MASTER

Progress Report for Research on

# HEAT TRANSFER TO TUBES IN FREEBOARD SPACE OF FLUIDIZED BED COMBUSTORS

Period of

August 1, 1980 to January 31, 1981

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## DISCLAIMER

## SCOPE OF THE PROJECT

The objective of the research is to carry out an experimental study of heat transfer characteristics for tubes located in the freeboard region of fluidized beds. The heat transfer coefficients will be measured and experimental techniques for determining solid-contact behavior on the surface of horizontal tubes will be evaluated. Data will be obtained for various bed particles over a range of fluidization air flow rates, and for a range of tube elevations. Tests will be performed in room temperature beds and in high temperature beds. The anticipated results are:

(a) Experimental data for heat transfer in freeboard region for single tubes, at room temperature and at FBC operating temperatures;

(b) Determination of effect of operating variables (gas velocity, tube elevation, particle size, particle density, etc.) on heat transfer coefficients;

(c) Indication of possible experimental method for determination of dominant heat transfer mechanisms in freeboard region, from solid contact measurements;

(d) Derivation of phenomenological models and correlations.

## PROGRESS IN THE PROJECT

This is the first progress report which covers the period of August 1, 1980 to January 31, 1981. A computer literature search was carried out to find the latest developments in this field. There was no new information outside the information given in the proposal of this research, except the very recent paper of Wood et al. [1]. They carried out tests with beds of silica sand (0.93 mm mean diameter) to measure the average heat transfer coefficient in freeboard region of fluidized beds with different static bed heights.

The existing room temperature fluidized bed facility was modified to carry out tests at high flow rates. A cyclone separator with recycling system was added as shown in Figure 1. To get rid of fluctuation problems caused by static charge on thermocouple readings, a new steam generator was added. The electrically heated steam generator uses 5 kw power, and steam is injected into the inlet air by a specially designed nozzle. It was observed that static charge difficulties were eliminated when relative humidity of inlet air is raised above 65 percent.

Tests were carried out to see the effect of increase in humidity of air on heat transfer coefficients. At low air flow rates, where the static charge effect is low, data without injection of steam were compared with the results for the same flow rates with steam injection. Increasing relative humidity to 65 percent changed the heat transfer coefficient by less than 2 percent.

There are several investigations on heat transfer behavior of tubes in the submerged region, some of which

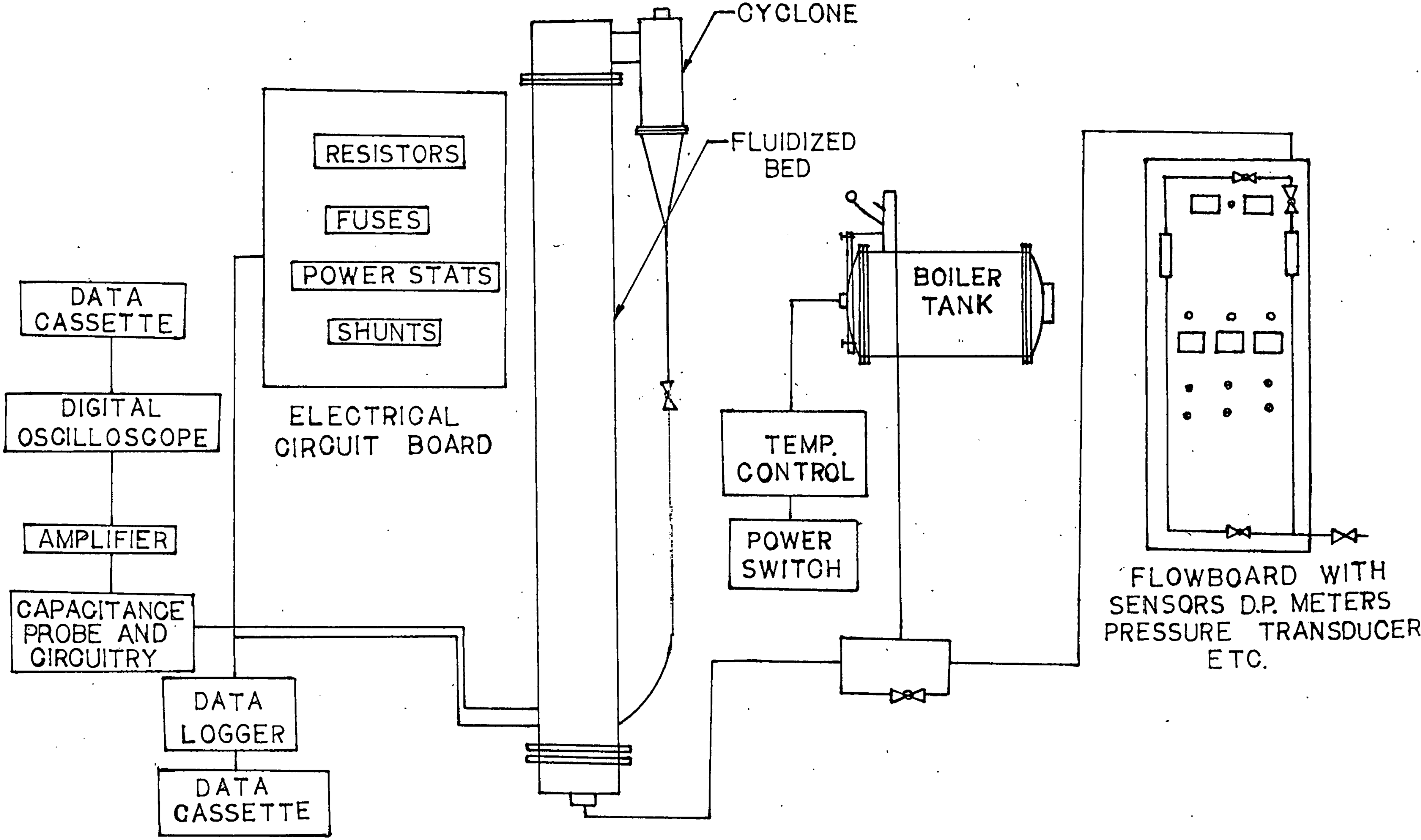


Figure 1 FLOW DIAGRAM OF THE FLUIDIZED BED TEST FACILITY AT LEHIGH UNIVERSITY

are given in references [2-4]. The data are mostly at low gas flow rates. Fluidized bed combustion operation calls for high gas flow rates. There are not enough data on heat transfer behavior of tubes in submerged region for high gas flow rates to compare with the anticipated results of this project in freeboard region at high flow rates. For this reason, the first tests were carried out with a submerged horizontal tube at midpoint of a 18 inches high static bed at room temperature. The test particles were 300 μm and 850 μm mean diameter glass beads. The maximum superficial gas velocity, $U_{sg}$, for 300 μm mean diameter particles was 2.5 m/sec and for 850 μm mean diameter particles was 3.5 m/sec.

At these high gas velocities, local and circumferentially averaged heat transfer coefficients were measured. They are tabulated in Table 1, for 300 μm mean diameter particles and in Table 2 for 850 μm mean diameter particles. To better visualize the variation heat transfer coefficient as a function of superficial gas velocity, they are plotted in Figures 2-6. Figure 2 shows the variation of average heat transfer coefficient for the two particles. The variation of local heat transfer coefficient around the tube at different gas flow velocities are shown in Figures 3, 4, 5 for the two particles.

TABLE 1

Heat Transfer Data for Glass Beads
with 300 μm Mean Diameter at Room Temperature

| $U_{sg}$, $\frac{m}{sec}$ | h, $w/m^2 {}^\circ C$ @ Angle β | | | | | | | | $\bar{h}$, $\frac{W}{m^2 {}^\circ C}$ |
|---|---|---|---|---|---|---|---|---|---|
| | 0° | 45° | 90° | 135° | 180° | 225° | 270° | 315° | |
| 0.385 | 206.5 | 135.6 | 236.2 | 233.0 | 222.7 | 216.6 | 202.7 | 182.5 | 204.5 |
| 0.542 | 465.3 | 236.6 | 282.3 | 308.2 | 318.1 | 217.6 | 212.5 | 266.8 | 296.4 |
| 0.692 | 424.5 | 242.3 | 271.3 | 267.6 | 284.9 | 245.5 | 231.1 | 263.5 | 278.9 |
| 0.821 | 373.5 | 333.3 | 271.7 | 288.0 | 325.6 | 241.8 | 226.9 | 257.1 | 270.2 |
| 0.946 | 427.0 | 232.7 | 229.6 | 285.4 | 286.0 | 243.8 | 193.5 | 255.7 | 269.3 |
| 1.107 | 442.2 | 229.6 | 220.1 | 274.2 | 284.5 | 240.6 | 187.0 | 252.6 | 265.5 |
| 1.238 | 440.8 | 234.4 | 221.1 | 271.4 | 283.1 | 245.0 | 195.8 | 260.2 | 269.0 |
| 1.345 | 442.8 | 233.7 | 219.6 | 292.2 | 299.5 | 254.0 | 189.3 | 261.0 | 274.0 |
| 1.577 | 436.3 | 223.3 | 205.7 | 276.9 | 301.1 | 242.4 | 184.2 | 257.9 | 266.0 |
| 1.803 | 456.4 | 232.7 | 222.3 | 280.3 | 306.4 | 231.7 | 183.4 | 263.3 | 272.0 |
| 1.967 | 459.1 | 239.6 | 209.3 | 288.0 | 340.3 | 261.5 | 189.0 | 268.0 | 281.9 |
| 2.367 | 460.1 | 243.5 | 212.6 | 263.4 | 305.9 | 231.9 | 183.6 | 264.9 | 269.8 |
| 2.585 | 460.1 | 229.7 | 199.6 | 250.2 | 315.0 | 230.7 | 170.9 | 245.7 | 262.7 |

TABLE 2

Heat Transfer Data for Glass Beads
with 850 μm Mean Diameter at Room Temperature

| $U_{sg}$, $\frac{m}{sec}$ | h, w/m² @ Angle β | | | | | | | | $\bar{h}$, $\frac{w}{m^2 °C}$ |
|---|---|---|---|---|---|---|---|---|---|
| | 0° | 45° | 90° | 135° | 180° | 225° | 270° | 315° | |
| 0.903 | 272.2 | 162.9 | 144.0 | 168.2 | 115.9 | 158.5 | 113.1 | 174.1 | 163.7 |
| 1.0075 | 250.6 | 146.0 | 162.2 | 125.1 | 135.8 | 118.2 | 129.0 | 156.1 | 152.9 |
| 1.355 | 266.0 | 145.5 | 154.1 | 133.0 | 148.7 | 125.5 | 127.0 | 158.5 | 157.3 |
| 2.255 | 250.5 | 134.4 | 141.6 | 139.6 | 163.9 | 132.7 | 116.2 | 151.2 | 153.8 |
| 2.369 | 215.6 | 106.5 | 114.4 | 135.7 | 197.5 | 128.1 | 93.6 | 111.3 | 137.8 |
| 2.678 | 228.6 | 114.7 | 120.3 | 136.3 | 197.2 | 128.9 | 80.5 | 112.4 | 139.9 |
| 2.891 | 237.4 | 120.4 | 127.4 | 131.2 | 180.6 | 123.5 | 95.6 | 114.7 | 141.4 |
| 2.914 | 240.8 | 117.3 | 139.9 | 132.9 | 184.4 | 126.9 | 103.3 | 114.7 | 145.0 |
| 3.169 | 243.3 | 115.4 | 133.6 | 139.4 | 176.1 | 128.4 | 100.4 | 106.4 | 142.9 |
| 3.468 | 240.3 | 114.9 | 123.4 | 150.7 | 185.7 | 138.8 | 120.2 | 115.7 | 148.7 |

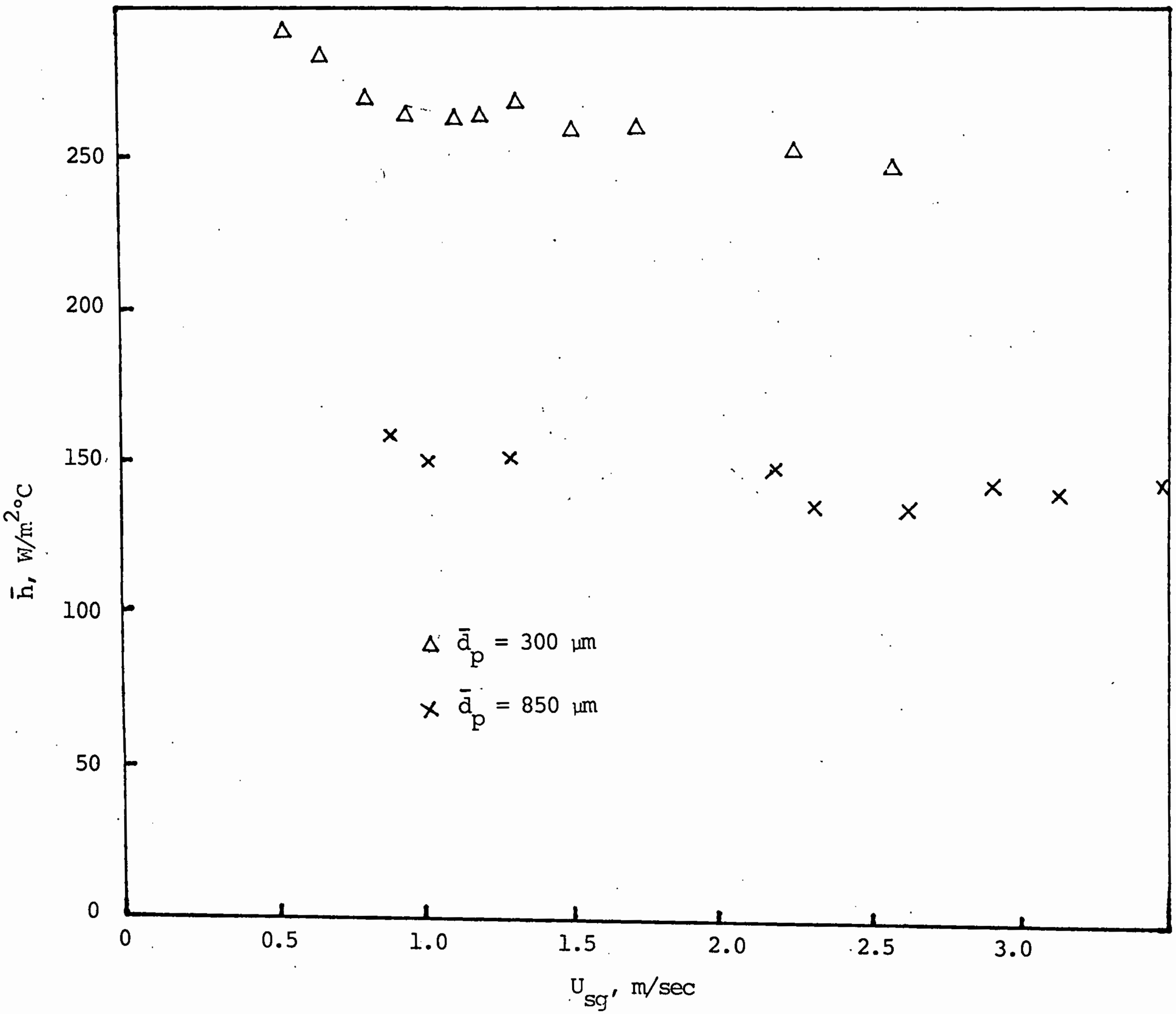


Figure 2 Variation of average heat transfer coefficient with superficial gas velocity

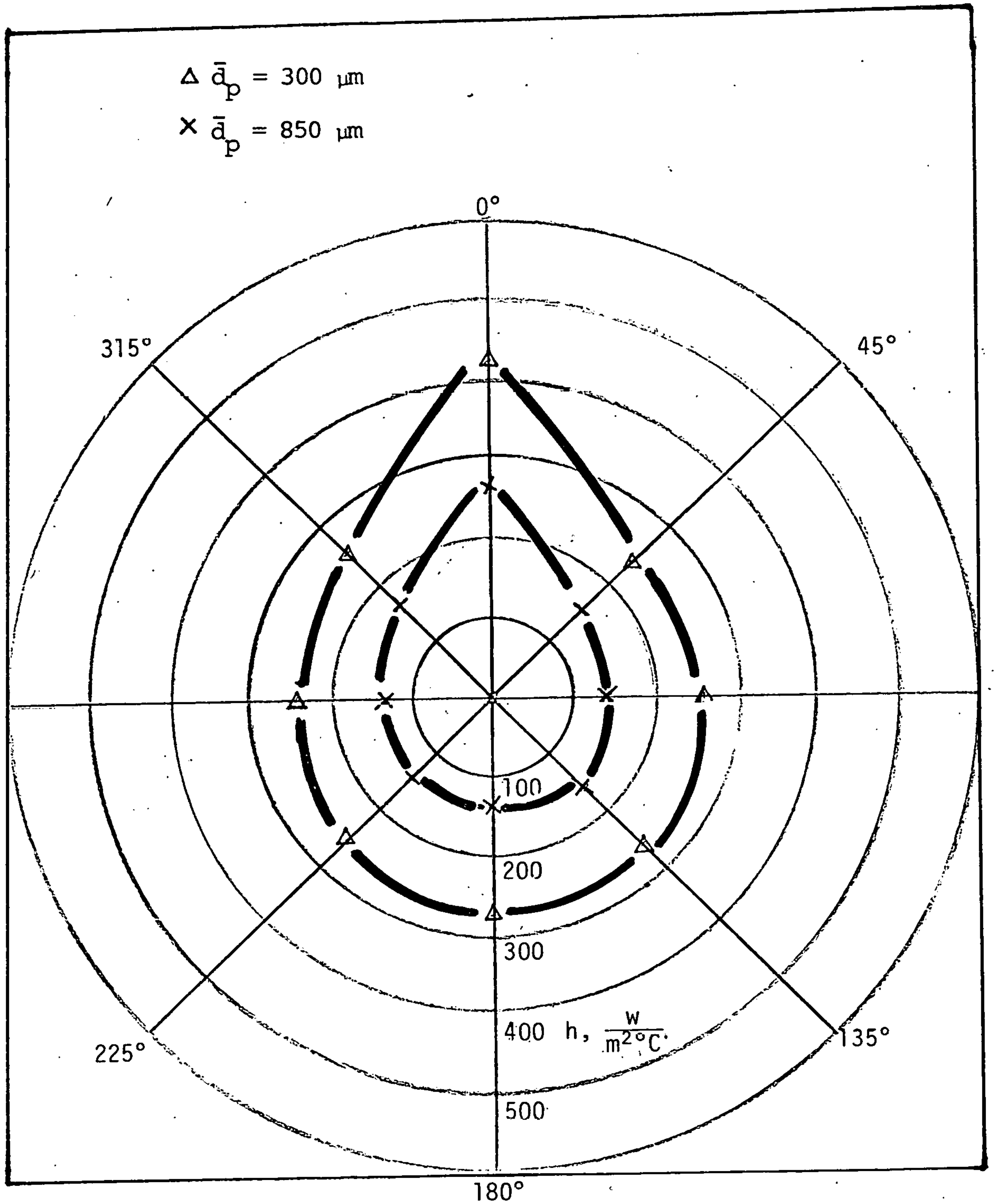


Figure 3 Local heat transfer coefficient at $U_{sg}$ = 0.7 m/sec around the horizontal tube, $0^o$ is the downstream direction

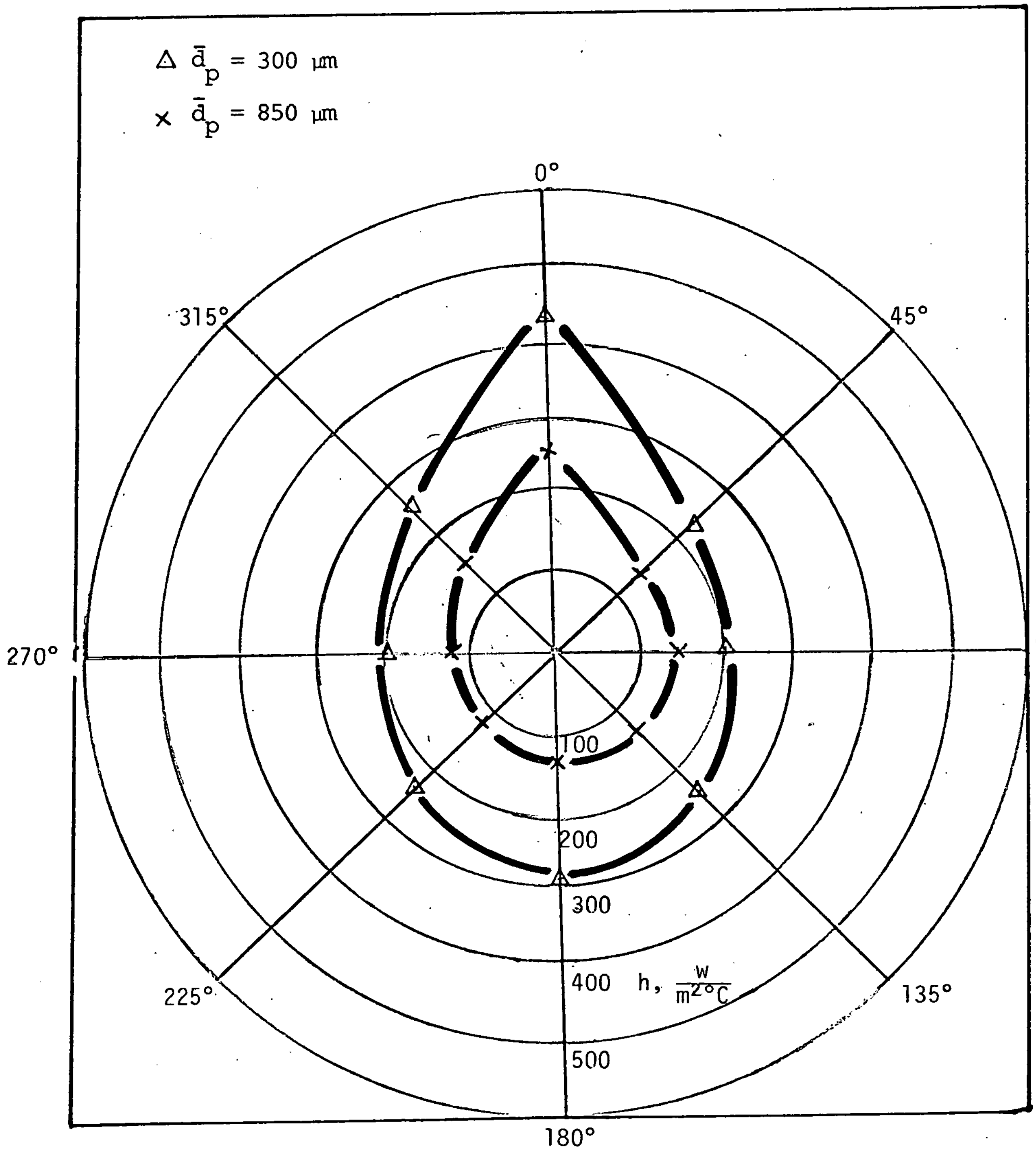


Figure 4 Local heat transfer coefficient at $U_{sg} = 1.5$ m/sec around the horizontal tube, $0^\circ$ is the downstream direction

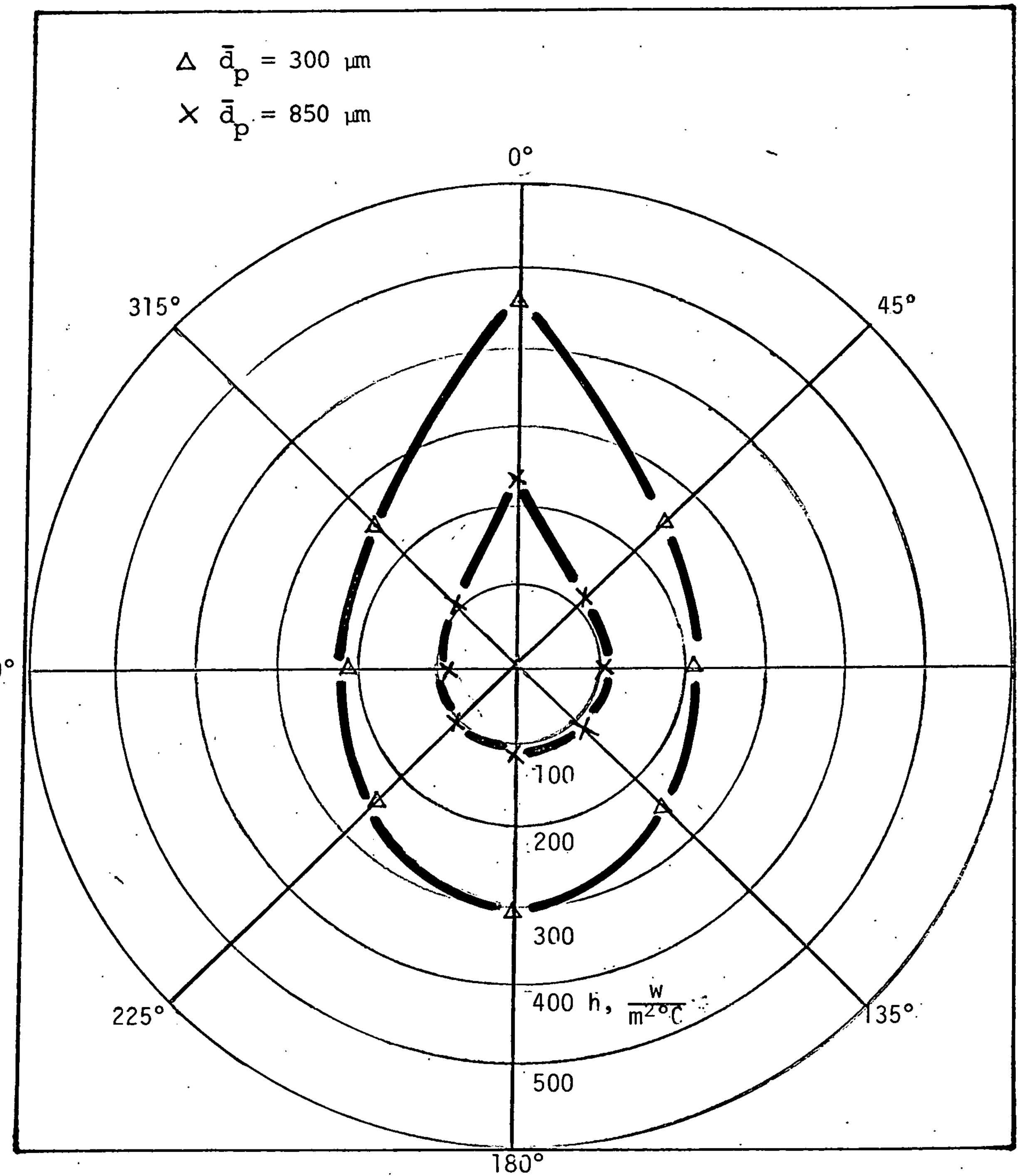


Figure 5 Local heat transfer coefficient at $U_{sg} = 2.5$ m/sec around the horizontal tube, 0° is the downstream direction

REFERENCES


1. R. T. Wood, M. Kuwata, and R. W. Staub, Heat Transfer to Horizontal Tube Banks in the Splash Zone of a Fluidized Bed of Large Particles, Proceedings of the 1980 International Fluidization Conference.

2. J. C. Chen, Heat Transfer to Tubes in Fluidized Beds, ASME Paper, 76-HT-75, National Heat Transfer Conference, St. Louis, August (1976).

3. F. W. Staub, Solids Circulation in Turbulent Fluidized Beds and Heat Transfer to Immersed Tube Banks, Journal of Heat Transfer v. 101, August (1979).

4. R. Chandran, J. C. Chen, F. W. Staub, Local Heat Transfer Coefficients Around Horizontal Tubes, Journal of Heat Transfer, vol. 102, February (1980).