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INDUSTRIAL APPLICATION FLUIDIZED BED COMBUSTION

CATEGORY III

INDIRECT FIRED HEATERS

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ABSTRACT

A program was initiated during July 1976 to evaluate the technical and economic potential for the application of fluidized bed combustion technology to refinery and petrochemical plant indirect fired process heaters. The strategy for the program is to build on the available boiler oriented FBC technology. Areas common to both steam generating boilers and process heaters will not be intentionally advanced by this program. However, the results of complimentary programs in the boiler area will be considered in the assessment of potential heater applications.

Two pertinent areas that are not being addressed in the on-going boiler oriented programs and which are being investigated here concern the effects of larger tube size and hydrocarbon coking. Phase I of the program consists of the design, construction and operation of three laboratory facilities to carry out these studies. Fluidized bed performance studies have been completed on tube bundle arrays of 2-inch and 6-inch diameter tubes and with tubes on nominal 2-diameter, 3-diameter and 4-diameter horizontal spacing. Conductive/convective heat transfer coefficients as a function of tube size, location and surface orientation have also been obtained on these same bundle configurations and on isolated single tubes.

A Process Stream Coking Test Unit has been assembled to study the primary process parameters affecting coke laydown on the internal surfaces of hydrocarbon containing tubes. Testing will begin during the next reporting Quarter.

Design and preliminary planning is complete for the third laboratory facility which will be a coal fired fluidized bed combustor. This facility will be used to study overall heat transfer coefficients and combustion performance.

1. Objectives and Scope of Work

The purpose of this program is to extend the state-of-the-art of fluidized bed coal combustion, which at present, addresses the generation of steam to applications where oil passing through immersed tubes in the bed will receive heat and be heated to a required condition. This purpose will be achieved by the successful completion of the following program objectives:

a. To conduct an R&D program necessary to provide the engineering data and know-how for designing a fluidized bed process heater.

b. To conduct an economic analysis necessary to evaluate the economic attractiveness of fluidized bed combustion for indirect fired process heater applications.

c. To demonstrate the operation of a coal fired fluidized bed heater in an actual refinery environment for an extended period of time.

d. To prepare a complete Design Specification and Control Cost Estimate for a commercial sized fluidized bed coal fired process heater.

The basic approach to be followed in pursuing the objectives of this program will be to build on the fluidized bed technology that is now available and under development by others in the related area of fluidized bed boiler applications. Effort in this program will be concentrated on doing the incremental work necessary to extrapolate the boiler oriented technology to refinery and petrochemical plant type indirect fired process heaters. The areas of technology common to both steam generating boilers and process heaters will not intentionally be advanced by this program. However, the state-of-the-art and the results of complimentary programs in the boiler area will be used in the overall technical and economic assessment of potential fluidized bed process heater applications.

The two principle areas of technology that have been identified as being peculiar to process heater applications and which are not being addressed in the on-going boiler orientated programs concern the effects of tube size and hydrocarbon coking. These two areas will be investigated in this program.

Indirect fired process heater tubes are conventionally two to five times larger in diameter than boiler tubes. A typical crude oil heater, for example, may have a multitude of 4" to 8" diameter tubes in the heat pick-up zones as contrasted to the 1" to 2" diameter tubes normally used in steam boilers. The effect that these larger tubes will have on fluidization characteristics and definition of the optimum or acceptable configuration of a tube bundle immersed within a fluidized bed must be investigated.

Similarly, the parameters affecting hydrocarbon coking must be investigated. When heating a hydrocarbon to 600°F+ (as required for separation by distillation or other typical processes) some degradation of the oil and coke laydown on the inside tube wall is unavoidable. The rate of coke laydown is affected primarily by the temperature of the hydrocarbon film on the inside wall of the tubes. This film temperature, in turn, is a function of several parameters relating inside film coefficient and heat transfer rate. Both overall average and localized conditions within the heat transfer zone must be examined.

The effects of tube size and coking described above will be investigated during the initial laboratory R&D phase of the program. This will be accomplished through the design, fabrication and operation of three separate laboratory test units. These units are designated as follows:

a. Two-Dimensional Flow Visualization Unit

b. Process Stream Coking Unit

c. High Temperature Heat Flux Unit

Other portions of the Phase I effort involve economic and operability evaluations of the technology and design of the Phase II Demonstration Unit followed by the Design Specification and Control Cost Estimate for a commercial-sized FBC process heater.

If, at the conclusion of Phase I, the technical and economic assessment of the data indicate favorable commercial potential, the program will be advanced to the demonstration phase. This will involve the installation of a 10-15 MBtu/Hr coal fired fluidized bed process heater in an Exxon refinery and its operation for a sufficient period of time to obtain the engineering data necessary to design a commercial-sized facility.

2. Summary of Progress to Date

The program got underway July 1, 1976 with an anticipated contract life of approximately 3 years. The program is structured into 10 Tasks or Cost Centers which are being used to monitor and report the progress of work. The overall schedule and identification of Tasks are shown in the Milestone Schedule Chart included here as Figure 1.

The program effort to date has been devoted to the design and construction of the three major laboratory units that will be used to generate most of the Phase I program data. The first of these units is the Two-Dimensional Flow Visualization Unit. This is an atmospheric pressure, transparent test chamber where fluidization and mixing characteristics of a fluidized bed containing immersed tubes can be visually observed and quantatively measured. A schematic of the facility is shown in Figure 2. The unit construction was completed and testing commenced in June 1977. To date four different bundle configurations have been tested and evaluated. The first consisted of 2-inch diameter tubes on 3-diameter center-to-center spacing. This configuration was identical to the bundle array installed in the Rivesville demonstration boiler and was intended to establish a baseline of performance against which comparisons could be made with data as it becomes available from that unit.

These tests were followed in turn by bundle configurations which were designed to evaluate the effects on fluidization performance of varying tube size and spacing. The additional bundles tested to date were all 6-inch diameter[1] tubes but arranged on 2-diameter, 3-diameter and 4-diameter center-to-center spacing and all using a scaleup of the Rivesville vertical and diagonal proportional spacing.

In conjunction with the fluidization performance testing, some conductive/convective heat transfer data were obtained on each of the bundle configurations. Variations in heat transfer coefficients as a function of peripheral tube surface orientation and tube location in the bundle have been determined. In addition, heat transfer data on single isolated tubes immersed in a fluidized bed have been measured using a range of tube sizes and bed materials to determine what affects the presence of adjacent tubes and varying bed materials (particularly bed particle size) have on heat transfer characteristics. These data are particularly useful in comparing the program results with data reported by other investigators who predominantly used single tubes and/or relatively fine bed materials. These single tube data are reported in some detail later in this report.

[1] Nearest plexiglas equivalent to nominal 6-inch pipe size or 6-1/2 inch O.D.

The second laboratory unit is the Process Stream Coking Unit. This unit consists of four single tube heat exchangers, an electric heat source and a gas fired crude preheater. A simplified flow plan is included as Figure 3. The objectives of these studies are to determine what effect the high heat flux levels available in a fluidized bed combustor will have on the coking rate of a hydrocarbon process stream and if these coking rates can be controlled within an acceptable range of operations. Comparative rates of coking at different heat flux levels, fluid mass velocities and bulk temperatures will be determined.

Assembly and transfer of this unit to its test site at Exxon's Bayway Refinery, Linden, N.J. has been completed. Final prestartup checkout of the unit and tie-ins to the Refinery crude feed supply and supporting utilities are nearing completion and testing will get underway during the 4th Quarter of 1977.

The third laboratory unit will be the High Temperature Heat Flux Unit in which maldistribution of heat input to a bundle of tubes immersed in a "hot" fluidized bed will be examined. Both peripheral and tube-to-tube heat flux patterns will be studied. The design of the unit has been completed and approved by the DOE. Vendor proposals have been received for all major equipment components and it is anticipated that orders will be released in the near future.

3. Discussion of Technical Progress

3.1 Two-Dimensional Flow Visualization Studies

3.1.1 Background Information

The Flow Visualization studies are being carried out in a two-dimensional atmospheric pressure, transparent fluidized bed chamber. The unit is approximately 1 ft. in depth by 7.5 ft. wide by 12 ft. high (see Figure 2). The facility is designed to accommodate a range of tube bundles assembled from tubes up to 6 inches in diameter and arranged on spacing up to 4 tube diameters on center.

Tube bundles will be immersed in the bed and the effects on fluidization of these relatively large tubes will be determined through a systematic study of the parameters of tube diameter, tube-to-tube spacing, tube-to-grid spacing and tube orientation. Other variables such as bed particle size, fluidization velocity, grid design and bed pressure drop will also be examined although these will be of secondary interest since they are being investigated by other boiler oriented programs.

A discussion of the overall Test Plan for this subtask including a description of the facility and the planned test sequence was included in the Program Quarterly Technical Report No. 1 dated October 19, 1976. The interested reader is referred to that report for more detailed background information.

3.1.2 Status of Work

A total of four different bundle configurations have been tested to date. The first bundle was the "Rivesville" configuration, so called because it duplicated the tube size and geometric arrangement installed at the Rivesville FBC boiler. This configuration was included in the Program test matrix, even though the primary program interest is in larger diameter tubes, in order to serve as a base for making subjective and comparative evaluation of later test results.

Subsequent tests have been run on three bundle configurations which were designed to evaluate the effects on fluidization performance of varying tube size and spacing. These bundles were each made up of nominal 6-inch diameter tubes but arranged on 2-diameter, 3-diameter and 4-diameter center-to-center tube spacing. Each used a direct scaleup of the Rivesville vertical and diagonal proportional spacing.

3.1.3 Test Results and Observations

The first step in the test sequence for each bundle was to determine bed fluidization characteristics over the full range of fluidization velocities. The bed material initially charged for each test was a 50-50 wt.% mixture of two grades of locally obtained limestone. Figure 4 indicates the resulting particle size distribution at the start of testing and at the conclusion of the test sequence. It can be seen that attrition during the operation of the unit generated some fines that were retained in the bed. Of course, some additional makeup material (the same 50-50 wt.% blend) was added periodically to replace the very fine material that was generated and discarded to the bag filter. But these data do represent the range of particle size distribution present in the bed during the course of the tests.

With this size distribution, observations were made to define the minimum (u_{mf}) and maximum superficial fluidization velocities. The pressure drop through the bed as a function of velocity for the Rivesville bundle test sequence are plotted on Figure 5 from which u_{mf} = 2.4 ft/sec can be determined. This is defined as the intersection of the rising pressure drop curve through the slumped bed and the essentially constant bed ΔP after fluidization is attained. This somewhat arbitrary definition of u_{mf} tends to wash out the data noise in the region of incipient fluidization which is probably caused by the wide range of particles present in the bed.

It must also be pointed out that when the same procedure was followed on subsequent tests with blended limestone bed material the results were somewhat inconsistent. Observed minimum fluidization velocities ranged from 1.6 to 2.4 ft./sec. It is hypothesized that the variations were caused by difference in bed packing, segregation and/or attrition. Some additional test work is planned to better define the effects of these variables.

The upper value of u was also not well defined but it was observed that above about u = 15-16 ft/sec the bed material tended to be carried out of the tube bundle and accumulate in a zone just above the top row of tubes. This would represent an upper practical limit of operation for this bed material and tube configuration.

The general observation was made that at low fluidization velocities (below about u = 5 ft/sec) the bed tended to segregate with the finer particles migrating to the top of the bed and the larger particles concentrating in the bottom tube rows and below the bundle. This observation could be significant in view of the effect that particle size is expected to have on overall heat transfer rates. It could represent another factor effecting maldistribution of heat input to tubes in different locations in a stacked horizontal tube bundle.

The general impression from observing the bed performance through the transparent walls was that "stagnation caps" tended to form on top of the tubes under essentially all fluidizing conditions. However, from close observations made by use of a boroscope through the inside wall of the tubes it was determined that these "stagnation caps" were not really stagnant but were very mobile and in almost constant motion. Individual particles generally had a very short residence time on the tube surface and constantly were being "washed off" and replaced with new particles.

Another step in the testing sequence was to determine the vertical and horizontal mixing characteristics of the bed. This information obtained on the Rivesville configuration will be used to establish a basis on which to make comparative mixing evaluations in later tests on bundles of larger diameter and more tightly packed tubes. In other words, we will be able to determine that mixing and particle migration characteristics of other configurations is either better than or poorer than available commercial experience and therefore would likely require more or fewer fuel injection points to maintain reasonable bed temperature distribution.

The mixing data are being obtained by injecting a sample of color tagged limestone particles into one side of the bed and, with stop action photography, following the migration of the particles as they mix both vertically and horizontally throughout the bed.

The primary significance of these data will only be apparent after additional data are available from all other tube bundles. However, to illustrate the type of data obtained, Figure 6 is included for reference. These data, which plot tracer concentration as a function of elapsed time after injection, were obtained during a test at 11.0 ft/sec fluidization velocity and at camera position #2 which is at the mid point in the bed and below the tube bundle. Similar data are obtained for each test at the 5 other indicated camera positions and at nominal fluidization velocities of 4 and 8 ft/sec.

These data indicate about a six second delay after injection until particles have reached the horizontal mid point in the bed and more than 60 seconds is required to reach equilibrium concentration. A more general observation from an analysis of all the data accumulated to date is that vertical mixing in the bed is very rapid but that horizontal mixing is relatively slow.

3.1.4 Planned Work

As indicated previously, a total of four tube bundle configurations have been tested. In addition to the Rivesville bundle of 2-inch diameter tubes, other bundles of tubes made up of 6-inch diameter tubes on 2-diameter, 3-diameter and 4-diameter tube center-to-center spacing have been evaluated. Remaining configurations to be tested include a 6-inch tube bundle on equilateral triangular pitch and finally a 4-inch diameter tube bundle with final spacing and pitch to be determined from an evaluation of optimum performance characteristics of the 6-inch tube configurations.

The analysis of fluidization performance of all the bundle testing will be included in a subsequent Quarterly Technical Report at the conclusion of this part of the program.

3.2 Process Stream Coking Studies

3.2.1 Background Information

The Process Stream Coking Studies are designed to determine what effect the high heat flux rates available in a fluidized bed combustor will have on the coking rate of a hydrocarbon stream and if these coking rates can be controlled within an acceptable range of operations. More specifically, they will establish a relative rate of carbon or coke deposition on the inside wall of a hydrocarbon containing tube as a function of bulk temperature, heat flux rate, mass velocity and inside film temperature.

The test facility that has been designed to carry out these studies will consist of four single tube heat exchangers, an electric heat source and a gas fired feed preheater. The basic scheme will be to pass a stream of hydrocarbon feed through each of the four exchangers. Each exchanger will be exposed to a different combination of process conditions (mass flow, bulk temperature and heat flux rate) and each will be carefully monitored for indications of coke deposition on the inside surface of the exchanger tube. In this way, comparative coking rates as a function of the varying process parameters can be determined.

A simplified process flow plan for this test facility is shown in Figure 3. A detailed description of the facility including a discussion of the planned test matrix and basis to be used for analysis of data is given in the Quarterly Technical Report No. 2 dated January 26, 1977. The reader is referred to that report for more detailed background information.

3.2.2 Status and Future Work

The Process Stream Coking Test unit was built on nine pallets in the ER&E Mechanical Division shops at Linden, N.J. Assembly was completed during this reporting period and the unit was operated with water as the process fluid. This technique permitted a complete simulated operation of the unit and checkout of all instrumentation and controls prior to moving into the Refinery.

On August 8th, the unit was transferred to its permanent test site within the Bayway Refinery. The transfer was accomplished without incident or damage to any of the equipment.

Since this test unit is situated within an operating refinery and will be an integral part of the processing sequence of a major refinery unit, special precautions and preparations were necessary to integrate the test unit into the refinery operating procedures.

While the test unit is basically designed to operate without an operator in attendance it will be necessary for the regular operating personnel to make periodic security checks of the unit and to react to any unforeseen occurrences or emergencies. A series of training sessions have been conducted for all responsible operating personnel. These sessions have included verbal instructions on test procedures and equipment plus some "hands-on" drills in operating the data logger and control instrumentation.

The third and final safety review with the Refinery Safe Operations Committee was also held at which time full agreement and approval was reached on the interface of responsibilities and procedures to be followed during the test period by both the Refinery and the ER&E personnel. An operating manual has also been prepared for the unit.

During the 4th Quarter of 1977 tie-ins to the refinery crude oil and support utilities will be completed and testing will get underway. A 28 to 32 week test period is anticipated to complete the planned program matrix.

3.3 Fluidized Bed Heat Transfer Studies

3.3.1 Background Information

The objective of the Fluidized Bed Heat Flux Studies is to quantitatively define both the peripheral and the tube-to-tube maldistributions of heat input to tubes immersed in a fluidized bed. The maldistribution patterns will be determined as a function of controllable design parameters including tube size, spacing, orientation and fluidization velocity.

The data to satisfy the requirements of this task will be obtained in two separate series of tests. The principle tests will be carried out in a "hot" fluidized bed facility. These tests will determine the overall level and pattern of heat transfer to tubes in a fluidized bed. Some complimentary ambient temperature studies, which are already underway will define the conductive/convective component of the heat transfer mechanism. By comparing results from the high temperature and ambient tests the radiation component can be determined by difference.

A detailed discussion of the facility designs and Task Plan for this part of the Program is given in the Quarterly Technical Report No. 3, dated April 25, 1977. The interested reader is referred to that report for additional information.

3.3.2 High Temperature Heat Flux Tests

A conceptual design of the Heat Flux Test Unit has been completed and approved by the DOE. Competitive vendor proposals have been received for the major equipment components. It is anticipated that procurement and construction activities can proceed during the 4th Quarter of 1977.

3.3.3 Ambient Temperature Heat Flux Tests

The primary facility being used for the ambient temperature heat flux tests is the Two-Dimensional Flow Visualization test unit that is described in Section 3.1 of this report. In this unit conductive/convective heat transfer data on both isolated single tubes and on multiple tubes arranged in various bundle configurations are being obtained. These tests all use a relatively coarse limestone bed material similar to that anticipated for use in later commercial installations.

Some additional data on the performance of small single tubes immersed in beds of 390μ and 1000μ spherical glass beads and 200-4000μ blended limestone were obtained in an 8-inch diameter test column. It is anticipated that by comparing the heat transfer data obtained in these three modes, i.e.

a. single tubes with fine glass beads.

b. single tubes with coarse limestone.

c. tube bundles with coarse limestone.

some conclusions can be drawn concerning the validity and applicability of heat transfer data reported by other investigators to the specific objectives of this program.

All of the single tube tests planned within the scope of this program have been completed and are reported in the following paragraphs. The testing and analysis of the various bundle configurations is continuing and will be covered in a subsequent report when these tests are completed.

3.3.3.1 Test Procedure

A detailed description of the test procedures, and particularly the test probe being used to obtain the heat transfer data was given in the Quarterly Technical Report No. 4 dated July 29, 1977. Briefly summarized, the procedure is as follows:

The probes consist of specially instrumented plexiglas tubes installed in the fluidized bed zone of the test unit. Each tube has a 1/4" x 6" x .005" thick Nichrome strip imbedded flush with the outside tube surface. One or two 40 BWG iron-constantan loop junction thermocouples are attached to the under surface of the strip to monitor strip temperature. (See Figure 7 for assembly details).

The Nichrome strip which has a resistance of approximately 0.2 ohms, is electric resistance heated to a temperature of 30° to 60° F above the ambient bed temperature. The power input required to maintain this differential temperature between the strip and the bed is monitored and is a direct function of the mean conductive/convective heat transfer coefficient over the area of the strip. By comparing the relative power required to maintain the temperature of the strip at various locations around the circumference of the tube and from one tube location to another a pattern of maldistribution as a function of tube location and surface orientation can be obtained.

A system is used to index or rotate the probes with the bed in operation so that a complete set of data at any number of desired locations around the tube circumference can be obtained using a single heated strip. This system has the advantage of greatly simplifying the construction of the probes and assuring that all data points for a given set will be directly comparable with each other since all data are obtained from the same test strip.

Probe Calibration

In order to compensate for the thermal losses other than those directly from the strip surface, each heater assembly is calibrated prior to use. The calibration is performed by placing the heated probe in a metered air stream and/or water where the heat transfer coefficient can be calculated with a reasonable degree of accuracy. By comparing this coefficient with that back calculated from the measured power input, a correction factor for each strip is generated. For imposed heat transfer coefficients greater than 30 Btu/hr ft^2 °F, the calibration experiments plus a computer modeling of the probe design indicated a probe error of less than 15%. The higher the imposed heat transfer coefficient, the better the probe accuracy. At an imposed coefficient of 100 Btu/hr ft^2 °F, probe error was less than 5%. Probe-to-probe correction factors varied only by 2 to 3%. In all calibration runs, the measured coefficients were higher than those predicted. All heat transfer coefficients reported here are as measured with no corrections applied since the coefficients generally are in the 30 Btu/hr ft^2 °F and higher range where probe accuracy is within other experimental variables.

Single Tube Tests Run

The following single tube tests were run in either the 8-inch diameter column or the Two-Dimensional Flow Visualization Unit:

Tests run in 8-inch column:

Test No. 1 - Two-inch diameter tube; 390μ spherical glass bead bed material. Fluidization velocities u = .34 to 1.6 ft/sec.

No. 2 - Two-inch diameter tube; 1000μ spherical glass beads; u = 1.8 to 4.0 ft/sec.

No. 3 - Two-inch diameter tube; 200μ-4000μ limestone (1000μ wt. avg. size); u = 2.1 - 8.4 ft/sec.

No. 4 - Two-inch diameter tube; 200μ-1800μ "ferilizer filler" graded limestone (650μ wt. avg. size); u = 2.2 - 3.6 ft/sec.

No. 5 - Two-inch diameter tube; 1000μ-4000μ "moon mountain grit" graded limestone (2500μ wt. avg. size); u = 6.3-8.0 ft/sec.

Test runs in Two-Dimensional Flow Visualization Unit:

No. 6 - Two-inch diameter tube; 200μ-400μ limestone (1000μ wt. avg. size); u = 11.-15.8 ft/sec.

No. 7 - Four-inch[(1)] diameter tube; 200μ-4000μ limestone; u = 10.5-14.6 ft/sec.

No. 8 - Six-inch[(1)] diameter tube; 200μ-400μ limestone; u = 11.4-15.0 ft/sec.

No. 9 - Six-inch[(1)] diameter tube; 200@-4000μ limestone (1000μ wt. avg. size); Bed depths of 15-inch and 21-inch above grid level. u = 11.4 ft. sec.

The reduced data displayed in the form of local heat transfer coefficients for these nine runs are tabled in Appendix I.

By analyzing the results from tests in the above matrix, various conclusions and observations can be made concerning the effects of fluidization velocity, bed particle size, tube diameter and bed depth. Each of these parameters will be discussed separately in the following paragraphs.

System Limitations

All single tube tests were run over a range of fluidization velocities - the range for each test varying depending on the bed particle size and the physical limitations of the test facilities being used. For example, in the 8-inch column tests with glass beads, maximum fluidization velocity was limited only by the entrainment velocity of the particles being used. With limestone, on the other hand, a system pressure head limitation restricted the region of investigation to a maximum of 8 ft/sec or about 3 times u_{mf}.

(1) 4-inch and 6-inch nearest plexiglas equivalent to nominal pipe sizes or 4-1/2 inch and 6-1/2 inch O.D. respectively.

When testing single tubes in the Two-Dimensional Flow Visualization Unit it was not possible to operate the unit at low fluidization velocities in the range from 2.4 ft/sec (u_{mf}) up to about 8 ft/sec. In this operating range the entire bed tended to alternately lift and collapse as a single mass, putting very high oscillatory forces on the plexiglas walls of the unit. Operating in this mode was judged to be inappropriate for the safety of this equipment.

It is interesting to note that when the unit is operated under the same conditions but with a tube bundle immersed in the bed, the same phenomenon is not observed.

3.3.3.2 Ambient Temperature Test Results

Visual Observations

During the course of running these tests in both the 8-inch column and the larger Two-Dimensional Flow Visualization Unit some visual observations were noted which may help to explain or interpret the data. These observations will therefore be described before discussing the test results.

Three distinct zones of solids flow or solids-to-tube contacting were observed around the periphery of the horizontal immersed tubes. Since there is general agreement among investigators that the relatively high heat transfer coefficients in a fluidized bed are a result of the scrubbing action or contacting of the solids particles on the tube wall, it would follow that these different patterns or zones of flow could be significant to our understanding of the heat transfer mechanism.

The first zone occurred on the lower portion of the tube - that is the side of the tube facing toward the direction of oncoming fluidization gas (identified as 180° position on all accompanying figures). This area has been described by others as a bubble shrouded zone; however, in this study, it did not appear as a complete gas pocket. The zone appeared to be lighter than normal bed density with very rapid density variation occurring. This zone can be contrasted to the second zone located on top of the tube (0°). In this second zone, the tube was covered by a dense "cap" of particles. The cap was not stagnant as oscillations of the cap occurred in a regular side to side sliding motion. In addition to the sliding, frequent bubbles crossed the upper surface causing solids replacement. This bubbling behavior regularly occurred only over the center two thirds of the tube, indicating a possible wall effect. This wall effect was also noted in the size of the cap. When the bed was defluidized and solids drained from the bed, the cap of material remaining on the heat transfer tube was observed to be saddle surfaced, the peak of the cap being two to three times higher nearer the wall than tube center. In the fluidized condition, it was not possible to measure the cap at the tube center. However, a qualitative judgement was made as to cap size by brightly back lighting the cap area and comparing light transmitted to the height of the cap at the wall. Again, the wall cap appeared taller than at the tube center. The location of the third zone varied with fluidization velocity, but was normally located at approximately the 90-45° position. This zone alternated between the dense cap and dilute phase, much like that of the advancing and retreating of ocean surf. As will be seen later, this "surf" consistently coincided with the area of highest heat transfer.

Average Heat Transfer vs. Fluidization Velocity and Particle Size

The heat transfer data as a function of fluidization velocity and particle size were analyzed on the basis of both overall average coefficients and the effects on peripheral maldistribution patterns. The overall average coefficients will be discussed first.

The average overall coefficients measured for the 390μ and 1000μ spherical glass beads and the 200-4000μ limestone blend are displayed on Figure 8. The data are plotted against a velocity parameter of

$$\frac{u}{u_{mf}}$$

which tends to normalize the data for the differences in fluidization characteristics imposed by differences in particle size. The u_{mf} used for the glass beads were calculated by the Leva method while a nominal u_{mf} of 2.4 ft/sec was used for the limestone blend.

At the higher end of the velocity ranges tested the limestone data compared very closely with the 1000μ glass bead data. It is interesting to note that the weight average particle size for the limestone blend was nearly 1000μ (see Figure 4 for the particle size history of the limestone bed material used in all tests).

At lower fluidization velocities, the resultant limestone heat transfer coefficients deviated substantially from the 1000μ line. At the lowest fluidization velocity at which data were collected, the limestone and 390μ glass beads had nearly identical measured heat transfer coefficients. Based on Leva's correlation for minimum fluidization velocity, much of the larger sized limestone would have segregated out at these low velocities with only particles of about 600μ or smaller still in a well fluidized state. Under these conditions the average size fluidized particle would have been 340μ. It is probable that multi-layer fluidization had occurred as described by Wen.[1] Wen's discussions center on a two-particle system; however, the same general trend should occur in multi-particle systems. Wen's theory predicts distinctly separated layers of fluidization when

$\frac{D_{P_1}}{D_{P_2}} > 1.3.$ In these tests the particle size range was significantly greater than 1.3.

It was observed that both the 390μ and 1000μ glass bead heat transfer coefficients increased rapidly with fluidization velocity while the limestone blend displayed much less sensitivity to fluidization velocity.

The coefficients measured for the glass spheres followed the inverse relationship with particle size of

$$h_{ave} \propto \left(\frac{1}{D_p}\right)^{0.36}$$

This relationship has also been reported by Zabrodsky.[2]

A few tests with 106μ glass spheres were also run which further confirmed this inverse relationship over the entire 10:1 particle size range.

Local Heat Transfer vs. Fluidization Velocity and Particle Size

The effect of fluidization velocity and bed particle size on local peripheral heat transfer can be seen from an examination of Figures 9a, 9b, and 9c. Figures 9a ans 9b are polar plots of peripheral heat transfer patterns for the 390μ and 1000μ glass beads respectively with velocity parameters of $\frac{u}{u_{mf}}$ in the range of 1.0 to 4.

An inspection of the data indicate that at relatively low fluidization velocities the highest rate of heat transfer occurs at the sides or 90°/270° positions on the tube where particle agitation is the most vigorous. The coefficient at the upper surface (0°) increases rapidly with an increase in fluidization velocity. This corresponds to the observed increase in mobility of the particles located in this "cap" region. As the fluidization velocity is increased the heat transfer profile becomes more circular and the cusps or flatness in the "cap" zone disappear. Very little change in heat transfer coefficient occurs at the bottom (180°) of the tube since this position was consistently contacted by dilute emulsion phase solids. The highest heat transfer rates at the intermediate and higher velocities were found approximately at the 45°/315° position with the exact maximum position depending on velocity. These high coefficients are consistent with the visual observation of the rapidly oscillating solids layer at the location of the previously described "surf line" and "cap". Rotation of the test probe into and out of the surf line region caused substantial and very rapid strip temperature variations at constant power input.

The local heat transfer data shown on Figure 9c were obtained in the same 8-inch column but using the limestone blend bed material. In this case runs were made at $\frac{u}{u_{mf}}$ ranging from .9 to 4.0. While the data are not conclusive, it would appear that at $\frac{u}{u_{mf}} \geq 1$ the maldistribution patterns are less sensitive to fluidization velocity for the limestone. This performance would be consistent with data discussed earlier for overall average coefficients and summarized on Figure 8.

Two additional brief tests were run in the 8-inch column to evaluate the effects of varying the particle size distribution of the limestone bed material. One test used a quarry graded "fertilizer filler" limestone with a size distribution of 200μ to 1800μ with a weight average size of 650μ. The second test used a "moon mountain grit" with a size distribution of 1000μ to 4000μ and a weight average size of 2500μ. (It is a 50-50 blend of these two grades of material that is used as the charge material for all tests run in the Two-Dimensional Flow Visualization Unit for both single tubes and bundle configuration tests).

The results of these two tests are compared on Figure 10 with the previously noted data on the limestone blend. The data appear to follow the same general pattern observed in the glass bead tests - namely that heat transfer coefficients increase with a decrease in weight average particle size. No more definitive conclusions are drawn from these limited data.

Heat Transfer Coefficients vs. Tube Diameter

The effects of tube diameter on overall heat transfer performance was also investigated. Since the 8-inch test column obviously could not accommodate tubes larger than 2-inch diameter these tests were carried out in the Two-Dimensional Flow Visualization Bed. Tests were run on single 2-inch, 4-inch and 6-inch diameter tubes. As can be seen on Figure 11, the 2-inch diameter tube data blended very well with the single overlapping data point obtained on the 8-inch column during the limestone bed tests on that unit.

In these tests heat transfer coefficients were measured at 15° increments around the tube circumference and averaged to obtain overall heat transfer data for each tube at each velocity.

The data show an increase in heat transfer coefficient with an increase in tube diameter. Dependence at constant fluidization velocity followed the relationship.

$$h_{ave} \propto (D_T)^{0.2}$$

These results are in general agreement with results reported by McLaren et al[3] and Kurochkin[4] but are not consistent with results reported by several other investigations.

Heat Transfer Coefficient vs. Bed Depth

One experiment was conducted to test the sensitivity of the heat transfer coefficient to bed depth. In this experiment a 6-inch diameter tube was positioned 18-inches above the grid while the defluidized bed depth was 15 inches in one run and 21 inches in the other. In other words, in the first case the slumped bed was even with the bottom tangent of the tube and in the second with the top tangent line. In both cases the tube was well inundated in the bed when fully fluidized.

The results of the heat transfer measurements are shown on Figure 12. The measured overall average coefficient was about 7% higher for the deeper bed. Also, there was a rather significant change in the profile or pattern of heat transfer. The coefficient at the 45°/315° "surf line" positions was measurably higher for the shallower bed. With the increase in bed depth, the region of highest heat transfer shifted to the top of the tube, indicating an increase of particle activity in the "cap" area.

3.3.4 Summary Observation and Conclusions

The following summary observations and conclusions can be made from an analysis of the single tube heat transfer tests reported here.

(1) Three distinct zones of particle activity or particle-to-tube contacting can be observed when a tube is immersed in a fluidized bed. These zones appear to bear a relationship to the heat transfer coefficient measured at each respective zone.

(2) The overall average heat transfer coefficient to a tube immersed in a fluidized bed appears to follow a relationship with particle size of $h_{ave} \alpha \left(\frac{1}{D_p}\right)^{.36}$

(3) For a bed containing a wide particle size distribution, the heat transfer coefficient is approximately governed by the weight average particle size of the portion of the bed that is in a well fluidized state.

(4) The heat transfer pattern in a bed composed of a range of particle sizes is less sensitive to fluidization velocity than a narrow graded size bed material.

(5) Patterns of peripheral heat transfer to an immersed tube become more symmetrical with an increase in fluidization velocity. The local coefficient at the bottom or 180° position on the tube is affected very little by changes in fluidization velocity.

(6) These experiments indicated a $h_{ave} \alpha (D_T)^{.2}$ relationship between heat transfer coefficient and tube diameter.

REFERENCES

(1) Wen, C. Y., Ya, Y. H. "Mechanics of Fluidization", Chemical Engineering Symposium Series, Vol. 62, 1966.

(2) Zabrodsky, S. S., Hydrodynamics and Heat Transfer in Fluidized Beds. M.I.T. Press, Cambridge, Mass. (1966).

(3) McLaren, J., Williams, D. F., Institute of Fuel, Vol. 42, Pg. 303 (1969).

(4) Kurochkin (as reported on: Modeling of a Fluidized Bed Combustor with Immersed Tubes. Dept. of Energy Engineering, University of Illinois, ERDA Quarterly Report Sept.-Nov., 1976. FE-1787-6).

APPENDIX I

Local Heat Transfer Measurements

a) Two-Inch Tube - Glass Beads

b) Two-Inch Tube - Limestone

c) Two-Four-Six-Inch Tube - Limestone

APPENDIX Ia

LOCAL HEAT TRANSFER COEFFICIENT MEASUREMENTS (BTU/HR FT^2 °F)

2" O.D. Tube

	390 μ Glass Beads Superficial Velocity (FPS)				1000 μ Glass Beads Superficial Velocity (FPS)		
(Deg.)	0.422	0.848	1.38	1.62	1.78	2.88	3.99
0	21.4	63.7	98.3	85.7	19.7	66.0	72.2
45	33.5	84.2	109.7	124.3	33.7	68.8	62.8
90	63.5	90.7	94.3	54.4	49.8	56.9	54.8
135	46.1	52.4	64.6	76.8	45.8	47.5	46.2
180	27.5	47.2	53.0	69.7	30.0	39.8	41.3
225	44.9	51.3	59.0	53.6	46.6	44.0	44.6
270	68.2	81.3	79.9	47.5	50.5	53.7	53.0
315	37.6	96.2	102.5	130.8	34.9	66.2	62.6
360	20.9	63.7	100.3	-	20.0	64.9	70.8

Measurements made in the 8" circular column slumped bed completely covers heat transfer tube.

APPENDIX Ib

LOCAL HEAT TRANSFER COEFFICIENT MEASUREMENTS (BTU/HR FT^2 °F)

2" O.D. Tube

	Fertilizer Filler Superficial Velocity (FPS)			Moon Mountain Grit Superficial Velocity (FPS)			Blend 50/50 Wt. Pct. Superficial Velocity (FPS)				
(Deg.)	2.23	2.73	3.64	6.33	6.89	7.94	2.11	2.81	4.14	6.42	8.39
0	31.4	57.3	77.6	25.3	19.8	53.3	25.8	35.8	75.9	84.3	79.6
45	54.8	76.3	76.1	55.6	53.9	52.3	31.4	49.4	61.7	68.6	61.6
90	50.0	48.3	48.3	55.5	57.1	49.1	50.6	54.2	54.4	55.5	52.6
135	43.2	42.1	40.1	46.6	50.9	41.7	39.3	44.3	45.7	44.8	47.3
180	32.6	31.5	34.2	41.8	41.9	33.7	32.9	36.6	35.5	38.2	38.9
225	40.6	39.3	38.5	49.5	46.7	40.1	37.4	44.8	48.3	41.9	43.9
270	49.6	48.5	45.2	56.2	53.0	46.4	53.4	53.3	52.5	48.8	52.7
315	55.2	71.0	70.6	53.3	56.9	51.1	43.9	59.6	55.5	66.4	58.5
360	29.7	56.8	-	23.9	20.0	52.7	-	35.6	-	83.5	72.9

	Particle Size Range	Wt. Avg. Size
Fertilizer Filler	650 μ	200-2500
Moon Mountain Grit	2500 μ	1800-4000
Blend 50/50 Wt. Pct.	1000 μ	200-4000

Measurements made in 8" circular column slumped bed completely covers heat transfer tube.

APPENDIX Ic

LOCAL HEAT TRANSFER MEASUREMENTS (BTU/HR FT^2 °F)

Bed Material: 50/50 Wt. Blend Fertilizer Filler/Moon Mountain Grit
Particle Size Range: 200-4000 μ
Wt. Average Size: 1000 μ

	2" O.D. Tube Superficial Velocity (FPS)			4" Nominal (4-1/2 inch O.D.) Superficial Velocity (FPS)			6" Nominal (6-1/2 inch O.D.) Superficial Velocity (FPS)		
Deg.	11.1	13.2	15.8	10.5	12.4	14.6	11.4*	11.4	15.0
0	71.8	63.1	65.2	104.0	84.1	73.0	69.2	101.7	54.4
15	68.1	67.2	66.8	98.5	80.7	72.3	68.5	98.7	73.2
30	64.3	62.9	64.6	97.9	86.1	80.2	84.6	89.9	93.1
45	61.1	57.9	59.5	91.5	94.5	91.2	95.4	92.2	103.8
60	56.1	55.0	58.1	75.7	72.2	82.1	85.7	78.9	94.4
75	52.3	52.5	52.6	55.0	56.8	59.8	65.2	59.3	63.9
90	49.7	50.2	51.2	52.9	48.9	51.7	51.3	53.1	51.5
105	48.6	49.1	50.9	46.0	47.7	58.1	47.2	50.2	48.2
120	48.0	48.9	50.6	45.6	44.3	47.0	47.7	47.5	48.2
135	48.1	48.6	49.7	44.5	45.1	45.8	47.6	47.6	47.0
150	47.3	48.0	49.6	43.8	44.7	45.6	44.7	47.2	46.0
165	47.0	49.0	49.7	42.7	44.1	44.6	44.2	47.1	43.8
180	47.7	47.3	49.9	42.7	42.3	44.1	43.0	46.2	42.0
195	47.8	48.9	49.3	43.2	41.5	44.3	42.0	47.4	44.0
210	46.6	49.5	51.6	44.7	42.2	43.4	45.0	47.9	45.2
225	48.9	49.3	52.1	44.4	43.4	44.3	45.6	49.5	46.8
240	49.8	51.4	53.5	44.3	45.7	45.5	46.4	51.3	46.8
255	49.0	50.8	54.0	43.9	46.4	45.3	49.7	52.2	47.9
270	49.8	50.3	53.3	45.5	45.0	47.1	49.2	51.9	45.9
285	50.5	51.0	53.0	55.7	50.0	53.1	65.0	60.1	57.7
300	55.9	54.7	53.9	68.4	64.3	67.8	86.0	74.5	81.5
315	56.4	53.5	54.2	89.1	73.8	80.6	97.6	90.1	94.7
330	64.3	55.1	58.6	94.9	83.0	88.0	85.2	91.3	96.5
345	65.7	59.1	61.2	98.9	91.5	82.9	73.9	96.4	76.5
360	65.3	60.6	63.9	97.9	90.1	75.8	66.4	-	56.0

Measurements performed in 2D Bed.
Tube to Grid Distance 18"
Slumped Limestone Bed Height 21"
* Slumped Limestone Bed Height 15"

FORM ERDA 535
(10 – 76)

U. S. ENERGY RESEARCH AND DEVELOPMENT ADMINISTRATION

MILESTONE PLAN AND MANAGEMENT REPORT

FORM APPROVED
OMB NO. 38R-

Page ___ of ___

1. Contract Identification: Industrial Application of Fluidized Bed Combustion - Category III Indirect Fired Heaters

2. Reporting Period: through Sept. 30, 1977

3. Contract Number: EX-76-C-01-2471

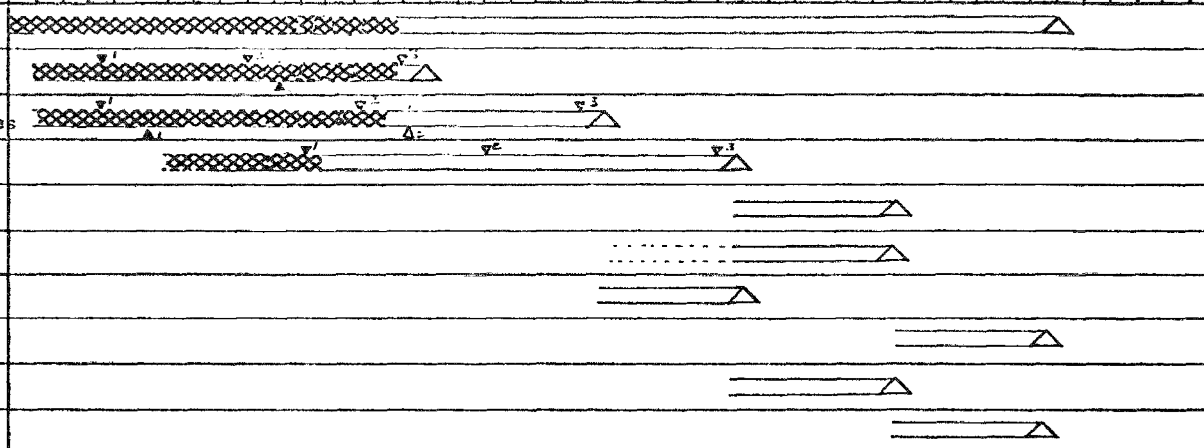
4. Contractor (name, address): Exxon Research and Engineering Company, P. O. Box 101, Florham Park, New Jersey 07932

5. Contract Start Date: June 30, 1976

6. Contract Completion Date: March 31, 1979

9. Fiscal Years and Months: T.Q.'77 (J A S); FY '77 (O N D J F M A M J J A S); FY '78 (O N D J F M A M J J A S); FY '79 (O N D J F M A M J J A S); FY '80 (O N D)

7. Identification Number	8. Reporting Category (e.g., contract line item or work breakdown structure element)	10. Percent Complete
1	Project Management	
2	2-Dimensional Flow	95
3	Process Stream Coking Studies	55
4	Fluidized Bed Flux Studies	8
5	Unit Operability Studies	
6	Econ. & Applications Eval.	
7	Demo. Unit Design	
8	Commercial Unit Design	
9	Demo. Unit Fab. & S/U	
10	Demo. Unit Operation	



11. Remarks
1. ERDA 90% design review
2. Construction complete
3. Testing complete

12. Signature of Contractor's Project Manager and Date: D. C. Cherrington [signature] 10/21/77

13. Signature of Government Technical Representative and Date:

Figure 1

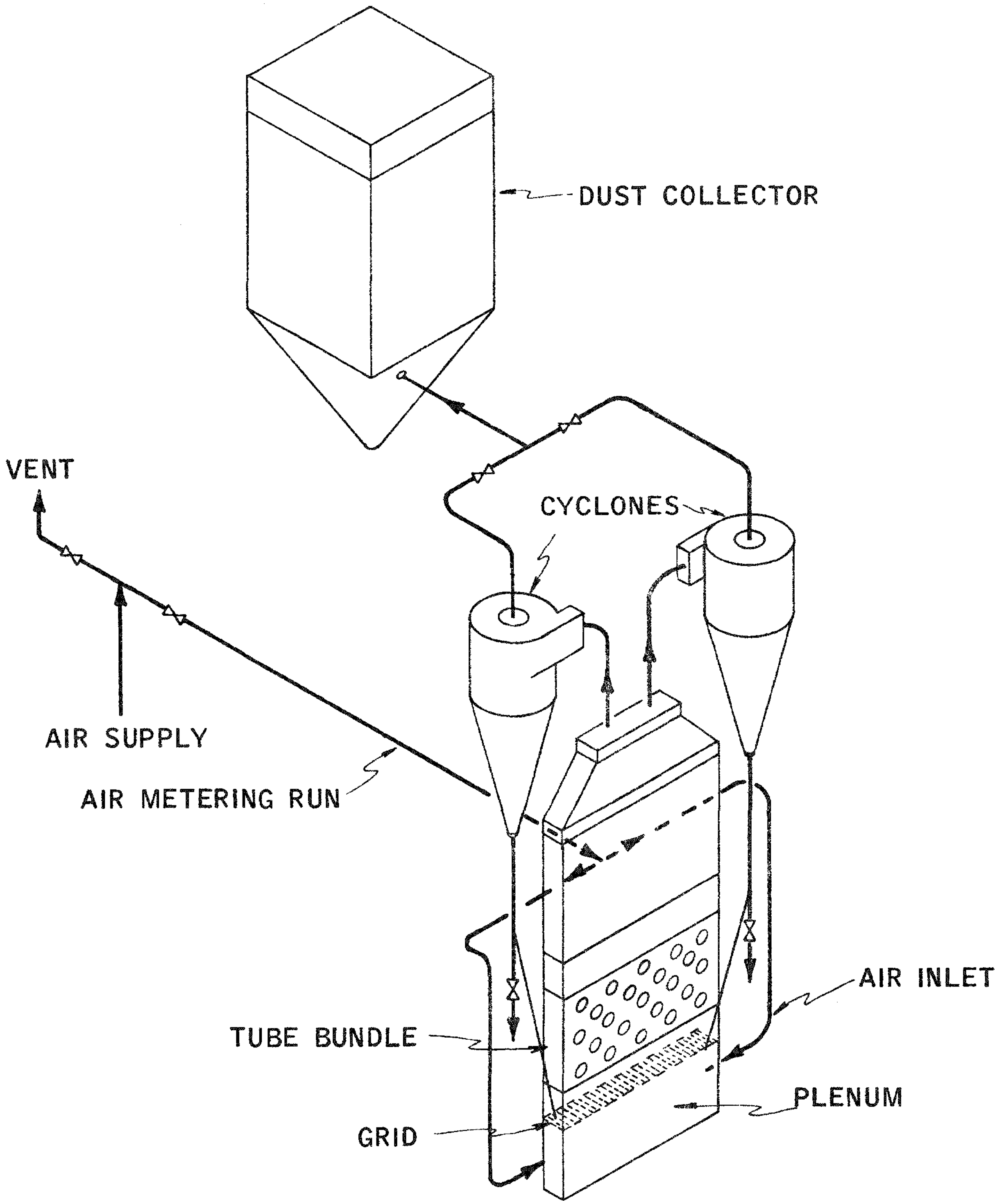


FIGURE 2
2-D FLOW VISUALIZATION UNIT

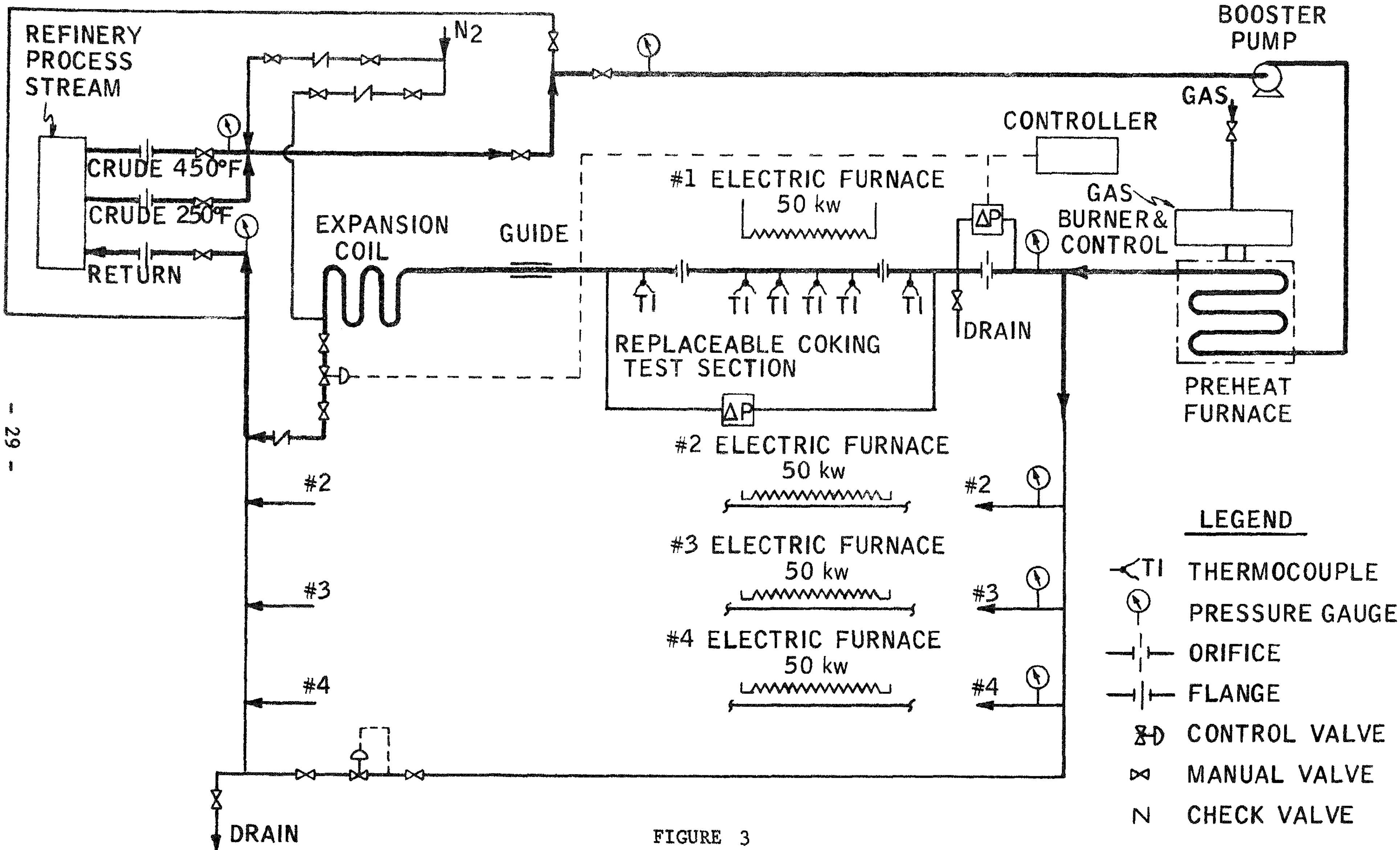


FIGURE 3
PROCESS STREAM COKING UNIT

PARTICLE SIZE DISTRIBUTION OF
LIMESTONE BED FOR "RIVESVILLE CONFIGURATION" TESTS

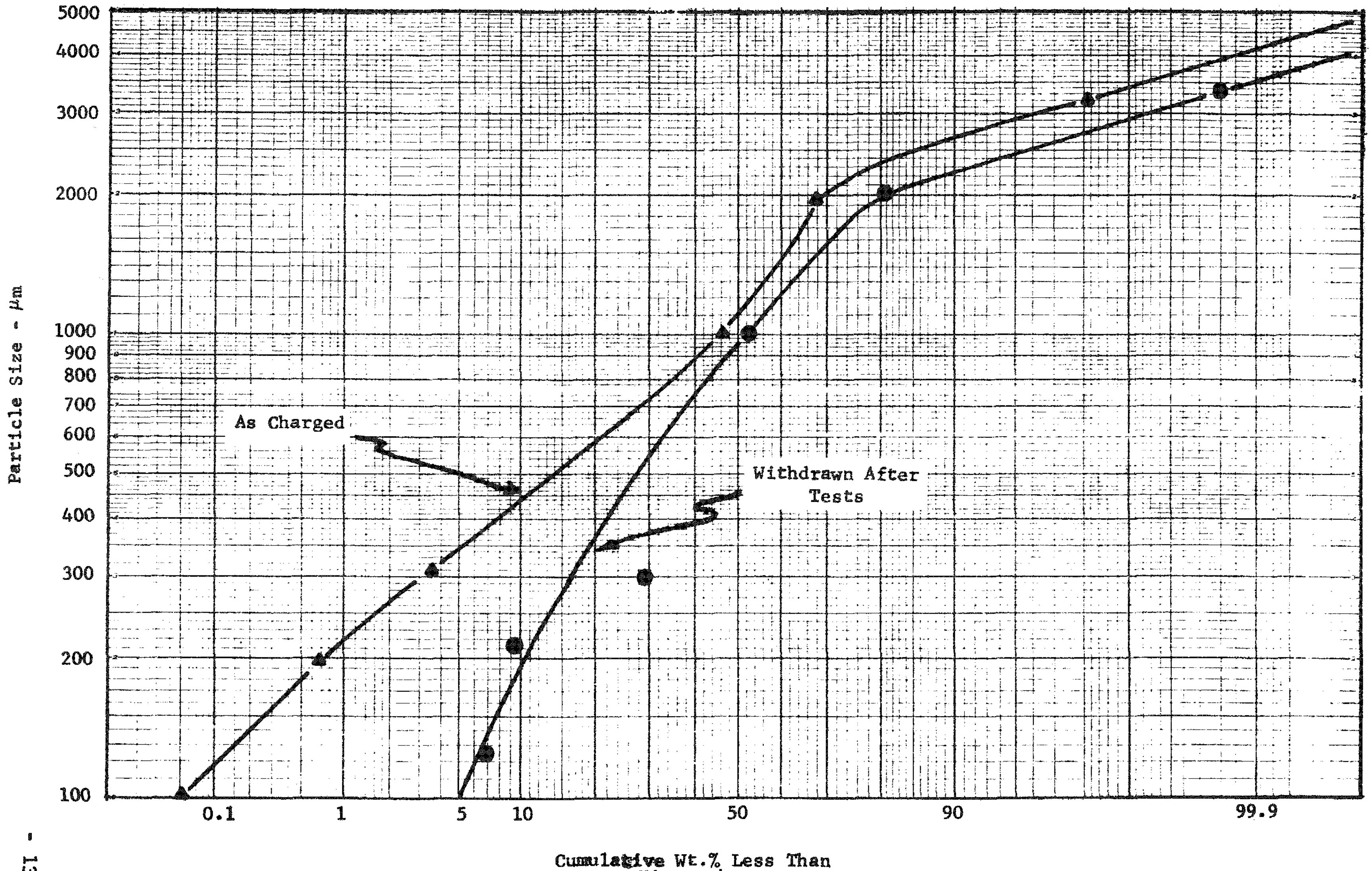


Figure 4

FLUIDIZATION VELOCITY VS. ΔP
RIVESVILLE CONFIGURATION

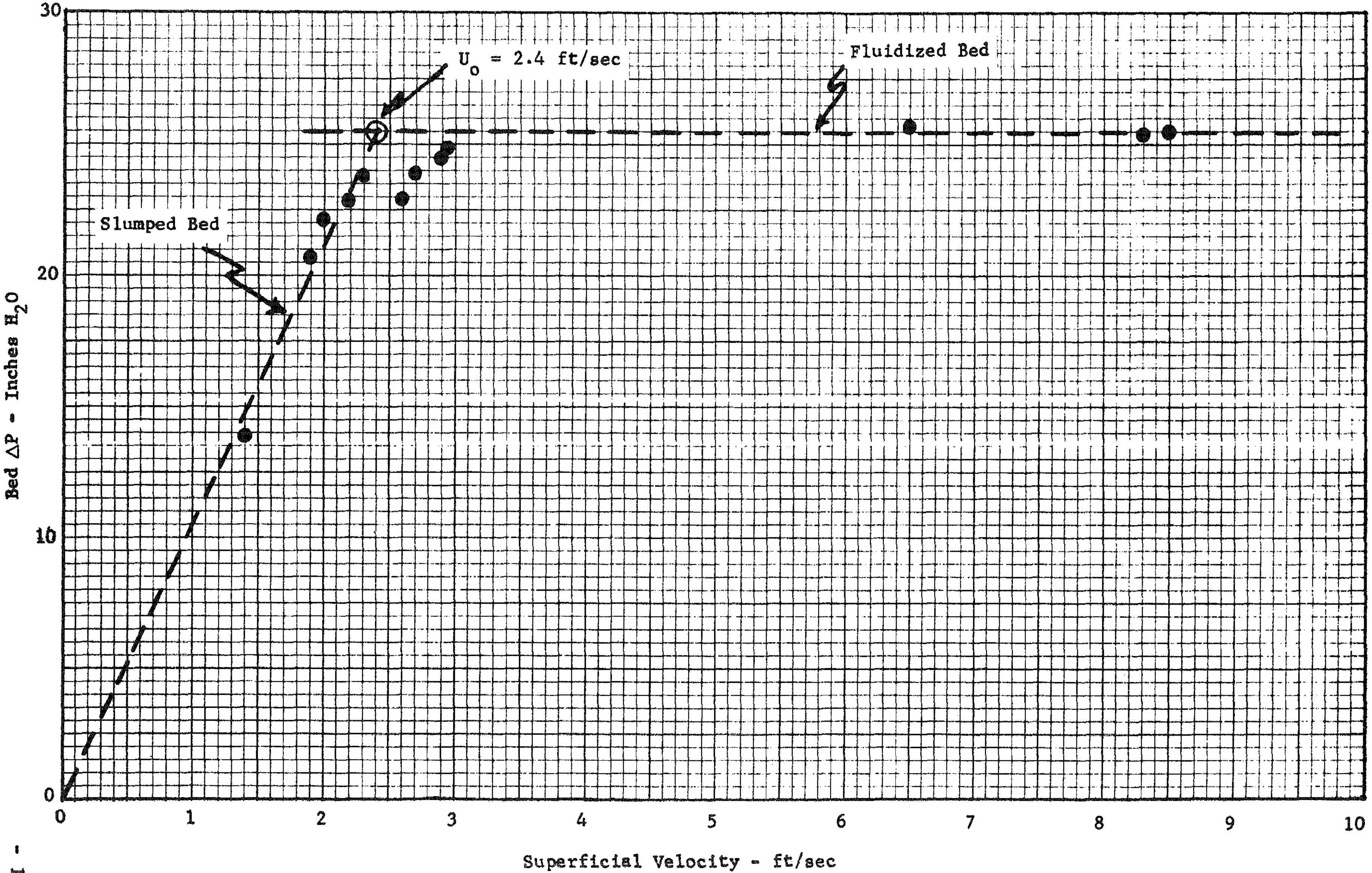


Figure 5

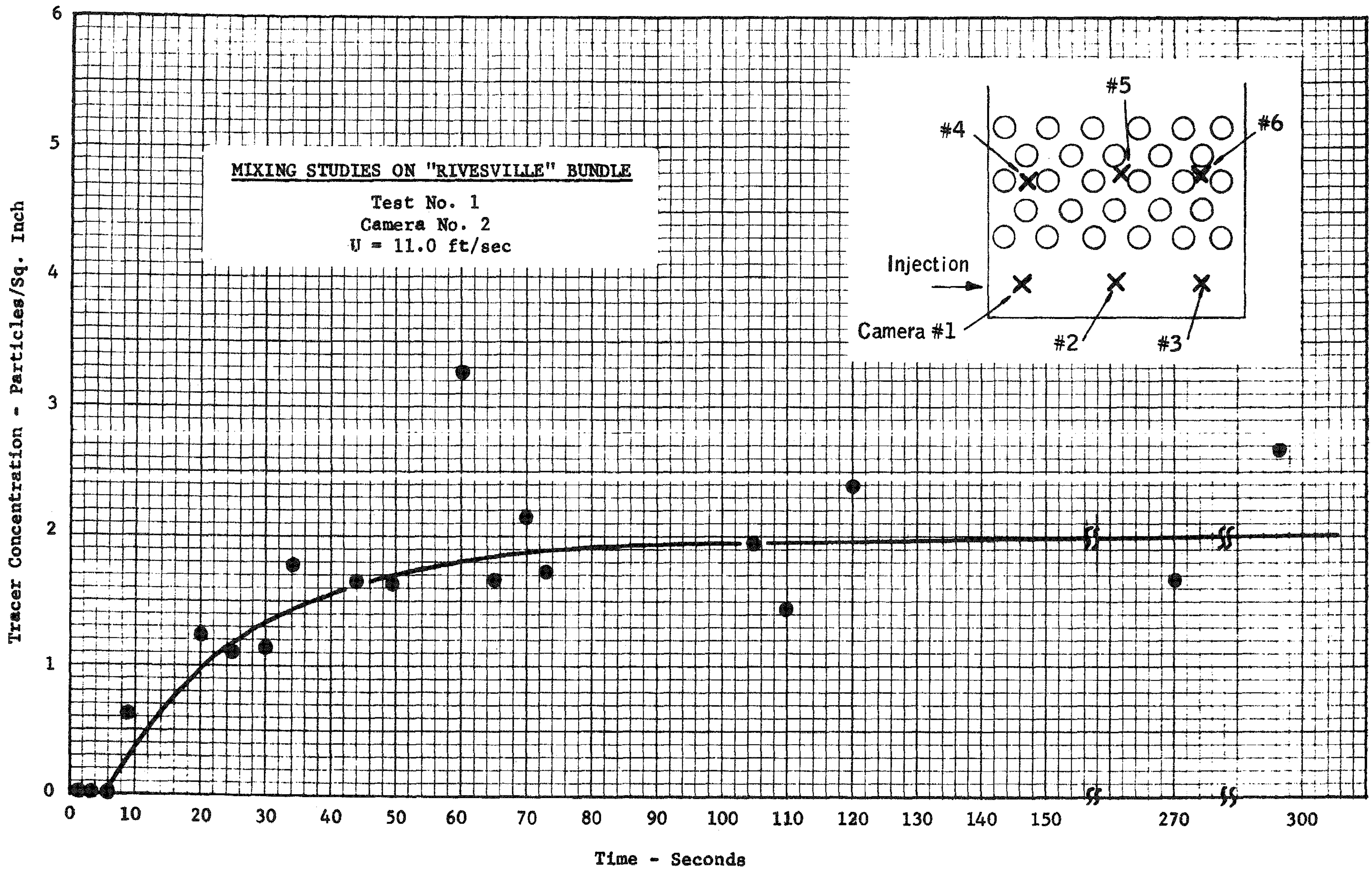
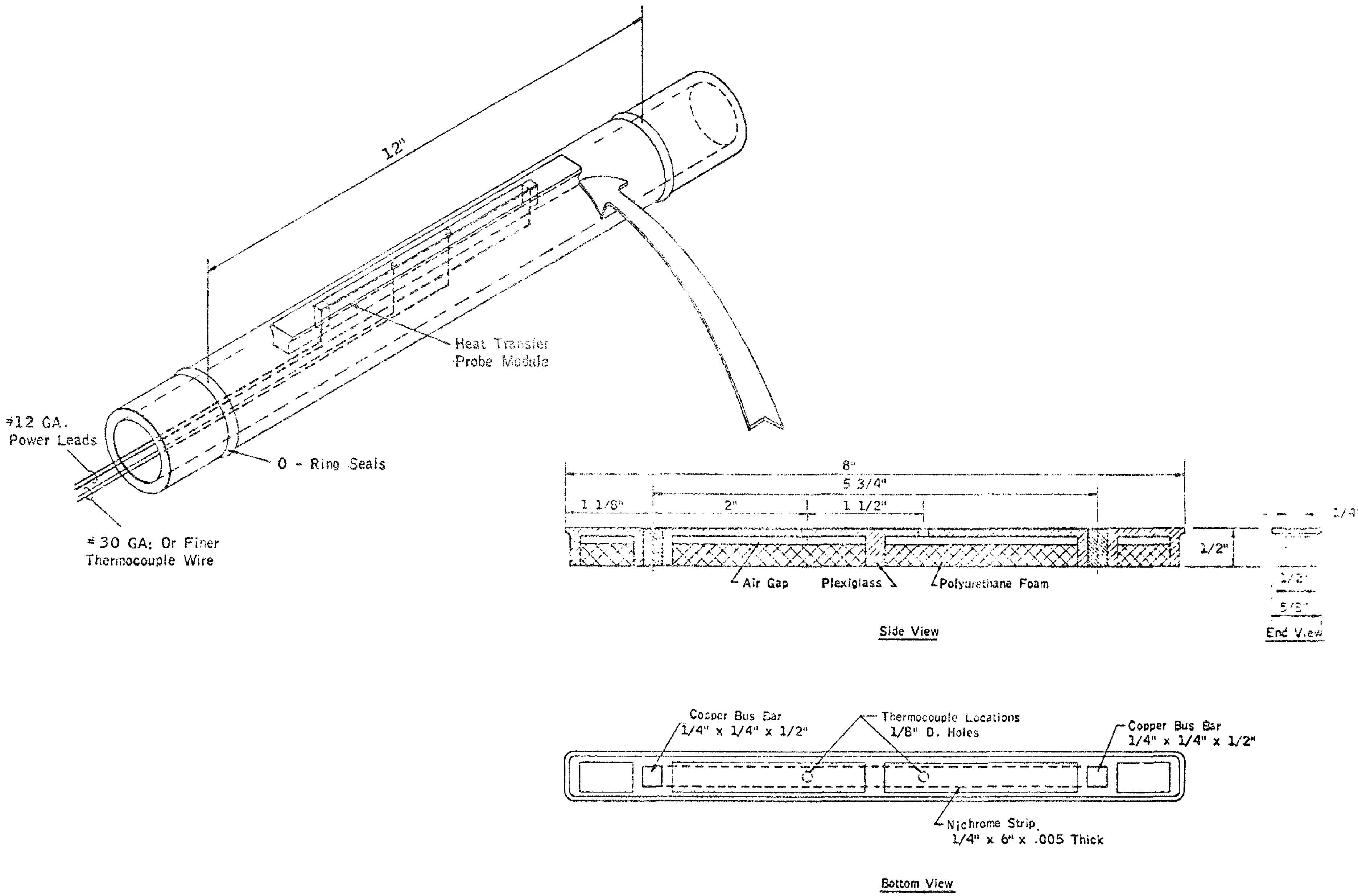

MIXING STUDIES ON "RIVESVILLE" BUNDLE
Test No. 1
Camera No. 2
U = 11.0 ft/sec
#4
#5
#6
Injection
Camera #1
#2
#3
Tracer Concentration - Particles/Sq. Inch
0
1
2
3
4
5
6
0
10
20
30
40
50
60
70
80
90
100
110
120
130
140
150
270
300
Time - Seconds

Figure 6



AMBIENT TEMPERATURE TEST PROBE

FIGURE 7

HEAT TRANSFER COEFFICIENT VS. BED PARTICLE SIZE

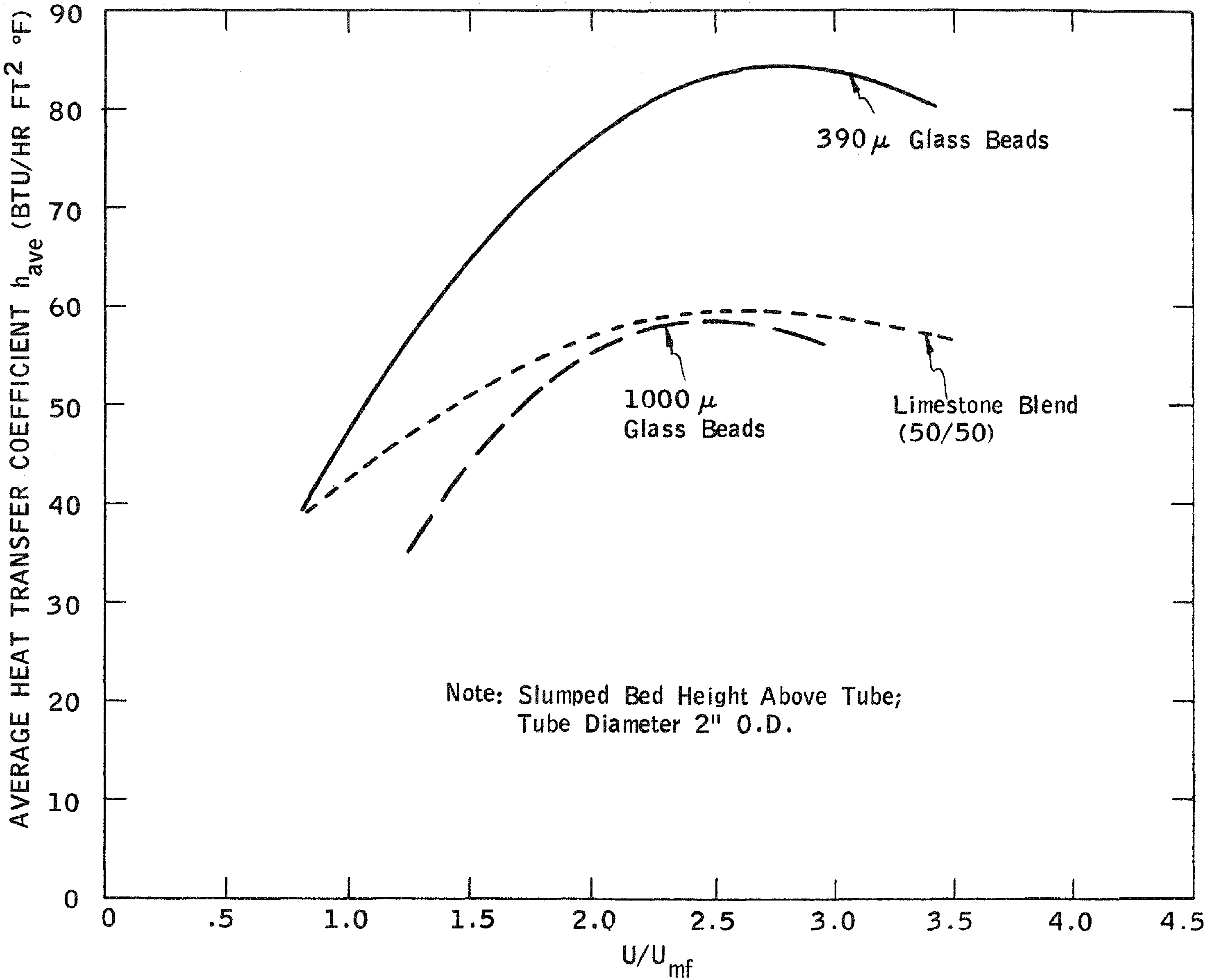


Figure 8

LOCAL HEAT TRANSFER COEFFICIENT VS. FLUIDIZATION VELOCITY
390 μ GLASS BEADS - 2" TUBE DIAMETER

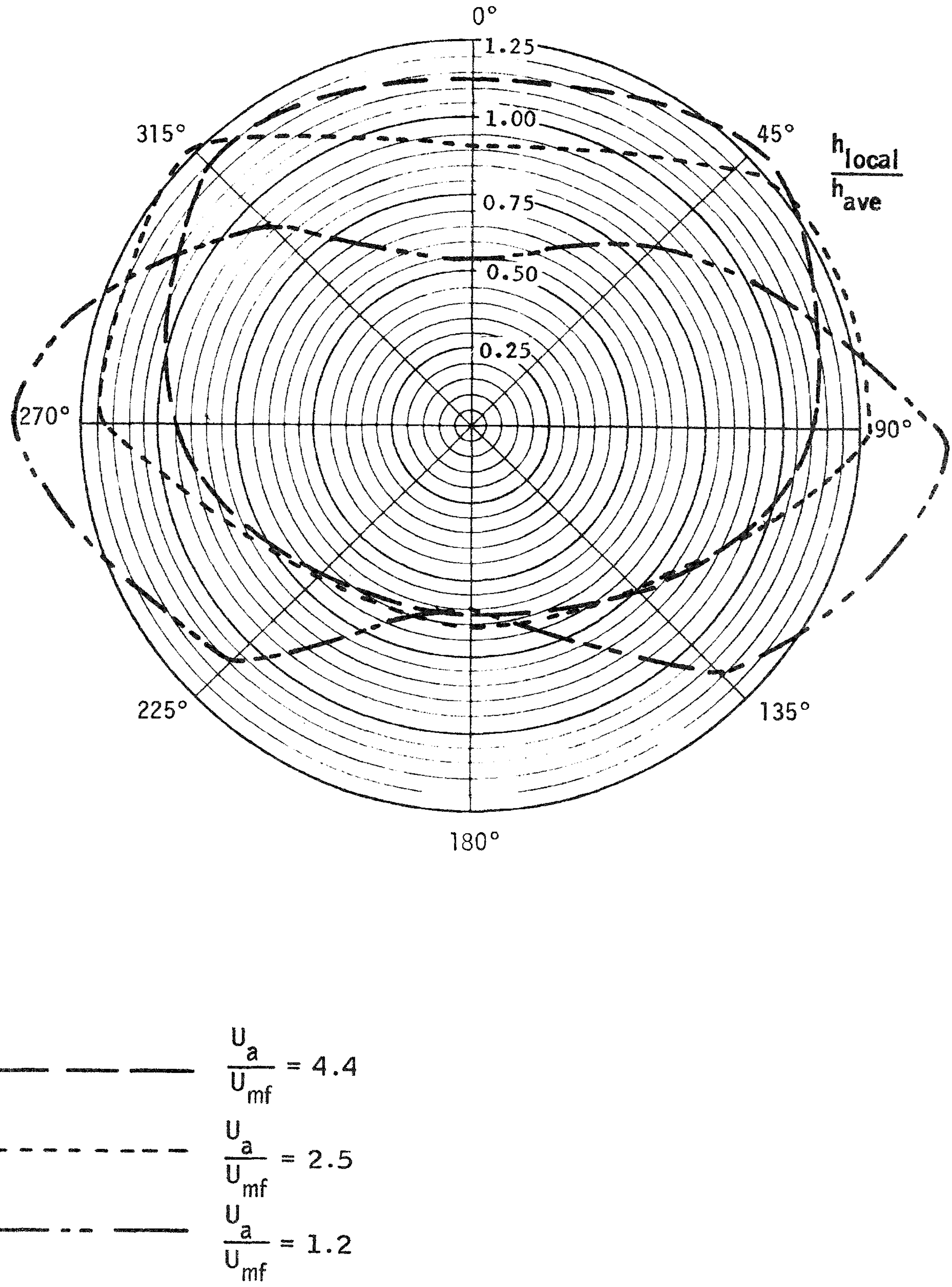
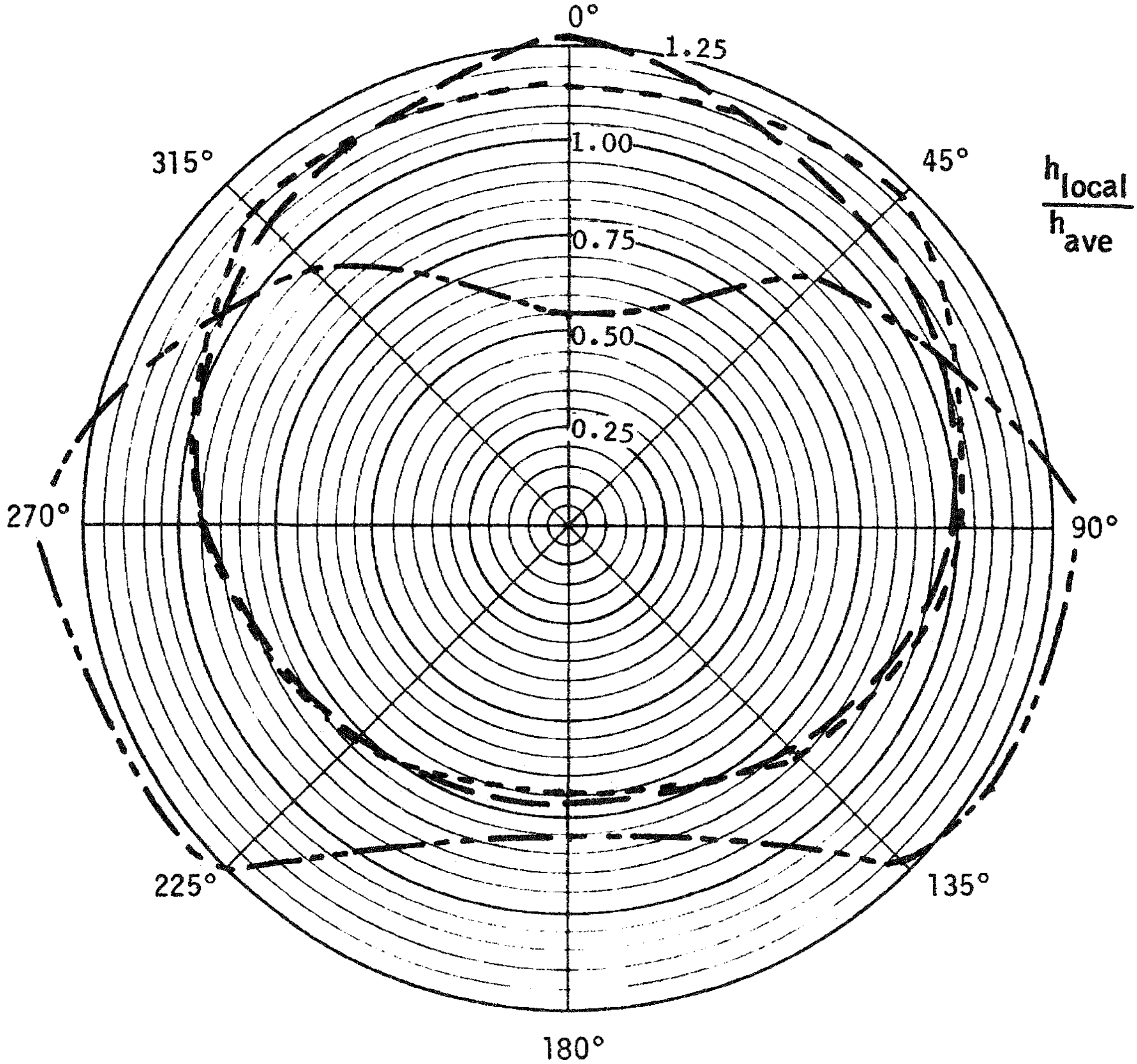


Figure 9a

LOCAL HEAT TRANSFER COEFFICIENT VS. FLUIDIZATION VELOCITY
1000μ GLASS BEADS - 2" TUBE DIAMETER



$\frac{U_a}{U_{mf}} = 4.1$

$\frac{U_a}{U_{mf}} = 3.1$

$\frac{U_a}{U_{mf}} = 1.9$

Figure 9b

LOCAL HEAT TRANSFER COEFFICIENT VS. FLUIDIZATION VELOCITY
BLENDED LIMESTONE - 2" TUBE DIAMETER

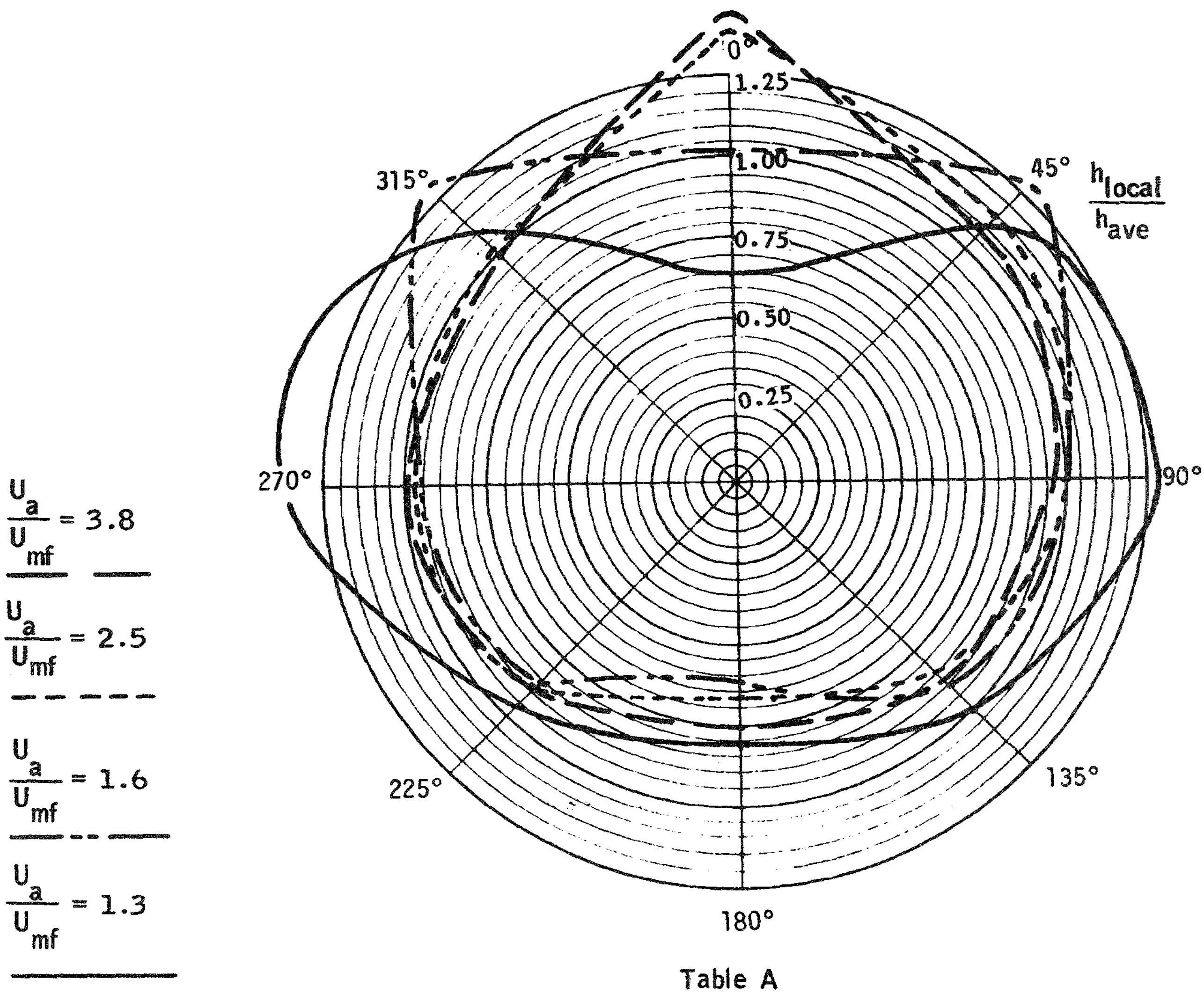


Table A

Characterization Of The Blended Limestone Bed As A Function Of Fluidization Velocity

$\frac{U_a}{U_{mf}}$	% Bed Material Fluidized	Largest Particle Fluidized/50% Particle
1.3	70	2000 μ /800 μ
1.6	95	3000 μ /1000μ
2.5	100	Exceeds Max. Particle In Bed/ 1100 μ
3.8	100	Exceeds Max. Particle In Bed/ 1100 μ

Figure 9c

HEAT TRANSFER COEFFICIENT LIMESTONE VS PARTICLE SIZE

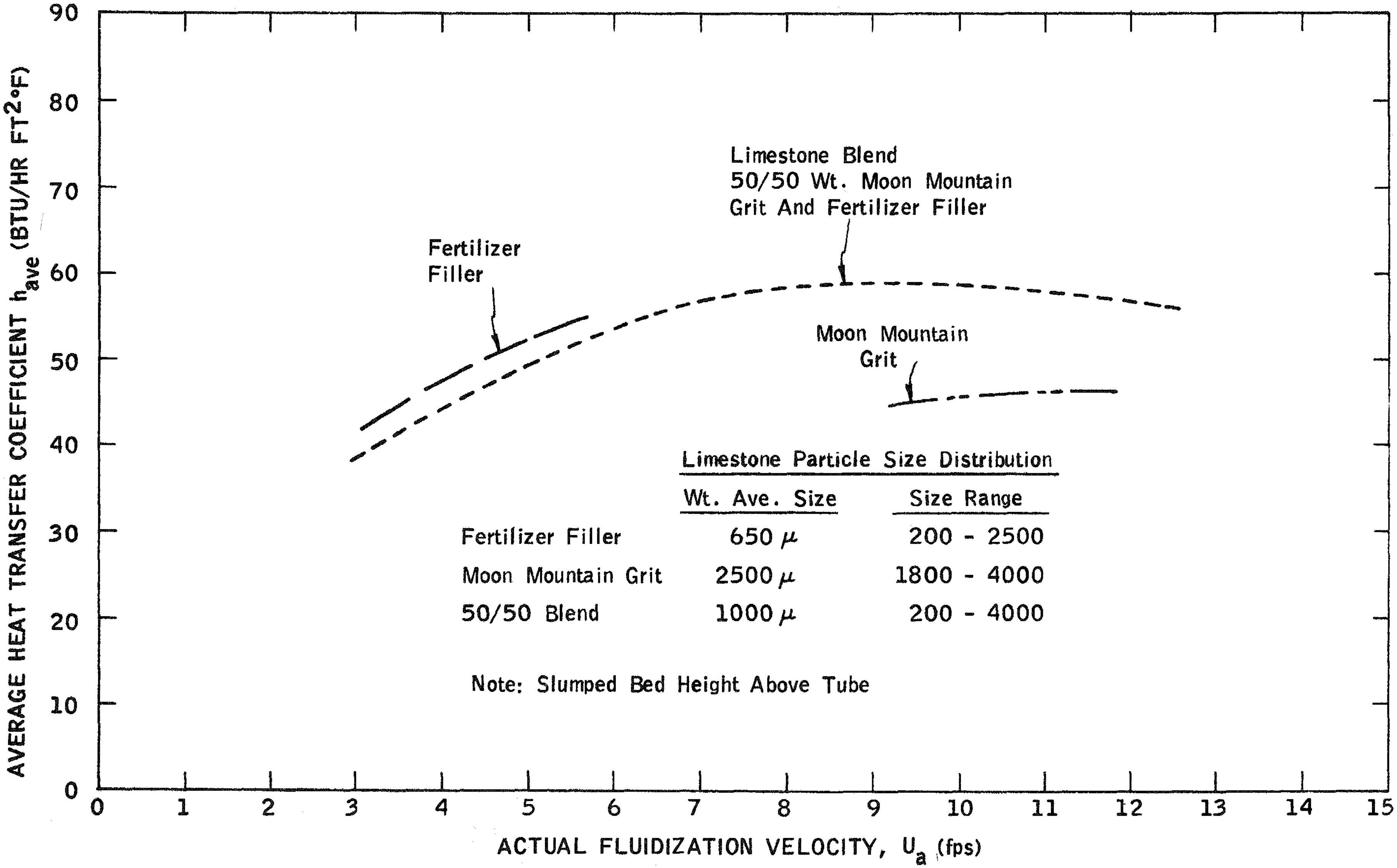


Figure 10

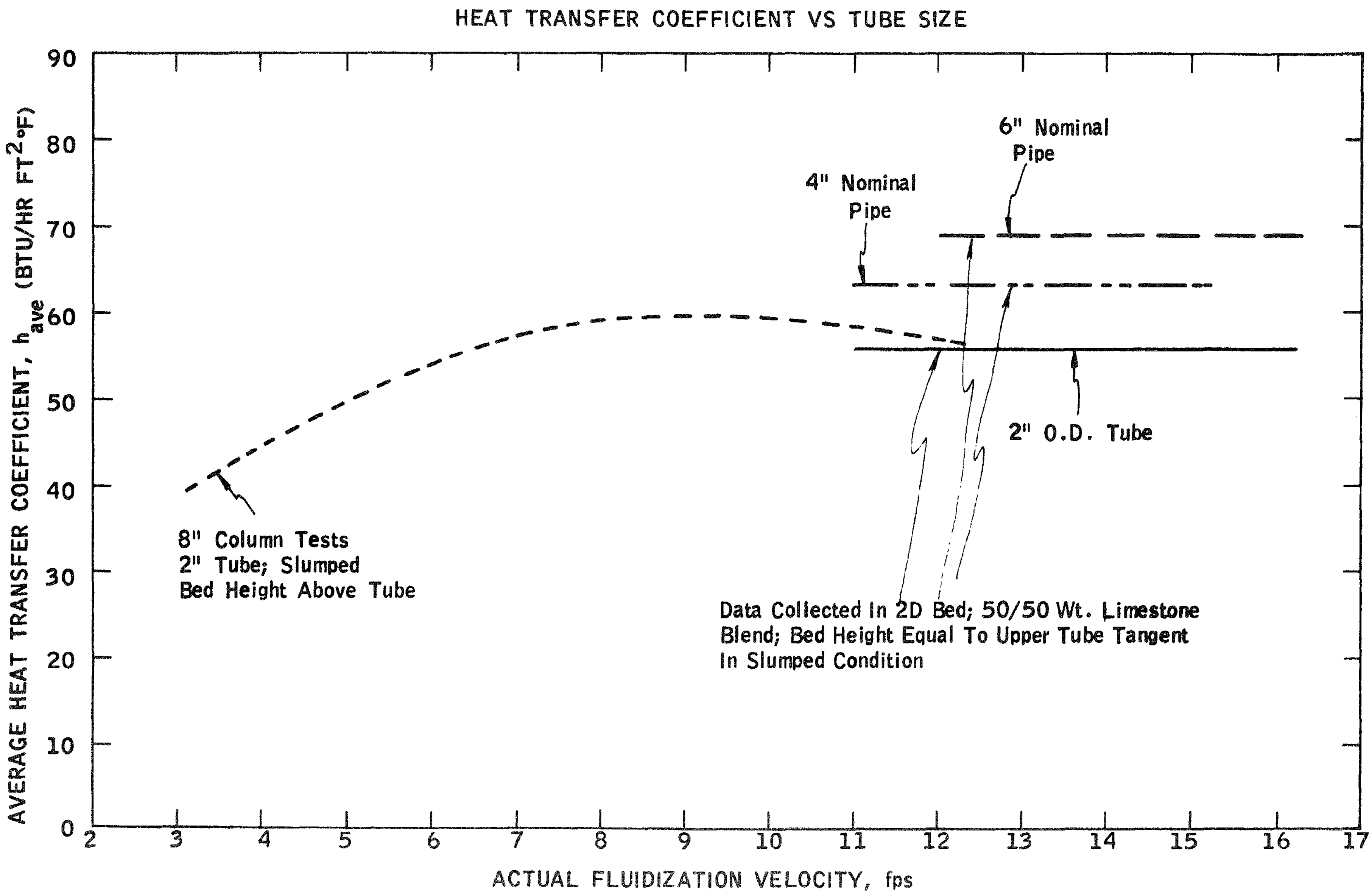
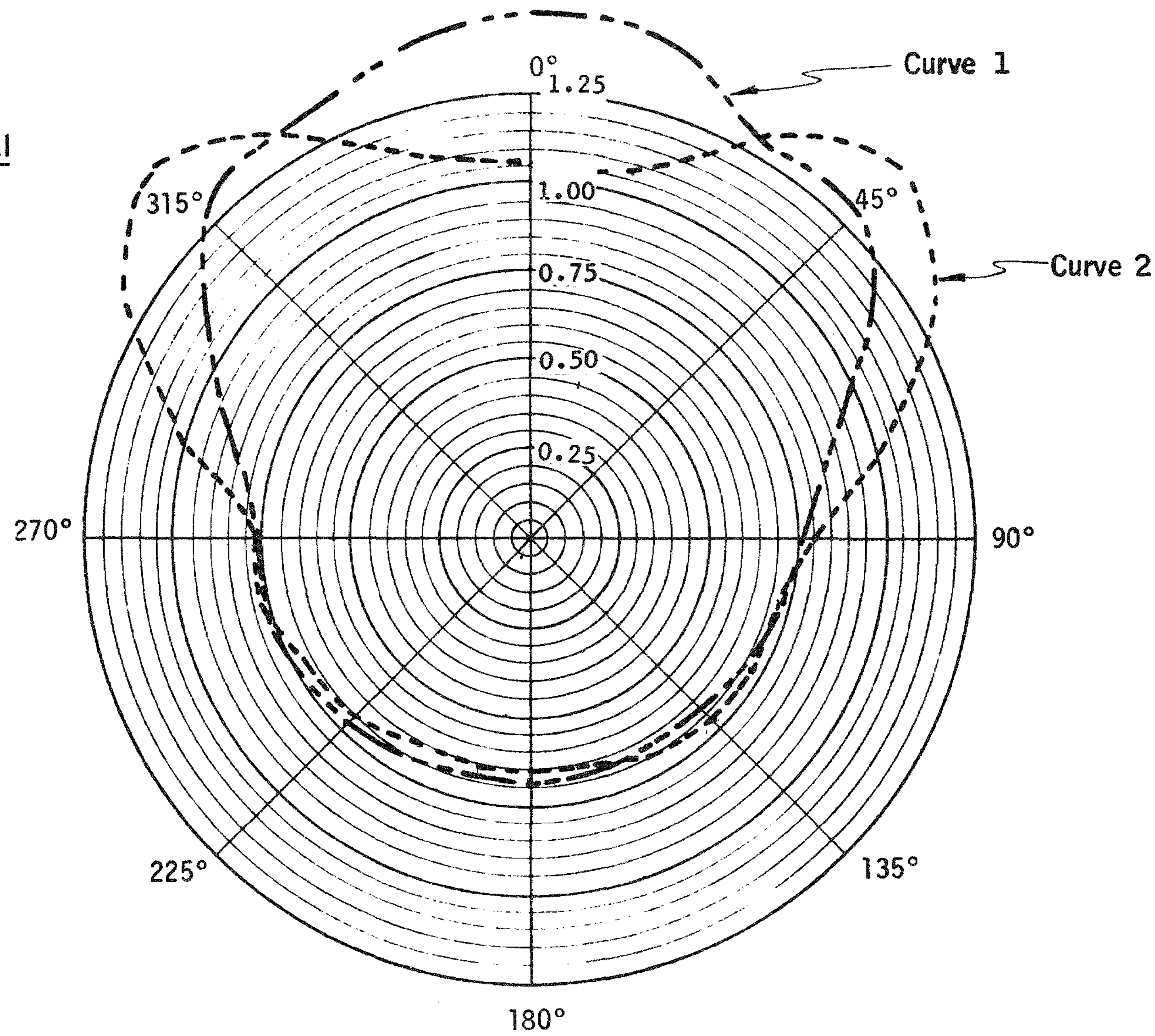

HEAT TRANSFER COEFFICIENT VS TUBE SIZE
AVERAGE HEAT TRANSFER COEFFICIENT, h_{ave} (BTU/HR FT^2°F)
ACTUAL FLUIDIZATION VELOCITY, fps
0
10
20
30
40
50
60
70
80
90
2
3
4
5
6
7
8
9
10
11
12
13
14
15
16
17
6" Nominal Pipe
4" Nominal Pipe
2" O.D. Tube
8" Column Tests
2" Tube; Slumped
Bed Height Above Tube
Data Collected In 2D Bed; 50/50 Wt. Limestone
Blend; Bed Height Equal To Upper Tube Tangent
In Slumped Condition

Figure 11

LOCAL HEAT TRANSFER COEFFICIENT VS. BED DEPTH
BLENDED LIMESTONE - 6" NOMINAL PIPE SIZE $\frac{U}{U_{mf}} = 4.4$



Explanatory Notes:

$\frac{h_{loc}}{h_{ave}}$

Curve 1 - Slumped Bed Depth 21 Inches
$h_{ave} = 68.9$ Btu/Hr Ft^2 °F

Curve 2 - Slumped Bed Depth 15 Inches
$h_{ave} = 64.4$ Btu/Hr Ft^2 °F

Curve 1 And 2 - Tube Center Line 18" From Grid

Figure 12

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