased continuously with the linear velocity of the sphere. However, the heat transfer coefficient leveled off at higher superficial air velocities.

2. The heat transfer coefficient increased by 4 to 13 times for a sphere moving at 7.5 cm/s near the minimum fluidizing condition as compared to the stationary sphere, depending on the glass particle size and the sphere diameter. However at higher superficial air velocities the difference between the heat transfer coefficient for a moving and a stationary sphere diminished.

3. The effect of the glass particle size on the heat transfer coefficient for a moving sphere was observed to be the same as in the case of a stationary sphere: increasing heat transfer coefficient with a decrease in the glass particle size.
4. A transitional value of a parameter (equation 4.4) was found to be equal to 191 μm below which all glass particle-sphere systems exhibited a similar trend in heat transfer coefficient; increasing or decreasing (Figure 5.1) trend with the superficial air velocity and above which the heat transfer coefficients were observed to be either leveling off or increased gradually.

5. The heat transfer correlations (equation 4.6) obtained for each glass particle size were found to give good agreement compared with the experimental data within ±20%.

6. A modified packet theory that included the contact resistance at the wall and the resistance in the emulsion packet was used to predict the theoretical heat transfer coefficient for the case of a sphere moving linearly in an air fluidized bed at a minimum fluidization condition. The predicted heat transfer coefficients compared favorably with the experimental values.

6.1.3 Oscillating sphere

The following important heat transfer results were observed for an oscillating sphere in an air fluidized bed.

1. The heat transfer coefficient increased 7.5 to 14 times for an oscillating sphere as compared to a stationary sphere at the same superficial air velocity. At lower peak-to-peak amplitude (1.8 cm) the heat transfer coefficient increased with an increase in the superficial air velocity. However, at higher peak-to-peak amplitude the heat transfer coefficient increased rapidly at lower su-
peripheral air velocity followed by a gradual increase or leveling off at higher superficial air velocities.

2. For the case of 5-44 μm glass particles and a superficial air velocity of 0.087 cm/s, the average heat transfer coefficient increased more than three times as the average translational sphere velocity increased from 3.96 cm/s to 39.33 cm/s. Whereas, at an superficial air velocity of 1.95 cm/s, the heat transfer coefficient increased only by 20% in the same range of the equivalent sphere velocity.

3. A different trend was observed for the case of 355-420 μm glass particles in which, the heat transfer coefficient decreased with an increase in equivalent sphere velocity.

4. The heat transfer coefficient for 5-44 μm glass particle system was observed to be about 260% higher for the linearly downward moving sphere when compared to the oscillating sphere with the same average translational sphere velocity at low superficial air velocity. However for the case of the 355-420 μm glass particle system, the heat transfer coefficient was found to be only 40% higher for a linearly moving sphere. At higher superficial air velocities the difference in the heat transfer coefficients for the oscillating and linearly moving sphere was found to be insignificant for all of the glass particles systems.
6.2 Recommendations for Future Work

Although the present study addresses many issues concerning heat transfer from an immersed sphere to an air fluidized bed, many more issues remain to be resolved. The following areas are recommended for further investigations:

1. Only three size ranges of the glass particles were available for the present (5-44, 126-147, and 355-420 μm). A more extensive and complete study would required many more glass particle size ranges. At least two to three glass powders should be used from each powder group (group A to group D) for better understanding of the effects of the particle size on the heat transfer coefficient.

2. The present study was conducted using only three sizes of relatively small diameter copper spheres (1.0, 1.4, and 2.0 cm). An industrially more useful study should include a complete heat transfer study involving the effects of size of the object (larger than used in the present study), orientation of the object, and shape of the object. Heat transfer in fluidized beds depends mainly on the hydrodynamics of the bed close to the object (e.g., the defluidized zone observed [Ginoux et al., 1974] is influenced by the size and the shape of the object).

3. This study indicated that the heat transfer coefficient could be a function of the average sphere temperature. In order to verify this result a different approach to this heat transfer study should be conducted. For example temperature fluctuation of an object with an internal heater can be used to study
the heat transfer coefficient as a function of the object temperature, the frequency of the bubbles passing by the immersed object, and the instantaneous heat transfer coefficient.

4. The residence time of the emulsion packet was estimated from the sphere diameter and sphere linear velocity for the minimum fluidization condition. In order to estimate the residence time for other values of the superficial air velocities, the computer program K-FIX (developed by Los Alamos and modified by IIT, Chicago) can be used to study the particulate flow around an immersed object. The later version of this software calculates the particulate slip velocity near the heat transfer surface. This slip velocity can be used to estimate the residence time of the particulate phase on the heat transfer surface for the various superficial air velocities.

5. A rigorous heat transfer study can be conducted using the K-FIX (with energy equation) program which can take the variable conductivity (due to variable voidage along the direction of heat transfer) of the particulate phase near the heat transfer surface into account and, thus, can calculate local heat transfer coefficient and an average heat transfer coefficient.

6. The sphere can be moved in upward direction inside the fluidized bed. In this case because of the absence of the defluidized region at the top portion of the sphere, it is expected that the heat transfer coefficient may be different (possibly being less) for this case compared to a sphere moving in downward direction.
7. The residence time of the emulsion packet can be measure experimentally for various superficial air velocities by recording the temperature fluctuation of a heated probe of a thin foil. The experimental value of the residence time can then be used to predict the heat transfer coefficient from the modified packet theory for the complete range of superficial air velocities.

8. Fluidized beds used in the industry are larger in size and have different behavior patterns in three major areas. (1) Properties of large gas distributors: the distributor plate used in the small scale fluidized beds is not practical due to its large pressure drop across the plate. Large fluidized beds employ multi-jet gas distributors of different design. (2) Bubble movement: bubble movement, bubble coalescence and bubble eruption are different in large fluidized beds. (3) Solids movement: multiple circulation patterns may be present along the radial direction of a large fluidized bed unlike in a small fluidized bed. These three factors are mainly responsible for different the heat transfer characteristics of a large fluidized bed. Thus, predictions of heat transfer behavior from testing of a small fluidized bed is often difficult and sometimes impossible. In order to generate useful information for application in industry, heat transfer study should be conducted on a large scale fluidized beds.
I wish to express my thanks to Professor Gerald M. Colver, under whose supervision this research was conducted. I am deeply appreciative of his confidence in my ability, respect for my opinion, and encouragement throughout the course of this research work. I am also grateful for the benefits I have received from his practical knowledge.

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8 BIBLIOGRAPHY


The uncertainty in experimental measurements occurs mainly because of three types of errors: illegitimate, systematic, and random. Illegitimate errors are caused by the error in reading instruments and performing calculation which can be reduced by using care and repetition of experiments and calculation. Systematic error occurs owing to improper calibration of instruments, which can be reduced by careful calibration of instruments. However, the random errors originate from a variety of causes such as fluctuating experimental conditions, or disturbances. Random errors cannot be usually be avoided since these errors are inherently present in any measuring system. The total uncertainty in any experiment can be estimated through the use of propagation of error analysis.

For a function Z,

\[ Z = x_1^a x_2^b x_3^c \ldots x_n^m \]  \hspace{1cm} (9.1)

where, the exponents a, b, c...m may be positive or negative. A simplified form of the propagation of error equation [Beckwith et al. 1982] for the above form can be given as;

\[ \left( \frac{UZ}{Z} \right)^2 = a^2 \left( \frac{UX_1}{X_1} \right)^2 + b^2 \left( \frac{UX_2}{X_2} \right)^2 + \ldots + m^2 \left( \frac{UX_m}{X_m} \right)^2 \]  \hspace{1cm} (9.2)

Applying the above propagation of error equation to the heat transfer correla-
tion (equation 4.5) which can be given in the following form,

\[
Num = C_{m1} \left( \frac{U}{U_{mf}} \right)^{C_{m2}} \left( \frac{gd_p^2 \rho_p (\rho_p - \rho_g)}{\mu_g^2} \right)^{C_{m3}} \left( \frac{D_{sph}}{d_p} \right)^{C_{m4}} \left( \frac{V_{sph}}{U_{mf}} \right)^{C_{m5}}
\]  

(9.3)

From above equations A.2 and A.3 the uncertainty in the heat transfer coefficient can be given as;

\[
\left( \frac{U_h}{h} \right)^2 = \left( \frac{UD_{sph}}{D_{sph}} \right)^2 + C_{m2}^2 \left( \frac{UU}{U} \right)^2 + C_{m2}^2 \left( \frac{UU_{mf}}{U_{mf}} \right)^2 + (3C_{m3})^2 \left( \frac{Ud_p}{d_p} \right)^2 + C_{m4}^2 \left( \frac{UD_{sph}}{D_{sph}} \right)^2 + C_{m4}^2 \left( \frac{Ud_p}{d_p} \right)^2 + C_{m5}^2 \left( \frac{UV_{sph}}{V_{sph}} \right)^2 + C_{m5}^2 \left( \frac{UU_{mf}}{U_{mf}} \right)^2
\]  

(9.4)

Using the following values of the above parameters;

\[
D_{sph} = 1.0 \text{ cm} \quad UD_{sph} = \pm 0.1 \text{ cm} \\
U = 10.0 \text{ cm/s} \quad UU = \pm 1.0 \text{ cm/s} \\
U_{mf} = 1.6 \text{ cm/s} \quad UU = \pm 0.20 \text{ cm/s} \\
d_p = 136.5 \mu m \quad Ud_p = \pm 10.0 \mu m \\
V_{sph} = 4.6 \text{ cm/s} \quad UV_{sph} = \pm 0.3 \text{ cm/s}
\]

the uncertainty in the heat transfer coefficient can be evaluated;

\[
\left( \frac{U_h}{h} \right)^2 = 0.0001 + 0.001797 + 0.0028 + 0.010044
\]
hence, the uncertainty in the heat transfer coefficient is about ±13.6 %. The uncertainty is seem to be mainly due to the uncertainty in measuring the mean diameter of the glass particle, \( d_p \).
The volumetric flow rate of a variable area flowmeters like rotameters used in this experiment can be given as follows.

\[ Q_r = A_w C_d \left( \frac{2gV_f (\rho_f - \rho_g)}{A_f \rho_g} \right)^{\frac{1}{2}} \]  

(10.1)

The float density \( \rho_f \) is much larger than the gas density \( \rho_g \) (usually, \( \frac{\rho_f}{\rho_g} \sim 10^{-3} \)). So the above rotameter equation can be approximated as,

\[ Q_r = A_w C_d \left( \frac{2gV_f \rho_f}{A_f \rho_g} \right)^{\frac{1}{2}} \]  

(10.2)

The relationship between the rotameter flowrate at calibrating condition and at the standard conditions at the same rotameter reading is given by the following expression:

\[ \frac{Q_{rc}}{Q_{rs}} = \left[ \frac{\rho_{std}}{\rho_{rc}} \right]^{\frac{1}{2}} \]  

(10.3)

In the calibration setup, the outlet of a rotameter was connected to the inlet of a wet test meter. Hence from the continuity equation

\[ Q_{rc} = \frac{\rho_{wm}}{\rho_{rc}} Q_{wm} \]  

(10.4)
During the calibration condition, the temperature of the air at rotameter and at wet test meter remained the same. Using the ideal gas law and equations A.3 and A.4,

\[ Q_{rs} = \left( \frac{P_s}{T_s} \right) \left( \frac{T_{rc}}{P_{rc}} \right) Q_{mw} \]  

(10.5)

For each position of the steel float of the rotameter, the above equation was used to obtained a calibration curve for air flow rate at the standard condition. The following wet test meters and airmeter were used to calibrate all the rotameters:

1. Precision Scientific, Chicago, ISU-313167, 0.1 \( f^3/\text{rev} \), maximum capacity = 24 \( f^3/\text{hr} \), minimum capacity = 2 \( f^3/\text{hr} \).

2. Precision Wet test meter, GCA Corporation, Chicago, ISU-335369, 1.0 \( f^3/\text{rev} \), maximum capacity = 110 \( f^3/\text{hr} \), minimum capacity = 10 \( f^3/\text{hr} \).

3. Airmeter, Dresser Measurement, model-1.5M.

Table A.1 lists all rotameters, the range of their applicability, and the calibration equation. Under the operating condition, when air is passed through a rotameter first and then through the fluidized bed, the flow rate at the operating condition can be given by,

\[ Q_{ro} = Q_{rs} \left( \frac{\rho_{std}}{\rho_o} \right) \]  

(10.6)

From the ideal gas law,

\[ Q_{ro} = Q_{rs} \left( \frac{P_s T_o}{T_s P_o} \right)^{\frac{1}{2}} \]  

(10.7)

Moreover, from the continuity equation, the air flow rate at the fluidized bed can be expressed as,

\[ Q_b = \left( \frac{\rho_o}{\rho_b} \right) Q_{ro} \]  

(10.8)
This equation can be rewritten using ideal gas relation, and considering constant temperature for rotameter section and fluidized bed section,

\[ Q_b = \left( \frac{P_0}{P_b} \right) Q_{ro} \]  \hspace{1cm} (10.9)
Table 10.1: List of rotameters used in the experiment

<table>
<thead>
<tr>
<th>Rotameter type</th>
<th>Range in cc/s</th>
<th>Calibration equation cc/s</th>
</tr>
</thead>
<tbody>
<tr>
<td>Lab Crest Div. F and P Co., Meter Tube Cat. no. 450-700</td>
<td>10 to 230</td>
<td>$Q_s = 1.5929R - 13.9910$</td>
</tr>
<tr>
<td>Devco Tube, Cat. no. 250-8</td>
<td>33 to 694</td>
<td>$Q_s = 27.9390R - 6.9009$</td>
</tr>
<tr>
<td>Brooks Instruments, Division of Emerson Co., Model no. 1114-08H2G1A</td>
<td>137 to 998</td>
<td>$Q_s = 9.6269R + 23.4293$</td>
</tr>
<tr>
<td>Brooks Instruments, Division of Emerson Co., Model no. 1110-01A1B1A</td>
<td>1.35 to 13.5</td>
<td>$Q_s = 8.10R^{'60}$</td>
</tr>
<tr>
<td>Brooks Instrument Division of Emerson Electric Co., Tube Size R7M25-1</td>
<td>80 to 713</td>
<td>$Q_s = 7.1395R - 4.0646$</td>
</tr>
<tr>
<td>Schtte and Koerting Co.</td>
<td>750 to 2794</td>
<td>$Q_s = 9.7284R + 440.9542$</td>
</tr>
</tbody>
</table>
The heat loss through the thermocouple wires may introduce some error in calculation of overall heat transfer coefficient for the fluidized bed system. Heat conduction through the thermocouple wire can be treated as conduction through a thin long fin. In this case the governing equation becomes:

$$\frac{d^2T}{dx^2} - \frac{h_{eq}C}{(KA)_{eq}}(T_{tb} - T_b) = 0$$ \hspace{1cm} (11.1)

The conduction heat loss through a very long fin can be given as follows:

$$q_{cl} = \sqrt{\frac{h_{eq}C(KA)_{eq}}{2\pi r_1}}(T_{tb} - T_b)$$ \hspace{1cm} (11.2)

where, $h_{eq}$ and $C$ are equivalent heat transfer coefficient for the convection losses along the fin surface and perimeter of the outer surface. The thermocouple wire along with the insulation and the equivalent radii are shown in Figures B.1a and B.1b. The equivalent radii and the thermal conductivity of the thermocouple wires are defined by Eckert and Goldstein [1976] as follows:

$$r_1 = \sqrt{2}r_w, \hspace{1cm} (11.3)$$
$$r_2 = \frac{L_1 + L_2}{4} \hspace{1cm} (11.4)$$
$$(KA)_{eq} = (K_{cu} + K_{cn})\pi r_w^2 \hspace{1cm} (11.5)$$
Figure 11.1: Thermocouple wire with the insulation (a) Thermocouple wires with insulation (b) equivalent radii
The thermocouple wire was passed through a hollow steel tube (O.D. = 2.0828 mm, I.D. = 1.5748 mm). This provided a relatively thick air gap between the outer boundary of the teflon insulation and the inside surface of the steel tube. This arrangement is shown in Figure B.2. From the thermal resistance circuit, the equivalent heat transfer coefficient for the transverse conduction from the wire to the outer surface of the steel tube can be expressed as:

\[
heq = \frac{1}{r_4 \left( \frac{\ln\left(\frac{r_1}{r_2}\right)}{K_{steel}} + \frac{\ln\left(\frac{r_3}{r_2}\right)}{K_{air}} + \frac{\ln\left(\frac{r_4}{r_2}\right)}{K_{ins}} \right) + \frac{1}{h_1}} \quad (11.6)
\]

As a numerical example for the 30 gauge copper-constantan thermocouple wire, \( r_1 = 0.1796 \text{mm}, \quad r_2 = 0.3229 \text{mm}, \quad r_3 = 0.7874 \text{mm}, \quad r_4 = 1.0414 \text{mm} \)

\( K_{steel} = 50 \text{ W/m}^\circ\text{K}, \quad K_{ins} = 0.35 \text{ W/m}^\circ\text{K}, \quad K_{cu} = 386.0 \text{ W/m}^\circ\text{K}, \quad K_{cn} = 22.7 \text{ W/m}^\circ\text{K}, \quad h_1 = 100.0 \text{ W/m}^2\circ\text{K} \)

\( C = 2\pi r_4 = 0.006543 \text{ m} \)

\( C heq = 0.1457 \frac{w}{m^\circ K} \)

\( (KA)_{eq} = 2.07091 \times 10^{-5} \frac{w}{m^\circ K} \)

For \( T_{lb} - T_{fb} = 100 \circ K \),

\( q_{cl} = 0.1736 W \)

Considering a typical value of heat transfer coefficient as 300 W/m^2\circ K for 1.0 cm copper sphere immersed in a fluidized bed with the temperature difference between the bed and the sphere of 100 \circ K,

\[
q_{conv} = h_{sph}(T_{sph} - T_{fb}) = 9.425 W. \quad (11.7)
\]
Figure 11.2: Thermocouple wires with equivalent radii inside a hollow steel tube.
Hence, the thermocouple conduction heat loss is about 1.8% of the total heat transfer from the sphere. This loss was considered to be negligible.
APPENDIX D: COMPUTER PROGRAMS

C*************************************************************************************************
C THIS DATA ACQUISITION PROGRAM CONNECTS IEEE AND HP DIGITAL VOLTMETER
C
DIMENSION DELTA(600),TEMPK(10)
CHARACTER*50 CHVOLT,COMM(5)
INTEGER*2 ISTAT(2)
REAL VOLTS(600),TIME(600)
CHARACTER*25 FNAME
OPEN(UNIT=2,NAME='FOROO2.DAT',STATUS='OLD')

CINITIALIZATION ROUTINE
CISTAT=STATUS CODE
CISYSC=1 > PC380 IS THE CONTROLLER
ClADR TAKEN AS DEFAULT
ClADR1 TAKEN AS DEFAULT
C
CIEFN=2 > STD EVENT FLAG
CMODE=1 > SYNCHRONOUS I/O (WAIT UNTILL THIS SUBROUTINE IS COMPLETED)
C
CALL IBINIT(ISTAT,1,,2,1)
C
C
CALL WAITFR(2)
PRINT *, 'ISTAT=', ISTAT
C------------------------------------------------------------------
C CALL SUBROUTINE TO CLEAR THE INTERFACE
C------------------------------------------------------------------
C CALL IBIIFC(ISTAT,2,1)
C------------------------------------------------------------------
C WAIT FOR EVENT FLAG(=2)
C------------------------------------------------------------------
C CALL WAITFR(2)
PRINT *, 'ISTAT INTERFACE CLEAR=', ISTAT
C------------------------------------------------------------------
C CALL SUBROUTINE TO ENABLE REMOTE
C------------------------------------------------------------------
C CALL IBREN(ISTAT,2,1)
C------------------------------------------------------------------
C WAIT FOR EVENT FLAG(=2)
CALL WAITFR(2)
PRINT *, 'ISTAT ENABLE REMOTE=', ISTAT
C------------------------------------------------------------------
C CALL SUBROUTINE TO CLEAR THE DEVICE
C------------------------------------------------------------------
C CALL IBDCL(ISTAT,2,1+2)
C------------------------------------------------------------------
C WAIT FOR EVENT FLAG(=2)
CALL WAITFR(2)
PRINT *, 'ISTAT CLEAR DEVICE=', ISTAT
C------------------------------------------------------------------
C SET UP THE HP-DIGITAL VOLTMETER
C------------------------------------------------------------------
READ(2,42,END=31)ND
READ(2,21,END=31)ENDTIM
READ(2,42,END=31)MDIR
READ(2,21,END=31)VSPH
READ(2,21,END=31)FBH
21 FORMAT(20X,G10.4)
279

```
42 FORMAT(20X,I4)
31 WRITE(6,22) ND,ENDTIM,MDIR,VSPH,FBH
22 FORMAT( ' ND,ENDTIM,MDIR,VSPH,FBH' 
* ,I4,1X,F10.4,1X,I4,2(1X,F10.4))
  TYPE 311
311 FORMAT( ' ENTER THE BED TEMPERATURE IN C>',$)
       READ(5,*) TFC
  TYPE 32
32 FORMAT( ' ENTER THE PRESSURE DROP ACROSS THE BED IN CMS>',$)
       READ(5,*) PD
  TYPE 33
33 FORMAT( ' ENTER THE PRESSURE OUTSIDE THE RATOMETER>',$)
       READ(5,*) RPOUT
  TYPE 34
34 FORMAT( ' ENTER RUN NUMBER>',$)
       READ(5,*) RUNNO
  TYPE 35
35 FORMAT( ' ENTER RATOMETER NO.(1 TO 6 )>',$)
       READ(5,*) ROTMNO
  TYPE 36
36 FORMAT( ' DO YOU WANT TO START 
* DATA ACQUISITION?(1-YES,0-NO)>',$)
       READ(5,*) ISTART
IF(ISTART.EQ.0) GO TO 37
COMM(1)='F1 R2 T2 Z0 FLO'
ICOUNT=15

CALL SUBROUTINE TO SEND DATA FROM USER BUFFER TO 
SPECIFIC LISTENER

CALL IBSEND(ISTAT,COMM(1),ICOUNT,2,1+2+8,,18)
CALL WATFR(2)
T1=SECNDS(0.0)

CODE TO BE TIMED
```
PRINT *, ' ISTAT SEND= ', ISTAT
DO 100 I = 1, ND

CALL SUBROUTINE IBGET TO EXECUTE TRIGGER

CALL IBGET(ISTAT, 2, 1+2, ., 18)
CALL WAITFR(2)

CALL SUBROUTINE IBRECV TO RECEIVE DATA

CALL IBRECV(ISTAT, CHVOLT, 14, 5, 1+2, ., 18)
CALL WAITFR(5)

DELTA(I) = SECND(T1)

READ(CHVOLT, 200) VOLTS(I)
TIME0 = DELTA(1)
TIME(I) = DELTA(I) - TIME0
WRITE(6, *) I, TIME(I), VOLTS(I)
IF(TIME(I) .GE. ENDTIM) GO TO 110

100 CONTINUE
200 FORMAT(E12.9)
110 ND = I
TIME0 = DELTA(1)

OPEN(UNIT = 1, TYPE = 'NEW', NAME = FNAME, ERR = 312)
WRITE(1, 336) FBH, PD, RPOUT, RUNNO, ROTMNO, TFC, MDIR, VSPH
336 FORMAT(F10.6/F10.6/F10.6/F10.6/F10.6/F10.6/I4/F10.6)
DO 400 I = 1, ND - 1
TIME0 = DELTA(1)
TIME(I) = DELTA(I) - TIMEO
WRITE(1, 300) TIME(I), VOLTS(I)
300 FORMAT(F10.6, 2X, F12.9)
400 CONTINUE
312 TYPE 314
314 FORMAT(' ERROR WHILE WRITING IN THE FILE')
I = 1
306 CALL MVTEMP(VOLTS(I), TEMPK(I))
IF (I .EQ. ND) GO TO 307
I = ND
GO TO 306
307 ND1 = ND - 1
WRITE(6, 402) N, TEMPK(1), TEMPK(ND).
402 FORMAT(' N, TINITIAL, TEND IN K', I3, 3X, F10.4, 3X, F10.4)
37 STOP
END

C
C
SUBROUTINE MVTEMP(E, TEMPK)
X1 = 0.1008609
X2 = 25727.94369
X3 = -767345.8295
X4 = 78025595.81
X5 = -9247486589.0
X6 = 697688000000.
X7 = -2.66192E+13
X8 = 3.94078E+14
TEMP1 = X1 + X2*E + X3*(E**2)
TEMP2 = X4*(E**3) + X5*(E**4)
TEMP3 = X6*(E**5) + X7*(E**6) + X8*(E**7)
TEMP = TEMP1 + TEMP2 + TEMP3
TEMPK = TEMP + 273.2
RETURN
END
C*------------------------------------------------------------*
C  THIS PROGRAM CALCULATES THE HEAT TRANSFER COEFFICIENT     *
C    FROM THE MILLIVOLTS READING OBTAINED BY THE DATA         *
C    ACQUISITION SYSTEM.                                     *
C                                                               *
C*--------------------------------------------------------------*
C
C VARIABLES
C
<table>
<thead>
<tr>
<th>Name</th>
<th>Type</th>
<th>Description</th>
</tr>
</thead>
<tbody>
<tr>
<td>AINPT</td>
<td>R*4</td>
<td>INTERCEPT OF REGRESSION LINE</td>
</tr>
<tr>
<td>AKAIR</td>
<td>R*4</td>
<td>CONDUCTIVITY OF AIR (W/M K)</td>
</tr>
<tr>
<td>AIR</td>
<td>R*4</td>
<td>AIR AT NORMAL TEMPERATURE</td>
</tr>
<tr>
<td>ANUEXP</td>
<td>R*4</td>
<td>EXPERIMENTAL NUSSELT NO. BASED ON SPHERE DIAMETER AND</td>
</tr>
<tr>
<td></td>
<td></td>
<td>GAS PROPERTIES</td>
</tr>
<tr>
<td>AREA</td>
<td>R*4</td>
<td>SURFACE AREA OF THE SPHERE</td>
</tr>
<tr>
<td>ATMP</td>
<td>R*4</td>
<td>ATMOSPHERIC PRESSURE</td>
</tr>
<tr>
<td>AVGHR</td>
<td>R*4</td>
<td>AVERAGE RADIATION HEAT TRANSFER COEFFICIENT</td>
</tr>
<tr>
<td>BOLTZ</td>
<td>R*4</td>
<td>BOLTZMAN'S CONSTANT</td>
</tr>
<tr>
<td>BSH</td>
<td>R*4</td>
<td>PACKED BED HEIGHT</td>
</tr>
<tr>
<td>CA</td>
<td>R*4</td>
<td>CONSTANT FOR THE EQUATION FOR THE SPECIFIC HEAT AS A</td>
</tr>
<tr>
<td></td>
<td></td>
<td>FUNCTION OF TEMPERATURE</td>
</tr>
<tr>
<td>CB</td>
<td>R*4</td>
<td>CONSTANT FOR THE EQUATION FOR THE SPECIFIC HEAT AS A</td>
</tr>
<tr>
<td></td>
<td></td>
<td>FUNCTION OF TEMPERATURE</td>
</tr>
<tr>
<td>DELTA</td>
<td>R*4</td>
<td>ALLOWABLE RANGE FOR THE INITIAL SPHERE TEMPERATURE</td>
</tr>
<tr>
<td>DIA</td>
<td>R*4</td>
<td>DIAMETER OF SPHERE</td>
</tr>
<tr>
<td>EMIS</td>
<td>R*4</td>
<td>EMISSIVITY OF THE SPHERE</td>
</tr>
<tr>
<td>EMIV</td>
<td>R*4</td>
<td>APPLIED VOLTAGE TO THE ELECTRIC MOTOR</td>
</tr>
<tr>
<td>FBDIA</td>
<td>R*4</td>
<td>FLUIDIZED BED INSIDE DIAMETER</td>
</tr>
<tr>
<td>FBH</td>
<td>R*4</td>
<td>FLUIDIZED BED HEIGHT</td>
</tr>
<tr>
<td>HCONV</td>
<td>R*4</td>
<td>CONVECTIVE HEAT TRANSFER COEFFICIENT</td>
</tr>
<tr>
<td>HT</td>
<td>R*4</td>
<td>TOTAL HEAT TRANSFER COEFFICIENT</td>
</tr>
<tr>
<td>MDIR</td>
<td>I*3</td>
<td>DIRECTION OF SPHERE MOTION IN THE FLUIDIZED BED</td>
</tr>
<tr>
<td></td>
<td></td>
<td>0-MOVING DOWN, 1-MOVING UP</td>
</tr>
</tbody>
</table>
PDIA  R*4  DIAMETER OF THE PULLEY
PD   R*4  PRESSURE DROP ACROSS THE BED
PLATNO R*4  DISTRIBUTOR PLATE NUMBER
POUT R*4  PRESSURE AT THE OUTLET OF ROTAMETER IN CMS OF Hg
PSTD R*4  STANDARD PRESSURE
QRS R*4  VOLUME FLOW RATE AT STD CONDITION
QOP R*4  VOLUME FLOW RATE AT THE OPERATING CONDITION
QBED R*4  VOLUME FLOW RATE IN SIDE THE BED
R  R*4  COEFFICIENT OF CORRELATION.
RHOS R*4  DENSITY OF SOLID SPHERE
RMTEMP R*4  ROOM TEMPERATURE
ROTM R*4  ROTAMETER READING
ROTMNO R*4  ROTAMETER CATALOG NUMBER
RPOUT R*4  PRESSURE AT THE OUTLET OF THE ROTAMETER IN CM OF WATER
RSQ R*4  SQUARE OF COEFFICIENT OF CORRELATION
RUNNO R*4  RUN NUMBER
SETTI R*4  SET INITIAL TEMPERATURE
SIZENO R*4  PARTICLE SIZE NUMBER REPRESENTING THE SIZE RANGE
SIZE1 R*4  UPPER SIZE RANGE OF THE PARTICLES IN MICRONS
SIZE2 R*4  LOWER SIZE RANGE OF THE PARTICLES IN MICRONS
SLOP R*4  SLOP OF THE COOLING CURVE ON A LOG-LOG BASES
SPHEAT R*4  SPECIFIC HEAT OF THE SPHERE
S1  R*4  SPHERE MATERIAL IDENTIFICATION KEY(S1=
S2  R*4  SPHERE SIZE IDENTIFICATION KEY
TAO  R*4  TIME CONSTANT (SEC)
TAO1 R*4  
TAVG R*4  AVERAGE TEMPERATURE
TFLUID R*4  BED TEMPERATURE IN K
TINI R*4  INITIAL TEMPERATURE OF THE SPHERE
TSTD R*4  STANDARD TEMPERATURE
TTIME R*4  TOTAL TIME
VOL R*4  VOLUME OF THE SPHERE
VSPH R*4  LONGITUDINAL SPHERE VELOCITY INSIDE
THE FLUIDIZED BED

VSPH1  R=4  SUPERFICIAL VELOCITY OF THE FLUIDIZING GAS

NAME TYPE
E    R=4  THERMOCOUPLE READING IN MILIVOLTS.
TEMP  R=4  TEMPERATURE OF THE SPHERE IN C
TEMPK R=4  TEMPERATURE OF THE SPHERE IN K
TIME  R=4  TIME IN SECS.

C******************************************************************

DIMENSION TIME(300),TEMP(300),E(300),TEMPK(300)
DIMENSION TR(300),TRL(300)
DIMENSION TRL1(300)
CHARACTER*40 RA,RB
INTEGER S1,S2,SIZE1,SIZE2,MDIR
OPEN(UNIT=1,FILE='FORO01.DAT',STATUS='OLD')
OPEN(UNIT=4,FILE='FOR004.DAT',STATUS='OLD')
OPEN(UNIT=12,FILE='FOR012.DAT',STATUS='OLD')
OPEN(UNIT=8,FILE='FOR008.DAT',STATUS='OLD')
PI=3.1415926
READ(8,100,END=130) SI
READ(8,100,END=130) S2
READ(8,110,END=130) SIZENO
READ(8,110,END=130) FBH
READ(8,110,END=130) BSH
READ(8,110,END=130) TFC
READ(8,110,END=130) PD
READ(8,110,END=130) RMTEMP
READ(8,110,END=130) ATMP
READ(8,110,END=130) RPOUT
READ(8,120,END=130) RUNNO
READ(8,120,END=130) EMIS
READ(8,120,END=130) SETTI
READ(8,120,END=130) DELTA
READ(8,120,END=130) TDIF
READ(8,120,END=130) CA
READ(8,120,END=130) CB
READ(8,120,END=130) ROTM
READ(8,120,END=130) PLATNO
READ(8,120,END=130) ROTMNO
READ(8,120,END=130) PSTD
READ(8,120,END=130) TSTD
READ(8,120,END=130) FBDIA
READ(8,120,END=130) EMIV
READ(8,120,END=130) PDIA
READ(8,120,END=130) VSPH1
READ(8,120,END=130) HG
READ(8,120,END=130) AMPL
READ(8,120,END=130) SPFREQ

100 FORMAT(42X,I4)
110 FORMAT(42X,G10.4)
120 FORMAT(42X,G10.4)
130 WRITE(6,140)
140 FORMAT(’ FOLLOWING ARE THE INPUT PARAMETERS’)
WRITE(6,150) S1,SIZEN0
150 FORMAT(’ S1=1 FOR BRONZ AND 0 FOR COPPER-’,I4,/)
 WRITE(6,160) FBH,BSH,TFC,S2,PD,RMTEMP,ATMP,RPOUT
160 FORMAT(’ BED HEIGHT IN FLUIDIZED STATE(CM)-’,F10.4/,)
2 ’ PACKED BED HEIGHT(CM)-’,F10.4/,
3 ’ BED TEMPERATURE (K)-’,F10.4/,
4 ’ SPHERE NUMBER-’,I4/,
5 ’ PRESSURE DROP ACROSS THE BED (CM OF WATER)-’,F10.4/,
6 ’ ROOM TEMPERATURE (C)-’,F10.4/,
7 ’ ATMOSPHERIC PRESSURE (CM OF Hg)-’,F10.4/,
8 ’ ROTAMETER OUTLET PRESSURE(CM OF WATER)-’,F10.4/)
'EMISSIVITY-',F10.4/,
'SET INITIAL TEMP.-',F10.4/,
'DELTA-',F10.4/,
'TDIF TEMP. DIFF. BET. A STEP-',F10.4/,
'ROTAMETER READING-',F10.4)
WRITE(6,170) PSTD,TSTD,FBDIA,EMIV,PDIA
170 FORMAT('STD PRESSURE IN CM OF Hg-',F10.4/,
'STD TEMPERATURE K-',F10.4/,
'INSIDE DIAMETER OF BED CM-',F10.4/,
'INPUT VOLTS TO THE MOTOR-',F10.4/,
'PULLEY DIAMETER CM-',F10.4)
IF(PLATNO.EQ.1.) GO TO 171
GO TO 190
171 WRITE(6,200)
200 FORMAT('ORIGINAL DISTRIBUTOR PLATE USED')
190 WRITE(6,*)S1,S2
TYPE 210
210 FORMAT('DO YOU NEED THE HEADING?,YES=1,N0=0',$',')
READ(5,*) N1
TYPE 220
220 FORMAT('ENTER THE ROTAMETER READINGS>',$',')
READ(5,*)ROTM
TYPE 230
230 FORMAT('ENTER SETTI IN K>',$',')
READ(5,*)SETTI
TYPE 240
240 FORMAT('ENTER TDIF IN K>',$',')
READ(5,*)TDIF
C
READ TIME AND THERMOCOUPLE VOLTAGE DATA FROM DATA FILE
C
READ(1,246)FBH,PD,RPOUT,RUNNO,ROTMNO,TFC
246 FORMAT(F10.6/F10.6/F10.6/F10.6/F10.6/F10.6)
WRITE(6,247) ROTMNO, TFC, MDIR, VSPH1
247 FORMAT(ROTMNO, TFC, MDIR, VSPH1', 2F10.4, I4, F10.4)
DO 250 K1 = 1, 9999
READ(1, 260, END = 270) TIME(K1), E(K1)
250 CONTINUE
260 FORMAT(F10.6, 2X, F12.9)
270 N = K1 - 1
N3 = 1
WRITE(6, 271) FBH, PD, RPOUT, RUNNO, ROTMNO, MDIR, VSPH1
271 FORMAT(F10.6/F10.6/F10.6/F10.6/F10.6/F10.6/I3/F10.6)
C
C-------------------------------
C IF N3 IS EQUAL TO 1 OPEN A NEW OUTPUT FILE
C
C-------------------------------
C IF (N1.EQ.1) GO TO 280
C
C-------------------------------
C IF N3 EQUAL TO 0 NO HEADING REQUIRED AND WRITE THE OUTPUT
C DATA AT THE END OF THE OUTPUT FILE FOR001.DAT
C-------------------------------
C
300 READ(4, 290, END = 310) RA
GO TO 300
310 CONTINUE
301 READ(12, 290, END = 302) RB
GO TO 301
302 CONTINUE
290 FORMAT(A30)
280 WRITE(6, *) SIZE
C
C-------------------------------
C THE KEY S1 SELECTS THE MATERIAL OF THE SPHERE. FOR BRONZE
C SPHERE S1 = 1
C AND FOR COPPER SPHERE S1 = 2
C
C----------------------------------------------------------------------
C
IF (Sl.EQ.1) GO TO 320
RHOS=8954.0
SPHEAT=383.1
GO TO 330
320 RHOS=8666.0
SPHEAT=343.0

C
C----------------------------------------------------------------------
C
THE KEY S2 SELECTS THE SIZE OF THE SPHERE
C
C----------------------------------------------------------------------
C
330 GO TO (340,350,360,370) S2
340 DIA=0.0095
    CDIA=0.00
    GO TO 380
350 DIA=0.01
    CDIA=0.0
    GO TO 380
360 DIA=0.014
    CDIA=0.0
    GO TO 380
370 DIA=0.02
    CDIA=0.0

C
C CALCULATE THE SURFACE AREA AND VOLUME OF THE SPHERE
C
C
380 VOL=4.0*PI*((DIA/2.)**3)/3.
AREA = 4 * PI * ((DIA/2.)**2) - PI * ((CDIA/2.)**2)
BOLTZC = 5.6700E-08
KEY1 = 0

C——— SELECT THE INITIAL TEMPERATURE RANGE ————

TFLUID = TFC + 273.2
SETTI1 = SETTI + DELTA
TIPD = SETTI - DELTA

C——— CONVERT THE THERMOCOUPLE VOLTAGE INTO TEMPERATURE IN K ————

DO 390 I = 1, N
CALL MVTEMP(E(I), TEMPK(I))
390 CONTINUE

C——— SELECT THE INITIAL INDEX I FROM THE INITIAL TEMPERATURE RANGE ————

DO 400 I = 1, N
IF ((TEMPK(I) .LE. SETTI1) .AND. (TEMPK(I) .GE. TIPD)) GO TO 410
400 CONTINUE
410 N3 = I
C------------
C CALCULATE THE TEMPERATURE RANGE FOR WHICH THE HEAT TRANSFER
C COEFFICIENT HAS TO BE CALCULATED
C------------

420 TSTEP=TEMPK(N3)-TDIF
TSTEPH=TSTEP+DELTA
TSTEPL=TSTEP-DELTA

C

WRITE(6,430) N3,TEMPK(N3)
430 FORMAT(' *****N3,T(N3)',I3,F10.4)

C------------
C SELECT THE FINAL INDEX N4, FROM THE TEMPERATURE STEP
C------------

DO 440 I=N3,N
   IF((TEMPK(I).LE.TSTEPH).AND.(TEMPK(I).GE.TSTEPL)) GO TO 450
440 CONTINUE
450 N4=I

C------------
C IF FINAL INDEX IS GREATER THEN NUMBER OF DATA POINTS TAKEN
C THEN SET N4 EQUAL TO THE NUMBER OF DATA POINTS
C------------

IF(KEY3.EQ.1) N4=N
IF(N4.GE.N) N4=N

C------------
C INITIALIZE PARAMETERS USED FOR CALCULATING THE SLOP
C------------
SIGXY = 0.0
SIGXSQ = 0.0
SIGX = 0.0
SIGY = 0.0
SIGYSQ = 0.0
SIGXYC = 0.0
SIGXSC = 0.0
SIGXC = 0.0
SIGYC = 0.0
SIGYSC = 0.0
WRITE(6,*) N3, N4

C
C---------------------------------------------------------------------
C
C CALCULATE TIME FROM THE DATA
C (USE THIS WHEN ADM DATA ACQUISITION PROGRAM IS USED)
C---------------------------------------------------------------------
C
C DUTIME = TIME(N3)
C DO 460 I3 = N3, N4
C WRITE(4,470) I3, DUTIME, TIME(I3), TIME(N3)
C 470 FORMAT(' I3, DUTIME, TIME(I3), TIME(N3)', I3, 1X, 3(2X, F10.4))
C TIME(I3) = TIME(I3) - DUTIME
C WRITE(4,480) I3, N3, N4, TIME(I3)
C 480 FORMAT(' I3, N3, N4, TIME(I3)', I3, 1X, I3, 1X, I3, 1X, F10.4, 2X, F10.4)
C 460 CONTINUE
C
C---------------------------------------------------------------------
C
C CALCULATE THE NECESSARY VARIABLES NEEDED FOR
C LINEAR REGRESSION
C---------------------------------------------------------------------
C
DO 490 I = N3, N4
TR(I) = (TFLUID - TEMPK(I))/(TFLUID - TEMPK(N3))
IF I=N3 THEN TR=0.0 SO AVOID LOG OF 0.0

IF(I.EQ.N3) GO TO 500
GO TO 510

500 TRL(N3)=0.0
TRL1(N3)=0.0
GO TO 520

FOR VARIABLE SPECIFIC HEAT OF THE COPPER SPHERE
CB IS NOT EQUAL TO ZERO

510 IF(CB.EQ.0.0) GO TO 530
TRL(I)=LOG(TR(I))*(CA+CB*TFLUID)+CB*(TEMPK(I)-TEMPK(N3))
TRL1(I)=LOG(TR(I))
GO TO 520

530 TRL(I)=LOG(TR(I))

520 E(I)=E(I)*1000.
TIME1=TIME(I)-TIME(N3)

SIGXY=SIGXY+TIME(I)*TRL(I)
SIGXSQ=SIGXSQ+TIME(I)^2
SIGX=SIGX+TIME(I)
SIGY=SIGY+TRL(I)
SIGYSQ=SIGYSQ+TRL(I)^2

REGRESSION ANALYSIS FOR THE CONSTANT SPECIFIC HEAT OF COPPER SPHERE

SIGXYC=SIGXYC+TIME(I)*TRL1(I)
SIGXSC=SIGXSC+TIME(I)*TIME(I)
SIGXC=SIGXC+TIME(I)
SIGYC=SIGYC+TRL1(I)
SIGYSC=SIGYSC+TRL1(I)^2

490 CONTINUE
C

C CALCULATE SLOP AND INTERCEPT
C

C

NN=N4-N3+1.0
DEN0=NN*SIGXSQ-SIGX*SIGX
DEN01=SQRT(DEN0*(NN*SIGYSQ-SIGY*SIGY))
SLOP1=SIGXY/SIGXSQ
SLOP=(NN*SIGXY-SIGX*SIGY)/DEN0
WRITE(6,491)SLOP1,SLOP

491 FORMAT(' SLOP1,SLOP',2(1X,F10.4))
AINT=(SIGY*SIGXSQ-SIGX*SIGXY)/DEN0

C CALCULATE SLOP AND INTERCEPT FOR THE CONSTANT SPECIFIC
C HEAT OF THE COPPER SPHERE
C

C

DENOC=NN*SIGXSC-SIGXC*SIGXC
DENOC1=SQRT(DENOC*(NN*SIGYSC-SIGYC*SIGYC))
SLOPC=(NN*SIGXYC-SIGXC*SIGYC)/DENOC
AINTPC=(SIGYC*SIGXSC-SIGXC*SIGXY)/DENOC

C

C CALCULATE COEFFICIENT OF REGRESSION AND TIME CONSTANT
C

C

R=(NN*SIGXY-SIGX*SIGY)/DEN01
RC=(NN*SIGXYC-SIGXC*SIGYC)/DENOC1
RSQ=R*R
TAO=1./ABS(SLOP)
TAO1=1.0/(ABS(SLOP1))
TAOC=1.0/(ABS(SLOPC))

C CALCULATE HEAT TRANSFER COEFFICIENT DUE TO RADIATION
C

TTIME=TIME(N4)-TIME(N3)
TINI=TEMPK(N3)
X1=0.5*TAO*(TFLUID*TFLUID-2.*TFLUID*TINI+TINI*TINI)
  *(EXP(-2.*TTIME/TAO))
X2=2.*TFLUID*TFLUID-2.*TFLUID*TAO*(TFLUID-TINI)
  *(EXP(-TTIME/TAO))
X3=2.*TFLUID+TAO*(TFLUID-TINI)*(EXP(-TTIME/TAO))
X4=0.5*TAO*(TFLUID*TFLUID-2.*TFLUID*TINI+TINI*TINI)
X5=2.*TFLUID*TFLUID-2.*TFLUID*TAO*(TFLUID-TINI)
X6=2.*TFLUID+TAO*(TFLUID-TINI)
AVGHR=BOLTZC*EMIS*((X2-X1)*(X3))-((X5-X4)*(X6)))/TTIME
  *(TEMPPK(N3)+TFLUID)

C TOTAL HEAT TRANSFER COEFFICIENT FOR VARIABLE SPECIFIC HEAT
C OF THE COPPER SPHERE
HT=RHOS*VOL/(TAO*AREA)
HT1=RHOS*VOL/(TAO1*AREA)
HTC=RHOS*SPHEAT*VOL/(TAO*AREA)

C TOTAL HEAT TRANSFER FOR CONSTANT SPECIFIC HEAT
C OF THE COPPER SPHERE
IF(CB.LE.0.0) HT=RHOS*SPHEAT*VOL/(TAO*AREA)
IF(CB.LE.0.0) HT1=RHOS*SPHEAT*VOL/(TAO1*AREA)

C CALCULATE THE CONVECTION PART OF THE HEAT TRANSFER COEFFICIENT
C


HCONV=HT-AVGHR
HCONVC=HTC-AVGHR
TFILM=(TEMPK(N3)+TFLUID)/2.0

C C C CALCULATE THE CONDUCTIVITY OF THE AIR FROM THE FILM
C TEMPERATURE
C
C CALL AIRPRO(TFILM,AKAIR)
ANUEXP=HCONV*DIA/AKAIR
TAVG=(TEMPK(N3)+TEMPK(N4))/2.0
DENOL=LOG((TEMPK(N4)-TFLUID)/(TEMPK(N3)-TFLUID))
TLMTD=(TEMPK(N4)-TEMPK(N3))/DENOL
TLMTD1=TLMTD+TFLUID
ALTMLT=LOG(TLMTD1)
TOF=TINI-TFLUID
ALHC=LOG(HT)
ALHM=LOG(HT1)
WRITE(6,650) AVGHR,HT,HT1,HCONV

650 FORMAT(' AVGHR,HT,HT1,HCONV',4(2X,F10.4))

C C C CALCULATE THE SUPERFICIAL VELOCITY OF FLUIDIZING GAS
C
C CALL VSUP(ROTMNO,ROTM,PSTD,RMTEMP,TSTD,ATMP,RPOUT,FBDIA *
,VSUF,PD,HG)
C
C C C CALCULATE THE SPHERE VELOCITY FOR THE CASE OF A MOVING SPHERE
C
IF(EMIV.GT.0.0) CALL SPHVEL(EMIV,PDIA,VSPH)

C
C IF N1=1 WRITE THE TITLE IN THE OUTPUT DATA FILE
C IF N1=0 SKIP THE TITLE
C
IF(N1.NE.1) GO TO 670
IF(KEY1.EQ.1) GO TO 670
IF(KEY3.EQ.1) GO TO 670

WRITE AN APPROPRIATE HEADING FOR A STATIONARY SPHERE

IF(EMIV.LE.0.0) GO TO 680
WRITE AN APPROPRIATE HEADING FOR COPPER/BRONZE SPHERE

IF(S1.EQ.1) GO TO 690
WRITE(4,700)
700 FORMAT(6X,' HEAT TRANSFER OF A OSCILLATING COPPER SPHERE',/)
GO TO 710
690 WRITE(4,720)
720 FORMAT(6X,' HEAT TRANSFER OF A MOVING BRONZE SPHERE',/)
710 WRITE(4,730) AMPL
730 FORMAT(6X,' AMPLITUDE OF OSCILLATION OF SPHERE (CM)=',F10.4)
WRITE(4,731) SPFREQ
731 FORMAT(6X,' FREQUENCY OF THE SPHERE (C/S)=',F10.4)
GO TO 740
680 IF(S1.EQ.1) GO TO 750
WRITE(4,760)
760 FORMAT(6X,' HEAT TRANSFER OF A STATIONARY COPPER SPHERE')
GO TO 740
750 WRITE(4,770)
HEAT TRANSFER OF A STATIONARY BRONZE SPHERE

IF (PLATNO.EQ.1.) GO TO 780
WRITE(4,790)
FORMAT(6X,' DISTRIBUTOR PLATE USED# A3420141'/
   * 6X,'STD PERMEABILITY=50, AVG PORE SIZE=100 MICRONS')
GO TO 800
WRITE(4,810)
FORMAT(6X,' INSIDE DIAMETER OF THE BED (CM) = ',F10.4)

C
C SELECT THE PARTICLE RANGE FROM THE VALUE OF SIZENO
C
GO TO (830,840,850,860,865) SIZENO

SIZE1 = 147
SIZE2 = 126
GO TO 870

SIZE1 = 44
SIZE2 = 0
GO TO 870

SIZE1 = 450
SIZE2 = 355
GO TO 870

SIZE1 = 2
SIZE2 = 2
GO TO 870

SIZE1 = 2
SIZE2 = 2

WRITE(4,880) SIZENO,SIZE1,SIZE2
FORMAT(6X,' SIZENO#',F3.1,3X,' GLASS BEADS OF /
   * SIZE = ',I4,'-',I4,' MICRONS')
GO TO (900,901,902,903,904,905,906) ROTMNO

WRITE(4,910)
FORMAT(6X,' ROTAMETER CATALOG# 450-700, CREST DIV.')
GO TO 920

WRITE(4,930)
FORMAT(6X,' ROTAMETER CATALOG# 250-8, DEVCO TUBE')
GO TO 920
298

902 WRITE(4,931)
931 FORMAT(6X,' ROTAMETER BROOKS INSTRUMENT DIV.')
GO TO 920

903 WRITE(4,932)
932 FORMAT(6X,' ROTAMETER F&P CO. PRECISION BORE FLOWMETER')
GO TO 920

904 WRITE(4,933)
933 FORMAT(6X,' BROOKS ROTAMETER -TUBE SIZE-R7M25-1')
GO TO 920

905 WRITE(4,934)
934 FORMAT(6X,' SCHUTTLE & KOERTING CO. ROTAMETER')

906 WRITE(4,907)
907 FORMAT(6X,' BROOK INSTRUMENT DIV.,(%*8.10CC/MIN)')

920 WRITE(4,940) S2
940 FORMAT(6X,' SPHERE NUMBER#=',14)

WRITE(4,950) RMTEMP
950 FORMAT(6X,' ROOM TEMPERATURE (C)=',F10.4)

WRITE(4,960) ATMP
960 FORMAT(6X,' ATMOSPHERIC PRESSURE(CM OF Hg)=',F10.4)

WRITE(4,970) PSTD
970 FORMAT(6X,' STD PRESSURE IN CM OF Hg=',F10.4)

WRITE(4,990) RHOS
990 FORMAT(6X,' DENSITY OF THE SPHERE MATERIAL(Kg/M3)=',F10.4)

WRITE(4,1000) SPHEAT
1000 FORMAT(6X,' SPECIFIC HEAT OF THE SPHERE MATERIAL()=',F10.4)

WRITE(4,1010) EMIS
1010 FORMAT(6X,' EMISSIVITY OF SPHERE=',F10.4)

WRITE(4,1020) DIA
1020 FORMAT(6X,' DIAMETER OF SPHERE(MTS)=',F10.4)

WRITE(4,1030) BSH
1030 FORMAT(6X,' PACKED BED HEIGHT (CMS)=',F10.4)

WRITE(4,1040) TFLUID
1040 FORMAT(6X,' BED TEMPERATURE(K)= ',F10.4)

WRITE(4,1050) CA,CB
1050 FORMAT(6X,' CA=',F10.4,3X,'CB=',F10.4//)

WRITE(4,1060)
299

1060 FORMAT(5X, 'ROTAMETER', 2X, 'VSUF', 3X, 'N3', 2X, 'N4', 6X,
     * 'TIME', 9X, 'R', 6X, 'TINITIAL', 3X, 'TEND')
WRITE(4, 1070)

     * , 4X, 'HT1'/5X, 'RUNNO', 5X, 'TAVG', 7X,
     * , 5X, 'FBH', 5X, 'ROTMNO', 3X, 'HCONVC', 3X, 'RC', 3X 'MDIR')

670 WRITE(4, 1080) ROTM, VSUF, N3, N4, TTIME, R, TEMPK(N3), TEMPK(N4),
     * HCONV, AVGHR, HT, ANUEXP, HT1, RUNNO, TAVG, TLMTD, ALTMLT, ALHC, ALHM
     * , PD, RPOUT, FBH, ROTMNO, HCONVC, RC, MDIR

1080 FORMAT(5X, F6.2, 4X, F6.2, 1X, I3, 2X, I3, 3X, F8.4, 3(2X, F8.4)
IF(KEY3.EQ.1) GO TO 1110
IF(N4.GE.N) GO TO 1100
N3=N4
KEY1=1
GO TO 420

1100 KEY3=1
GO TO 421

1110 WRITE(4, 1090)

1090 FORMAT(6X, '--------------------------------')
WRITE(12, 640) VSUF, HCONV

640 FORMAT(F10.4, 1X, F10.4)
CLOSE(UNIT=4)
CLOSE(UNIT=12)
CLOSE(UNIT=3)
CLOSE(UNIT=8)
CLOSE(UNIT=1)
END

C
C

SUBROUTINE MVTMP(E, TEMPK)
X1=0.1008609
X2=25727.94369
X3=-767345.8295
300

X4=78025595.81
X5=-9247486589.0
X6=697688000000.
X7=-2.66192E+13
X8=3.94078E+14
TEMP1=X1+X2*E+X3*(E**2)
TEMP2=X4*(E**3)+X5*(E**4)
TEMP3=X6*(E**5)+X7*(E**6)+X8*(E**7)
TEMP=TEMP1+TEMP2+TEMP3
TEMPK=TEMP+273.2
RETURN
END

C

SUBROUTINE AIRPRO(TFILM,AKAIR)
B=7402.0E-08
A=3945.0E-06
AKAIR=B*TFILM+A
RETURN
END

SUBROUTINE VSUP(ROTMNO,ROTM,PSTD,RMTEMP,TSTD,ATMP,RPOUT,FBDIA
* VSUF,PD,HG)
PI=3.1415926
IF (HG.EQ.1.0) GO TO 100
POUT=ATMP+0.074*RPOUT
GO TO 110
100 POUT=ATMP+RPOUT
110 FBAREA=(PI/4.0)*((FBDIA*FBDIA)
GO TO (10,20,30,40,50,60,80) ROTMNO
10 QRS=1.5929*ROTM-13.9910
GO TO 70
C 20 IF (ROTM.GT.3.0) GO TO 200
C QRS=-2.4760*ROTM*ROTM+34.5873*ROTM+0.4431
C GO TO 70
20 QRS=27.9390*ROTM-6.9009
GO TO 70
30 QRS=9.6269*ROTM+23.4293
GO TO 70

C

SUBROUTINE AIRPRO(TFILM,AKAIR)
B=7402.0E-08
A=3945.0E-06
AKAIR=B*TFILM+A
RETURN
END

SUBROUTINE VSUP(ROTMNO,ROTM,PSTD,RMTEMP,TSTD,ATMP,RPOUT,FBDIA
* ,VSUF,PD,HG)
PI=3.1415926
IF (HG.EQ.1.0) GO TO 100
POUT=ATMP+0.074*RPOUT
GO TO 110
100 POUT=ATMP+RPOUT
110 FBAREA=(PI/4.0)*((FBDIA*FBDIA)
GO TO (10,20,30,40,50,60,80) ROTMNO
10 QRS=1.5929*ROTM-13.9910
GO TO 70
C 20 IF (ROTM.GT.3.0) GO TO 200
C QRS=-2.4760*ROTM*ROTM+34.5873*ROTM+0.4431
C GO TO 70
20 QRS=27.9390*ROTM-6.9009
GO TO 70
30 QRS=9.6269*ROTM+23.4293
GO TO 70
40  QRS=22.9046*ROTM+68.6256  
   GO TO 70
50  QRS=7.1395*ROTM-4.0646  
   GO TO 70
60  QRS=9.7284*ROTM+440.9562  
   GO TO 70
80  QRS=8.10*ROTM/60.00  
    QROP=(((PSTD/POUT)*(273.2+RMTEMP)/(294.1))*QRS  
   GO TO 90

C-----------------------------------------------
C    QROP-OPERATION CONDITION ROTAMETER FLOW RATE CC/SEC
C-----------------------------------------------
C
70  QROP=(((PSTD/TSTD)*(273.2+RMTEMP)/(POUT))**0.5)*QRS
90  PBED=(PD*0.074+ATMP+ATMP)*.5  
    QBED=(POUT/PBED)*QROP  
    VSUF=QBED/FBAREA  
    RETURN  
END

C
C USE THIS ONLY FOR RUBBER PULLEY DRIVEN SPHERE
C (OLD ARRANGEMENT)
C
SUBROUTINE SPHVEL(EMIV,PDIA,VSPH)  
PI=3.1415926  
RPM=6.037*EMIV-1.1515  
RPS=RPM/60.0  
VSPH=PI*PDIA*RPS  
RETURN  
END
Program IADM

Language: FORTRAN-77

Interactive Analog Data Module (IADM) program.

Function:
This program allows the user to interactively acquire analog data using the PRTIL ADM subroutines.

Algorithms Used:
This program uses the voltage-level external trigger mode to start analog data acquisition. The user is prompted for the appropriate channel to receive the trigger.

Error Conditions:
Most error conditions are handled internally.

Subroutines Referenced:
Functions YES, GETTIM
Subroutines GETFRQ, GETCHN
Subroutine WAIT (from P/0S system library);
ADM Subroutines AIN, AINIT, CLKFRQ, CNVDIF, ERASE, PLOT, PLOTM.

Declare and define all constants:

INTEGER*2 MAXPTS ! Declare the maximum number of data.
PARAMETER (MAXPTS=2000)! Maximum number of data points = 2000.
INTEGER*2 IEFN ! Declare default event flag number.
PARAMETER (IEFN=10) ! Define default event flag number.
INTEGER*2 MDSYN  ! Declare synchronous I/O mode constant.
PARAMETER (MDSYN=1)  ! Define synchronous I/O mode constant.
INTEGER*2 IFORM  ! Declare constant, analog data format.
PARAMETER (IFORM=2)  ! Define double-word analog data format.
INTEGER*2 ITRIG  ! Declare constant, analog trigger type.
PARAMETER (ITRIG=0)  ! Define software trigger type.
INTEGER*2 ITIME  ! Define timeout parameter.
PARAMETER (ITIME=0)  ! Declare no timeout value.
INTEGER*2 SBNOP  ! Declare SBNOP status code.
PARAMETER (SBNOP="40000")  ! Define SBNOP status code.
INTEGER*2 SBTRI  ! Declare SBTRIGC0M status code.
PARAMETER (SBTRI="40040")  ! Define SBTRIGC0M status code.

C******************************************************************
C Declare external functions
C******************************************************************
C
EXTERNAL YES  ! Declare name of external routine.
LOGICAL*2 YES  ! Declare type of external routine.

C******************************************************************
C Declare variable storage locations:
C******************************************************************
C
BYTE FNAME(40)  ! Declare storage for disk file name.
INTEGER*2 ISTAT(2)  ! Declare PRTIL status buffer holder.
INTEGER*2 JDATA(MAXPTS)  ! Declare storage, double integer data.
INTEGER*2 ICOUNT  ! Declare storage, acquired data count.
INTEGER*2 ICHAN(8)  ! Declare desired channel array for AIN.
INTEGER*2 ICONV  ! Declare storage for color descriptors.
INTEGER*2 ICONV  ! Declare storage for conversion format.
INTEGER*2 LENGTH  ! Declare storage for file name length.
INTEGER*2 NUMCHN  ! Declare storage, total # active chans.
INTEGER*2 TRGCHN  ! Declare storage for trigger channel #.
INTEGER*2 MARRAY(4)  ! Declare storage for plotting modes.
INTEGER*2 IPLCHN(8)  ! Declare storage for desired plot chans.
INTEGER*2 NUMPCH  ! Declare storage total # plotted chans.
INTEGER*2 PTSPCH  ! Declare storage # points per channel.
INTEGER*2 CHNPLT  ! Declare storage plotted channel count.
INTEGER*2 NPTS !Declare storage number of data values.
REAL*4 XDELTA !Declare storage for plot x axis increment.
REAL*4 YORG !Declare storage for plot y axis origin.
REAL*4 YDELTA !Declare storage for plot y axis increment.
REAL*4 RDATA(MAXPTS)!Declare storage for real numbered data.
REAL*4 XPLOT(2) !Declare storage for temp plot buffer.
REAL*4 YPLOT(2) !Declare storage for temp plot buffer.
REAL*4 CFREQ !Declare storage for desired frequency.
REAL*4 TFREQ !Declare storage for true frequency.
REAL*4 TINC !Declare storage for time increment.
REAL*4 SECT !Declare storage for total elapsed time.
INTEGER*2 I,J,K0,K1 !Declare index variables

C Let JDATA and RDATA share the same space
EQUIVALENCE(RDATA,JDATA)

C*****************************************************************
C The next section begins the executable portion of the program.
C First identify the program:
C*****************************************************************

C 1 TYPE 9000
9000 FORMAT(//,16X,'General Purpose Real-Time Data Acquisition
iProgram',//)

C*****************************************************************
C Prompt the user for sampling frequency, conversion format, channel
C selection and total test time:
C*****************************************************************

C Get desired, true frequency from user.
CALL GETFRQ(CFREQ,TFREQ)

C*****************************************************************

C TYPE 9010
9010 FORMAT(/' Enable autoranging ?')
IF(YES())
1 THEN
   ICONV=5 !Enable autoranging.
ELSE
ICONV='0' !Disable autoranging.
ENDIF

CALL GETCHN(ICHAN,NUMCHN) !Get desired channel selection.

C Get desired sample count.
CALL GETTIM(TFREQ,NUMCHN,MAXPTS,ICOUNT,SECET)
XDELTA=SECET/5.0
YDELTA=2.0
YORG=-5.0

C*****************************************************************
C                        C*****************************************************************
C                        C*****************************************************************
CALL AINIT(ISTAT,IEFN) !Initialize analog conversion system.
TYPE 9020 !Print status identification.
TYPE 9030,ISTAT(1),ISTAT(2) !Print two-word status array.
TYPE 9040,TRGCHN !Inform user about trigger.

9020 FORMAT(/' AINIT status was: '), 
9030 FORMAT(' ISTAT(1) (octal) = ',06/, 
* ' ISTAT(2) (decimal) = ',16)
9040 FORMAT(/' Hit return when ready ',',I1, ' when ready.')

IF(ISTAT(1) .NE. SBNOP)GO TO 32767
IF(NUMCHN.GT.1)GO TO 32767

1 THEN
   TRGCHN=0
ELSE
   DO 10 I=1,8 !Loop to find the only single channel.
      IF(ICHAN(I).EQ.0) GO TO 10
   TRGCHN=I-1
10 CONTINUE !Bottom of loop to find only channel.
ENDIF

C CALL AIN(ISTAT, 
2 JDATA, !Specify standard status buffer.
3 ICOUNT, !Specify double-word integer data buffer.
4 IEFN, !Specify desired sampled data count.

MDSYN, ! Specify synchronous mode, timeout off.
ICHAN, ! Specify desired channel array.
ICONV, ! Specify desired conversion format.
IFORM, ! Specify double-word integer format.
ITRIG, ! Specify software trigger.
ITIME, ! Timeout based on data acquisition.
CPREQ) ! Specify desired clock frequency.

TYPE 9050 ! Print status identification.

9050 FORMAT(' AIN status was:')
TYPE 9030,ISTAT(1),ISTAT(2) ! Print two-word status array.
IF(ISTAT(1) .EQ. SBNOP .OR. ISTAT(1) .EQ. SBTI) GO TO 15
GO TO 32767 ! End prog due to error
15 NPTS=ISTAT(2) ! NPTS holds the no. of converted values.

C******************************************************************************

C Convert the raw acquired data from the double-integer
C format to standard floating-point format:

C******************************************************************************

CALL CNVDIF(JDATA,RDATA,NPTS)

C******************************************************************************

9060 FORMAT(' Do you want to plot any data ?')
IF(.NOT.YES()) GO TO 40
GO TO 17

C******************************************************************************

5 TYPE 9063
9063 FORMAT(' Do you want to rescale the plot axes ?')
IF(YES()) THEN
   GO TO 6
ELSE
   ENDIF
GO TO 17

6 TYPE 9064,XDELTA,YORG,YDELTA
9064 FORMAT(' The old values are :',/,' x axis delta value = ',
1 1PE14.6,/' y axis origin value = ',1PE14.6,/,'
2    ' y axis delta value = ',1PE14.6,/
TYPE 9065
9065   FORMAT(' Enter a new x axis delta value,
1   y axis origin value,' ,/, 
2   '$em.d y aucis delta value : ')
ACCEPT *,XDELTA,YORG,YDELTA
17   NUMPCH=0
DO 25 I=1,8
    IPLCHN(I)=0
    IF(ICHAN(I).EQ.0) GO TO 25 
    TYPE 9070,I=1-1
9070   FORMAT(' Do you 
vaut 
20   to plot channel ',11,' ? ')
    IF(YES()) THEN
        IPLCHN(I)=1
        NUMPCH=NUMPCH+1
25      TYPE 9080
9080   FORMAT('$Enter Color number (1 to 7) from manual : ')
    ACCEPT *,ICOLOR(I)
    IF(ICOLOR(I) .GT. 0 .AND. ICOLOR(I) .LE. 7) GO TO 25
    TYPE 9090,7
9090   FORMAT(A1,' Improper input .. try again !!!')
    GO TO 20
ELSE
    ENDIF
25     CONTINUE
    IF(NUMPCH.EQ.0) GO TO 40
    TYPE 9100 !Inform user.
9100   FORMAT(/' After the plot is complete you
1   will have 20 seconds'
2   /' to press the PRINT SCREEN key on the PRO... After'
3   /' 20 seconds the screen will be erased and you will'
4   /' be prompted for further plotting.' ,/, 
5   ' PRESS THE RETURN KEY TO CONTINUE ')
C
C
    ACCEPT 9012,I,K !Get a dummy character
9012   FORMAT(Q,A1) !Dummy input
C
C******************************************************************************
C Draw and label the grid for plotting:
C******************************************************************************
C
TINC=1.0/TFREQ
MARRAY(1)=2
MARRAY(2)=7
MARRAY(3)=1
MARRAY(4)=7
PTSPCH=NPTS/NUMCHN

CALL PLOT(ISTAT) !Re-use old status holder for plot stat.
MARRAY,
0.0,
0.0,
0,
0.0,
XDELTA,
YORG,
YDELTA,
' TIME (SECONDS) ',
' VOLTS ')

IF(ISTAT(1) .EQ. 0) GO TO 30
TYPE 9105,ISTAT(1)
GO TO 32767

C******************************************************************************
C Plot the data one channel at a time:
C******************************************************************************
C
CHNPLT=0 !Initialize channels plotted counter.
DO 35 I=1,8
IF(IPLCHN(I).EQ.0) GO TO 35
CHNPLT=CHNPLT+1 !Increment channels plotted counter.
MARRAY(2)=ICOLOR(I)               !Transfer color descriptor.
XPLOT(2)=0.0                     !Initialize x axis values

DO 32 J=1,PTSPCH-1               !Loop for all points per channel.
  XPLOT(1)=XPLOT(2)               !Use old value
  XPLOT(2)=XPLOT(1)+TINC          !Increment the next value
  KO=(J-1)*NUMCHN+CHNPLT         !Calculate index
  K1=J*NUMCHN+CHNPLT             !Calculate index
  YPLOT(1)=RDATA(KO)             !Get first point in segment
  YPLOT(2)=RDATA(K1)             !Get second point in segment
  CALL PLOTM(ISTAT,              '
    2. MARRAY(2),               !Specify plot mode array subset.
    3. XPLOT,                  !Specify x axis explicit values
    4. YPLOT,                 !Specify floating point data to plot.
    5. 2)                     !Specify 2 points to plot.
    IF(ISTAT(1).EQ.0) GO TO 32
    TYPE 9105,ISTAT(1)
    9105 FORMAT('ISTAT(1) (decimal) = ',I8)
    GO TO 32767                !And leave loop
  32 CONTINUE

C Provide a 20-second wait before prompting for more plotting:
  CALL WAIT(20,                   !Specify wait of 20 seconds
    2,                           
    3 ISTAT)
  IF(ISTAT(1).NE.1) GO TO 32767  !Exit on fatal errors.

C
  CALL ERASE
  TYPE 9110
  9110 FORMAT(/' Do you want to replot any data ?'
    IF(YES()) GO TO 5            !Go back to plot more data.

C Allow the user to optionally store the data in a file:
C******************************************************************************
  40 TYPE 9120
  9120 FORMAT(/' Do you want to store the data on disk ?'
    IF(.NOT.YES()) GO TO 70     !No disk storage, check for re-start.
310

45  TYPE 9130
9130  FORMAT(:'/Enter filename (up to 40 characters): ')    
     ACCEPT 8000,LENGTH,FNAME
8000  FORMAT(Q,40A1)
     DO 50 I=LENGTH+1,40
     FNAME(I)=0  !Clear unused locations in name string.
50  CONTINUE
     C!Open file for ASCII data storage.
     OPEN(UNIT=1,TYPE='NEW',NAME=FNAME,ERR=60)
     WRITE(1,9140) TFREQ  !Actual true clock frequency used.
9140  FORMAT(1X,1PE16.7)
     WRITE(1,9140) FLOAT(NUMCHN)
     DO 55 I=1,NPTS
     WRITE(1,9140)RDATA(I)  !All acquired data to disk.
55  CONTINUE
     CLOSE(UNIT=1)  !Close the file.
     GO TO 70
60  TYPE 9150
9150  FORMAT('/ Improper disk attribute .. try again !!!')
     GO TO 45
70  TYPE 9160
9160  FORMAT('/ Do you want to start again ?')
     IF(YES()) GO TO 1  !Go back to start again.
32767  STOP  ' End of program IADM.'
     END
APPENDIX E: EXPERIMENTAL DATA
Table 13.1: Heat transfer coefficient for stationary spheres immersed in a fluidized bed of 355 - 420μm glass particle

<table>
<thead>
<tr>
<th>super-</th>
<th>$D_{sph} = 1.0$</th>
<th>$D_{sph} = 1.4$</th>
<th>$D_{sph} = 2.0$</th>
<th>Bed Height</th>
</tr>
</thead>
<tbody>
<tr>
<td>ficial air velocity cm/s</td>
<td>cm</td>
<td>cm</td>
<td>cm</td>
<td>cm</td>
</tr>
<tr>
<td>1.7000</td>
<td>70.00</td>
<td>50.00</td>
<td>50.00</td>
<td>23.40</td>
</tr>
<tr>
<td>5.3000</td>
<td>65.00</td>
<td>48.00</td>
<td>48.00</td>
<td>23.40</td>
</tr>
<tr>
<td>8.5000</td>
<td>69.00</td>
<td>43.00</td>
<td>43.00</td>
<td>23.40</td>
</tr>
<tr>
<td>11.5582</td>
<td>79.70</td>
<td>58.65</td>
<td>63.00</td>
<td>23.40</td>
</tr>
<tr>
<td>13.1700</td>
<td>101.7</td>
<td>75.30</td>
<td>58.00</td>
<td>23.40</td>
</tr>
<tr>
<td>14.9443</td>
<td>189.9</td>
<td>221.8</td>
<td>204.0</td>
<td>23.40</td>
</tr>
<tr>
<td>16.6835</td>
<td>294.3</td>
<td>286.6</td>
<td>246.0</td>
<td>23.70</td>
</tr>
<tr>
<td>18.5214</td>
<td>351.7</td>
<td>287.8</td>
<td>258.0</td>
<td>24.40</td>
</tr>
<tr>
<td>20.4428</td>
<td>391.9</td>
<td>334.1</td>
<td>309.0</td>
<td>25.10</td>
</tr>
<tr>
<td>22.4324</td>
<td>413.8</td>
<td>372.0</td>
<td>332.0</td>
<td>25.70</td>
</tr>
<tr>
<td>24.2129</td>
<td>430.1</td>
<td>400.1</td>
<td>355.0</td>
<td>26.50</td>
</tr>
<tr>
<td>26.6532</td>
<td>437.1</td>
<td>423.2</td>
<td>375.0</td>
<td>27.30</td>
</tr>
<tr>
<td>28.4243</td>
<td>451.4</td>
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Table 13.2: Heat transfer coefficient for stationary spheres immersed in a fluidized bed of 126 – 147μm glass particle

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Table 13.3: Heat transfer coefficient for stationary spheres immersed in a fluidized bed of 5 – 44μm glass particle

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<th>$D_{sp}h$ cm</th>
<th>$D_{sp}h$ cm</th>
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Table 13.4: Heat transfer coefficient for linearly downward moving 1.0 cm diameter sphere immersed in a fluidized bed of 355 – 420μm glass particle

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<th>( V_{sph} ) cm/s</th>
<th>( V_{sph} ) cm/s</th>
<th>( V_{sph} ) cm/s</th>
<th>( V_{sph} ) cm/s</th>
<th>( V_{sph} ) cm/s</th>
<th>( V_{sph} ) cm/s</th>
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<td>443.9</td>
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</table>
Table 13.5: Heat transfer coefficient for linearly downward moving 1.4 cm diameter sphere immersed in a fluidized bed of 355 – 420μm glass particle

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<td>399.0</td>
<td>390.0</td>
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<td>399.0</td>
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Table 13.6: Heat transfer coefficient for linearly downward moving 2.0 cm diameter sphere immersed in a fluidized bed of 355 – 420μm glass particle

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<th>$V_{sph}$ cm/s</th>
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Table 13.7: Heat transfer coefficient for linearly downward moving 1.0 cm diameter sphere immersed in a fluidized bed of 126 – 147μm glass particle

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Table 13.8: Heat transfer coefficient for linearly downward moving 1.4 cm diameter sphere immersed in a fluidized bed of 126 – 147µm glass particle

<table>
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<tr>
<th>superficial air velocity cm/s</th>
<th>$V_{sph}$</th>
<th>$V_{sph}$</th>
<th>$V_{sph}$</th>
<th>$V_{sph}$</th>
<th>$V_{sph}$</th>
<th>$V_{sph}$</th>
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</tr>
<tr>
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<td>332.0</td>
<td>415.0</td>
<td>447.0</td>
<td>524.0</td>
</tr>
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<td>441.0</td>
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<td>370.0</td>
<td>399.0</td>
<td>454.0</td>
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<td>517.0</td>
<td>530.0</td>
</tr>
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<td>514.0</td>
<td>550.0</td>
<td>549.0</td>
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<td>552.0</td>
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<td>579.0</td>
</tr>
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<td>595.0</td>
<td>602.0</td>
<td>623.0</td>
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<td>636.0</td>
<td>628.0</td>
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<td>634.0</td>
<td>654.0</td>
</tr>
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<td>657.0</td>
<td>637.0</td>
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<td>653.0</td>
<td>665.0</td>
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</table>
Table 13.9: Heat transfer coefficient for linearly downward moving 2.0 cm diameter sphere immersed in a fluidized bed of 126 – 147μm glass particle

<table>
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<tr>
<th>superficial air velocity (cm/s)</th>
<th>( V'_{sph} )</th>
<th>( V_{sph} )</th>
<th>( V'_sph )</th>
<th>( V''_{sph} )</th>
<th>( V'''_{sph} )</th>
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</thead>
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<td>361.0</td>
<td>398.0</td>
<td>355.0</td>
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<td>523.0</td>
<td>514.0</td>
<td>492.0</td>
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<td>539.0</td>
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<td>528.0</td>
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<td>561.0</td>
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<td>594.0</td>
<td>608.0</td>
<td>611.0</td>
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<td>617.0</td>
<td>607.0</td>
<td>610.0</td>
<td>628.0</td>
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<td>624.0</td>
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<td>643.0</td>
<td>632.0</td>
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Table 13.10: Heat transfer coefficient for linearly downward moving 1.0 cm diameter sphere immersed in a fluidized bed of 5 – 44μm glass particle

<table>
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<tr>
<th>Superficial air velocity cm/s</th>
<th>$V_{sph}$ 0.4</th>
<th>$V_{sph}$ 1.1</th>
<th>$V_{sph}$ 1.9</th>
<th>$V_{sph}$ 3.0</th>
<th>$V_{sph}$ 4.6</th>
<th>$V_{sph}$ 7.5</th>
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</thead>
<tbody>
<tr>
<td>0.04150</td>
<td>173.4</td>
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<td>384.4</td>
<td>478.7</td>
<td>546.3</td>
<td>649.5</td>
</tr>
<tr>
<td>0.08690</td>
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<td>283.6</td>
<td>363.9</td>
<td>464.2</td>
<td>563.2</td>
<td>680.7</td>
</tr>
<tr>
<td>0.1245</td>
<td>151.0</td>
<td>316.0</td>
<td>365.0</td>
<td>438.0</td>
<td>453.0</td>
<td>670.0</td>
</tr>
<tr>
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<td>363.0</td>
<td>401.0</td>
<td>469.0</td>
<td>672.0</td>
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<td>458.0</td>
<td>650.0</td>
</tr>
<tr>
<td>0.2687</td>
<td>209.6</td>
<td>303.0</td>
<td>361.0</td>
<td>434.0</td>
<td>516.0</td>
<td>614.0</td>
</tr>
<tr>
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<td>314.0</td>
<td>315.0</td>
<td>373.0</td>
<td>459.0</td>
<td>527.0</td>
</tr>
<tr>
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<td>400.0</td>
<td>540.0</td>
<td>429.0</td>
<td>537.0</td>
</tr>
<tr>
<td>0.9874</td>
<td>553.0</td>
<td>561.0</td>
<td>548.0</td>
<td>607.0</td>
<td>605.0</td>
<td>618.0</td>
</tr>
<tr>
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<td>609.0</td>
<td>654.0</td>
<td>633.0</td>
<td>673.0</td>
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<td>710.0</td>
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<td>698.0</td>
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<td>776.0</td>
<td>732.0</td>
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</table>
Table 13.11: Heat transfer coefficient for linearly downward moving 1.4 cm diameter sphere immersed in a fluidized bed of 5 - 44 µm glass particle

<table>
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<tr>
<th>Superficial air velocity cm/s</th>
<th>( V_{sph} = 0.4 ) cm/s</th>
<th>( V_{sph} = 1.1 ) cm/s</th>
<th>( V_{sph} = 1.9 ) cm/s</th>
<th>( V_{sph} = 3.0 ) cm/s</th>
<th>( V_{sph} = 4.6 ) cm/s</th>
<th>( V_{sph} = 7.5 ) cm/s</th>
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</thead>
<tbody>
<tr>
<td>0.04150</td>
<td>104.4</td>
<td>307.3</td>
<td>346.9</td>
<td>415.8</td>
<td>504.3</td>
<td>597.5</td>
</tr>
<tr>
<td>0.08690</td>
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<td>304.2</td>
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<td>454.4</td>
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<td>277.0</td>
<td>356.0</td>
<td>480.0</td>
<td>662.0</td>
</tr>
<tr>
<td>0.1521</td>
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<td>334.0</td>
<td>328.0</td>
<td>333.0</td>
<td>489.0</td>
<td>612.0</td>
</tr>
<tr>
<td>0.1857</td>
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<td>330.0</td>
<td>359.0</td>
<td>481.0</td>
<td>624.0</td>
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<td>620.0</td>
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<td>673.0</td>
<td>712.0</td>
<td>681.0</td>
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<td>700.0</td>
<td>698.0</td>
<td>763.0</td>
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<td>767.0</td>
<td>808.0</td>
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Table 13.12: Heat transfer coefficient for linearly downward moving 2.0 cm diameter sphere immersed in a fluidized bed of 5 - 44μm glass particle

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<th>superficial air velocity cm/s</th>
<th>$V_{sph} = 0.4$</th>
<th>$V_{sph} = 1.1$</th>
<th>$V_{sph} = 1.9$</th>
<th>$V_{sph} = 3.0$</th>
<th>$V_{sph} = 4.6$</th>
<th>$V_{sph} = 7.5$</th>
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<td>0.04150</td>
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<td>324.5</td>
<td>389.5</td>
<td>423.6</td>
<td>547.6</td>
</tr>
<tr>
<td>0.08690</td>
<td>102.5</td>
<td>237.7</td>
<td>312.3</td>
<td>365.4</td>
<td>457.9</td>
<td>615.6</td>
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<td>294.0</td>
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<td>429.0</td>
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<td>485.0</td>
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<td>321.0</td>
<td>435.0</td>
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<td>481.0</td>
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<td>590.0</td>
<td>626.0</td>
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<tr>
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<td>607.0</td>
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<td>562.0</td>
<td>643.0</td>
<td>632.0</td>
<td>645.0</td>
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<td>700.0</td>
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Table 13.13: Heat transfer coefficient for a copper sphere of 1.0 cm diameter, oscillating at constant peak-to-peak amplitude of 6.9 cm immersed in a fluidized bed of 5 – 44μm glass particle

<table>
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<th>Superficial air velocity cm/s</th>
<th>Frequency =1.11 Hz</th>
<th>Frequency =2.0 Hz</th>
<th>Frequency =2.85 Hz</th>
</tr>
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<tr>
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<td>514.4596</td>
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<td>720.5953</td>
</tr>
<tr>
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<td>732.4918</td>
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<td>913.5311</td>
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Table 13.14: Heat transfer coefficient for a copper sphere of 1.0 cm diameter, oscillating at constant peak-to-peak amplitude of 4.0 cm immersed in a fluidized bed of 5 – 44μm glass particle

<table>
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<th>Superficial air velocity cm/s</th>
<th>Frequency = 1.11 Hz</th>
<th>Frequency = 2.0 Hz</th>
<th>Frequency = 2.85 Hz</th>
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</thead>
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<td>818.2127</td>
<td>816.3910</td>
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Table 13.15: Heat transfer coefficient for a copper sphere of 1.0 cm diameter, oscillating at constant peak-to-peak amplitude of 1.8 cm immersed in a fluidized bed of 5 – 44μm glass particle.

<table>
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<th>Frequency =1.11 Hz</th>
<th>Frequency =2.0 Hz</th>
<th>Frequency =2.85 Hz</th>
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Table 13.16: Heat transfer coefficient for a copper sphere of 1.4 cm diameter, oscillating at constant peak-to-peak amplitude of 6.9 cm immersed in a fluidized bed of 5 – 44μm glass particle

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<th>Frequency =1.11 Hz</th>
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<th>Frequency =2.85 Hz</th>
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<tr>
<td>0.1698</td>
<td>435.5582</td>
<td>680.1333</td>
<td>840.4732</td>
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<td>884.2593</td>
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<td>885.5651</td>
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<td>876.8582</td>
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Table 13.17: Heat transfer coefficient for a copper sphere of 1.4 cm diameter, oscillating at constant peak-to-peak amplitude of 4.0 cm immersed in a fluidized bed of 5 – 44μm glass particle

<table>
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<th>superficial air velocity cm/s</th>
<th>Frequency = 1.11 Hz</th>
<th>Frequency = 2.0 Hz</th>
<th>Frequency = 2.85 Hz</th>
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</thead>
<tbody>
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<td>472.7295</td>
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<td>653.2668</td>
</tr>
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<td>356.8204</td>
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<td>813.3058</td>
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<td>544.1761</td>
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<td>872.9606</td>
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Table 13.18: Heat transfer coefficient for a copper sphere of 1.4 cm diameter, oscillating at constant peak-to-peak amplitude of 1.8 cm immersed in a fluidized bed of 5 - 44µm glass particle

<table>
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Table 13.19: Heat transfer coefficient for a copper sphere of 2.0 cm diameter, oscillating at constant peak-to-peak amplitude of 6.9 cm immersed in a fluidized bed of 5 – 44μm glass particle

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Table 13.20: Heat transfer coefficient for a copper sphere of 2.0 cm diameter, oscillating at constant peak-to-peak amplitude of 4.0 cm immersed in a fluidized bed of 5—44μm glass particle

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<th>Frequency = 2.85 Hz</th>
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Table 13.21: Heat transfer coefficient for a copper sphere of 2.0 cm diameter, oscillating at constant peak-to-peak amplitude of 1.8 cm immersed in a fluidized bed of 5 – 44\(\mu\)m glass particle

<table>
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<th>Frequency =2.85 Hz</th>
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Table 13.22: Heat transfer coefficient for a copper sphere of 1.0 cm diameter, oscillating at constant peak-to-peak amplitude of 6.9 cm immersed in a fluidized bed of 126 – 147μm glass particle

<table>
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<th>Frequency =2.85 Hz</th>
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<td>Frequency $=2.0$ Hz</td>
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Table 13.24: Heat transfer coefficient for a copper sphere of 1.0 cm diameter, oscillating at constant peak-to-peak amplitude of 1.8 cm immersed in a fluidized bed of 126 – 147μm glass particle

<table>
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<tr>
<th>Superficial air velocity cm/s</th>
<th>Frequency =1.11 Hz</th>
<th>Frequency =2.0 Hz</th>
<th>Frequency =2.85 Hz</th>
</tr>
</thead>
<tbody>
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Table 13.25: Heat transfer coefficient for a copper sphere of 1.4 cm diameter, oscillating at constant peak-to-peak amplitude of 6.9 cm immersed in a fluidized bed of 126 – 147μm glass particle

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<th>Frequency =2.0 Hz</th>
<th>Frequency =2.85 Hz</th>
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<td>518.0382</td>
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<td>500.7810</td>
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Table 13.26: Heat transfer coefficient for a copper sphere of 1.4 cm diameter, oscillating at constant peak-to-peak amplitude of 4.0 cm immersed in a fluidized bed of 126 – 147μm glass particle

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<td>658.4618</td>
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Table 13.27: Heat transfer coefficient for a copper sphere of 1.4 cm diameter, oscillating at constant peak-to-peak amplitude of 1.8 cm immersed in a fluidized bed of 126 – 147 μm glass particle

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<th>Frequency =2.0 Hz</th>
<th>Frequency =2.85 Hz</th>
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<td>499.0000</td>
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Table 13.28: Heat transfer coefficient for a copper sphere of 2.0 cm diameter, oscillating at constant peak-to-peak amplitude of 6.9 cm immersed in a fluidized bed of 126 – 147μm glass particle

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<th>Frequency = 2.0 Hz</th>
<th>Frequency = 2.85 Hz</th>
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Table 13.29: Heat transfer coefficient for a copper sphere of 2.0 cm diameter, oscillating at constant peak-to-peak amplitude of 4.0 cm immersed in a fluidized bed of 126 – 147μm glass particle

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Table 13.30: Heat transfer coefficient for a copper sphere of 2.0 cm diameter, oscillating at constant peak-to-peak amplitude of 1.8 cm immersed in a fluidized bed of 126 – 147 μm glass particle

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Table 13.31: Heat transfer coefficient for a copper sphere of 1.0 cm diameter, oscillating at constant peak-to-peak amplitude of 6.9 cm immersed in a fluidized bed of 355 – 420μm glass particle

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<th>Frequency =2.85 Hz</th>
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Table 13.32: Heat transfer coefficient for a copper sphere of 1.0 cm diameter, oscillating at constant peak-to-peak amplitude of 4.0 cm immersed in a fluidized bed of 355 – 420μm glass particle

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Table 13.33: Heat transfer coefficient for a copper sphere of 1.0 cm diameter, oscillating at constant peak-to-peak amplitude of 1.8 cm immersed in a fluidized bed of 355 - 420μm glass particle

<table>
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<th>Frequency = 2.85 Hz</th>
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<td>503.7747</td>
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Table 13.34: Heat transfer coefficient for a copper sphere of 1.4 cm diameter, oscillating at constant peak-to-peak amplitude of 6.9 cm immersed in a fluidized bed of 355 – 420μm glass particle

<table>
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Table 13.35: Heat transfer coefficient for a copper sphere of 1.4 cm diameter, oscillating at constant peak-to-peak amplitude of 4.0 cm immersed in a fluidized bed of 355 – 420μm glass particle

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<th>Frequency =2.0 Hz</th>
<th>Frequency =2.85 Hz</th>
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<td>33.2488</td>
<td>491.7166</td>
<td>503.5739</td>
<td>486.6116</td>
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</table>
Table 13.36: Heat transfer coefficient for a copper sphere of 1.4 cm diameter, oscillating at constant peak-to-peak amplitude of 1.8 cm immersed in a fluidized bed of 355 – 420μm glass particle

<table>
<thead>
<tr>
<th>Superficial air velocity cm/s</th>
<th>Frequency =1.11 Hz</th>
<th>Frequency =2.0 Hz</th>
<th>Frequency =2.85 Hz</th>
</tr>
</thead>
<tbody>
<tr>
<td>11.5403</td>
<td>339.0155</td>
<td>357.3980</td>
<td>395.5992</td>
</tr>
<tr>
<td>13.1960</td>
<td>363.2838</td>
<td>378.5270</td>
<td>386.3303</td>
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<td>358.1369</td>
<td>382.7383</td>
<td>411.7512</td>
</tr>
<tr>
<td>16.7374</td>
<td>415.3810</td>
<td>421.4320</td>
<td>466.5674</td>
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<td>452.3858</td>
<td>455.5369</td>
<td>472.5960</td>
</tr>
<tr>
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<td>476.7601</td>
<td>486.1807</td>
</tr>
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<td>508.1780</td>
<td>497.3738</td>
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<td>28.6316</td>
<td>496.6038</td>
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<td>482.8270</td>
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<td>33.2488</td>
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<td>501.4396</td>
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Table 13.37: Heat transfer coefficient for a copper sphere of 2.0 cm diameter, oscillating at constant peak-to-peak amplitude of 6.9 cm immersed in a fluidized bed of 355 – 420μm glass particle

<table>
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<th>super-ficial air velocity cm/s</th>
<th>Frequency =1.11 Hz</th>
<th>Frequency =2.0 Hz</th>
<th>Frequency =2.85 Hz</th>
</tr>
</thead>
<tbody>
<tr>
<td>11.5403</td>
<td>359.7212</td>
<td>397.3535</td>
<td>329.6180</td>
</tr>
<tr>
<td>13.1960</td>
<td>350.6171</td>
<td>360.7378</td>
<td>314.8083</td>
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<td>341.7200</td>
<td>351.2399</td>
<td>327.9600</td>
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<tr>
<td>16.7374</td>
<td>360.2961</td>
<td>368.4415</td>
<td>334.7017</td>
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<td>383.9271</td>
<td>345.6426</td>
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<td>20.4246</td>
<td>398.0002</td>
<td>396.1328</td>
<td>364.3485</td>
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<td>410.3565</td>
<td>420.1016</td>
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<td>401.3488</td>
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<td>410.0000</td>
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Table 13.38: Heat transfer coefficient for a copper sphere of 2.0 cm diameter, oscillating at constant peak-to-peak amplitude of 4.0 cm immersed in a fluidized bed of 355–420μm glass particle

<table>
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<tr>
<th>Superficial air velocity (cm/s)</th>
<th>Frequency (Hz)</th>
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<th>Frequency (Hz)</th>
</tr>
</thead>
<tbody>
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</tr>
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<td>11.5403</td>
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<td>345.6180</td>
</tr>
<tr>
<td>13.1960</td>
<td>299.6171</td>
<td>365.7378</td>
<td>338.8083</td>
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<td>14.9766</td>
<td>334.7200</td>
<td>373.2399</td>
<td>354.9600</td>
</tr>
<tr>
<td>16.7374</td>
<td>334.2961</td>
<td>395.4415</td>
<td>380.7017</td>
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<td>18.5453</td>
<td>381.7827</td>
<td>408.9271</td>
<td>405.6426</td>
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<td>428.3485</td>
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<td>399.3565</td>
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<td>411.4318</td>
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<td>438.3488</td>
</tr>
<tr>
<td>33.2488</td>
<td>422.9253</td>
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<td>443.0000</td>
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</table>
Table 13.39: Heat transfer coefficient for a copper sphere of 2.0 cm diameter, oscillating at constant peak-to-peak amplitude of 1.8 cm immersed in a fluidized bed of 355 – 420μm glass particle

<table>
<thead>
<tr>
<th>Superficial air velocity (cm/s)</th>
<th>Frequency = 1.11 Hz</th>
<th>Frequency = 2.0 Hz</th>
<th>Frequency = 2.85 Hz</th>
</tr>
</thead>
<tbody>
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<td>11.5403</td>
<td>302.1776</td>
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<td>364.8348</td>
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