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Heat Transfer Studies to Vertical Tubes in a Circulating Fluidized Bed

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The Mississippi Mineral Resources Institute
University, Mississippi 38677

FINAL REPORT

HEAT TRANSFER STUDIES TO VERTICAL TUBES IN A CIRCULATING FLUIDIZED BED

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ABSTRACT

Circulating fluidized bed combustors (CFBC) employing fine limestone offers many potential advantages over the design approach using a shallow bed of coarse limestone, which has dominated efforts to date. Potential advantages include higher combustion efficiencies, reduced limestone needs, simpler fuel and limestone feed system, and elimination of unacceptable heat transfer surface erosion and fouling and better turn-down and response rate.

Previous work with fluidized bed combustion (4) of Mississippi lignite has lead to problems with ash agglomeration within the bed. This problem will potentially be eliminated with a circulating bed design. Because of the softness of Mississippi lignite, grinding results in a high fraction of fines. Fines are not acceptable in the conventional design, but are easily accommodated in the CFBC.

Design of heat transfer surfaces is critical in the overall equipment design and the operation and control of circulating fluidized beds. This study investigates the heat transfer characteristics from the bed to the wall. Heat transfer - coefficients will be determined as a function of gas velocity, solid suspension concentration, and solids circulation rate. Empirical correlations will be developed.

Specific objectives of this research project which includes the work of MMRI 86-2S are:

1. To construct an experimental facility for studying the heat transfer characteristics in a circulating fluidized bed combustor.
2. To measure the bed-to-wall heat transfer coefficient as a function of gas velocity, solid concentration in the bed and solids recirculation rate in a circulating fluidized bed and to develop a correlation based on the experimental results.

3. To measure heat transfer coefficients for heat transfer from a vertical tube as a function of gas velocity, solid concentration in the bed and solids recirculation rate in a circulating fluidized bed and to develop a correlation based on the experimental results.

Phase I of the project, which was the construction of the experimental facility, has been completed. Experimental data for Phase II is presented in this report. Construction of the experimental facility for Phase III is near completion. Since this project is continuing this year, without support from MMRI, this will be the final report. In this report, the design and construction of the experimental facility is discussed and the methodology used to extract the heat transfer coefficient from the data is presented.

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ACKNOWLEDGMENT

The project direcrors acknowledge the support of this research project by the Mississippi Mineral Resources Institute (Project Numbers MMRI 86-2S and MMRI 87-8F). The efforts put forth by our Graduate Students Mr. Ming-Shau Lu and Mr. Anuchit Sekthira, in the design, construction and operation of the experimental facility is acknowledged. The help in construction of the experimental equipment by Mr. Arthur E. Bowles, our technician, is greatly appreciated. Finally, many thanks go to Mrs. Amy A. Burrow for her typing of the progress reports and of this final report.

NOMENCLATURE

A_i = inner surface area of inner pipe

C = parametric expression

$$= 1.5 \frac{r_i}{r_p}(1 - e)$$

Cp = heat capacity

d_1 = outside diameter of inner pipe

d_2 = inside diameter of outer pipe

de = $d_2 - d_1$

h_i = inside heat transfer coefficient

h_o = water side heat transfer coefficient

κ = thermal conductivity of steel pipe

l = length of water jacket

$\dot{m}$ = mass flow rate of water

q = heat transfer rate

qr = radiant heat transfer rate

r_i = inside radius of inner pipe

r_o = outside radius of inner pipe

r_p = radius of bed particle

T_a = inner surface temperature of inner pipe

T_b = outer surface temperature of inner pipe

T_1 = bed temperature of CFB

Greek Letters

e = void fraction

e_1 = emissivity of particles

ε_2 = emissivity of inner surface of inner pipe

σ = Stefan-Boltzman constant

μ = viscosity

CHAPTER I

INTRODUCTION

Fluidized bed combustion is the only practical means of burning coal or lignite (1) in an environmentally acceptable manner without flue gas scrubbing for SO-

While the outlook for industrial conventional fluidized bed combustors (FBC) appears promising, the evolution of these combustors for utility applications is less certain (2). This may be because the early work ”locked in” a single design which has some major short comings. This conventional design uses a bed of coarse limestone particles operating with gas velocities around 10 ft/s. The bed contains arrays of heat exchange tubes submerged in it. The FBC has limited combustion efficiency because of char particle carry over and short residence time in the bed. The combustor is unable to accommodate coal or lignite fines because fines are quickly carried out of the combustor. High Ca/S ratios are required because limestone particles are large and gas residence time is short. The performance of the FBC can be improved by the recycle of limestone, ash and char and an extended freeboard space.

Recycle and extended reactor space are two of several important features of the circulating fluidized bed combustor (CFBC). In the CFBC the fuel is burned by air in a suspension of finely divided limestone at gas velocities of 20-30 ft/s and at temperatures around 1500-1600^{o}F. The suspension leaving the top of the combustor is circulated back to its bottom by cyclones and a large standpipe. The circulation rate of the suspension is many times greater than the combined limestone and fuel feed.rate. A smaller stream of solids is withdrawn from the system to remove ash and spent sorbents.

The circulating fluid bed combustor offers a number of potential advantages over the conventional designJ

(a) Accommodation of fines

(b) Enhanced combustion efficiency

(c) Higher efficiency of limestone utilization

(d) Good capability of load following

(e) Easier distribution of air, fuel and limestone

(f) Lower potential for heat exchange surface erosion, corrosion and fouling

(g) Much lower potential for ash agglomeration

A conceptual design of a CFBC is shown in Figure 1. Small industrial CFBC have been in operation (1,3) for steam production. In these units the combustion chamber (bed) is water jacketed. Much of the extracted energy is transferred in the super heater or boiler bank. In the combustion chamber heat transfer surfaces are parallel to the flow of air and the suspensed solids. This greatly reduces erosion of the heat transfer surfaces when compared to the conventional design and also reduces the potential for ash agglomeration.

For large utility applications, multiple combustors or a combustor with suspended heat transfer surfaces will be needed. Applications with Mississippi lignite which has a very high percentage of volatiles may need more than a water jacketed combustor for heat exchange in the combustors. Even for smaller units, the required heat transfer area can be reduced by the suspended surfaces in the combustor. These surfaces would need to be vertical tubes so that the flow of gases and suspended solids would be parallel to the surfaces. This arrangement will minimize erosion and ash agglomeration.

The combustion of Mississippi lignite in a CFBC brings some questions to surface that need to be studied. Some of these are:

(a) Heat exchange rates in the combustor

(b) SO^ removal efficiency with ash and limestone

(c) Combustion efficiency

(d) Ash agglomeration in the combustor and cyclones.

The work proposed in this study has dealt with heat transfer in the circulating fluidized bed. Limited information is available for heat transfer coefficients in a circulating bed. In this study, the heat transfer coefficients to the bed wall were measured as a function of gas velocities and suspended solid concentration rate. In the next stage of the project, experimental studies on heat transfer to the immersed surfaces in the CFBC will be performed. In addition, a correlation of the heat transfer coefficient to the bed wall as well as immersed surfaces as a function of operating parameters will both be determined. This information is instrumental to a better design of a CFBC.

Specific objectives of this research project were:

1. To construct an experimental facility for studying the heat transfer characteristics in a circulating fluizied bed combustor.
2. To measure the bed-to-wall heat transfer coefficient as a function of gas velocity, solid concentration in the bed and solids recirculation rate in a circulating fluidized bed and to develop a correlation based on the experimental results.
3. To measure heat transfer coefficients for heat transfer from a vertical tube as a function of gas velocity, solid concentration in the bed and solids recirculation rate in a circulating fluidized bed and to develop a correlation based on the experimental results.

Phase I of the project, which was the construction of the experimental facility, has been completed. Experimental data for Phase II is presented in this report. The construction of equipment for Phase III has been completed. Experimental results will be obtained in coming months without support from MMRI. In this report, Chapter I and II are devoted to the discussion of the objectives of this research and review of the published literature on this subject. The presentation of experimental results of Phase II will be presented in Chapter V.

CHAPTER II

SCIENTIFIC DISCUSSION

2.1 Evolution Of The CFBC

Experimental work directed by the principal investigators (4) has shown that the conventional FBC is difficult to operate with Mississippi lignite. The two major problems with burning Mississippi lignite in a conventional FBC are ash agglomeration within the bed and the generation of fines during grinding.

Because of the softness of Mississippi lignite it appears impossible to get a large fraction of lignite particles in the 1 to 6 mm particle range required by a conventional FBC. Large fraction of fines are generated. Fines cannot be used in a conventional FBC because they are carried out of the bed before significant devolatilization and/or combustion can occur. This problem can be partially corrected by the recycle of carryover material.

Mississippi lignite ash has been shown (4) to have a low melting and/or softening point. This results in ash agglomeration within the fluidized bed centering around the submerged heat transfer tubes. As the agglomeration grows the bed becomes defluidized and the combustor fails to operate.

Our work has also shown (4) that even under the best of operating conditions Mississippi lignite ash will not capture enough SO^ to meet environmental standards. However, under favorable oxygen levels the capture of SO^ by the ash can significantly reduce limestone usage.

The evolution of the circulating fluidized bed combustor (CFBC) in the past few years has demonstrated (3) a new design path. It appears that the new design accommodate the disadvantages of the conventional design and maintains many of the advantages of the conventional FBC.

2.2 Advantages of The CFBC

2.2.1 Limestone Utilization

In the CFBC finely divided limestone particles are used as bed material. The limestone, lignite and ash will be circulated through the combustion chamber (see Figure 1). This will provide good contacting between the limestone/ash and the SO_2 generated during combustion. The conventional design required about a 3 to 1 ratio of calcium to sulfur. It appears that the calcium to sulfur ratio can be reduced (5,6) to 1.5 in a CFBC.

Under the best of conditions it can be assumed (4) that the ash will capture 40% of the sulfur. To meet the environmental standard of 1.2 lb SO_2/10^6 Btu, using the optimistic ash capture assumption, the conventional design would require about 250 lb of limestone per ton of dry lignite. The CFBC would require about 125 lb of limestone per ton of dry lignite. This is a significant savings in cost and material handling. The above calculation is for a Mississippi lignite with 3 percent sulfur.

2.2.2 Lignite Fines and Combustion Efficiency

The generation of a very large fraction of fines during grinding of Mississippi lignite makes the use of a conventional FBC almost impossible. The fines are in the bed such a short period of time that combustion efficiency is lowered. The fines are carried to the freeboard and out of the system. The O_2 concentration in the freeboard of the conventional design is low; therefore, combustion is not promoted in the freeboard.

In the CFBC fines are continually circulated through the combustion chamber and secondary air is used at higher elevations in the chamber to promote combustion. High combustion efficiencies on other fuels have been demonstrated (6).

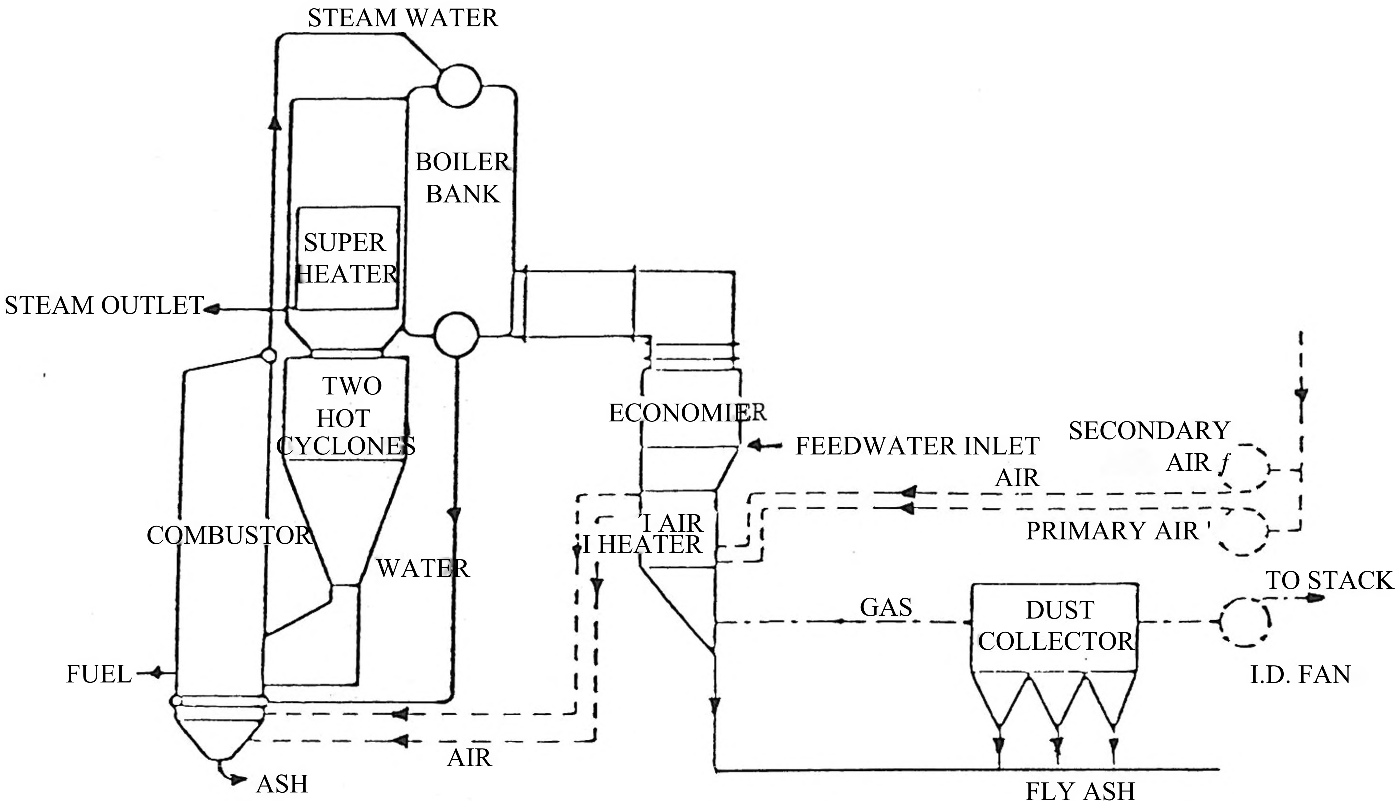


Figure 1. Circulating Fluidized Bed Boiler System (1)

2.2.3 Ash Agglomeration On Heat Transfer Surfaces

In the CFBC high heat transfer rates are not limited to a bed of fixed height. In designs to date, heat transfer has been facilitated by a water jacketed combustion chamber. The ash fines will be carried through the chamber at relatively high velocities and will probably not agglomerate to the chamber walls. Without heat transfer internals perpendicular to the flow of gas and solid suspensions, the potential for ash agglomeration will be greatly reduced.

2.2.4 Heat Exchange Surface Erosion

With heat transfer surfaces parallel to the flow of gas and suspended solids, erosion of heat transfer surfaces with a CFBC will be greatly reduced. This differs from the conventioal FBC where heat transfer surfaces are generally perpendicular to the flow of solids.

2.2.5 Turn Down And Response Rate

The CFBC can use greater air velocities because the entrained particles are separated from the hot gases in a cyclone collector and reinjected into the bottom of the combustion chamber. Turn down ratios of 3 to 1 can be achieved merely by changing flow rates and the fuel feed. Fifty percent load change within three minutes have been demonstrated (3). This flexibility does not exist in the conventional design.

2.2.6 Fuel Feeding

Burning Mississippi lignite in a conventional FBC leads to volatile combustion in the freeboard where heat transfer coefficients are much lower (7). This was observed to occur (4,8) with both under-bed and over-bed feeding because lignite particles that are devolatilizing in a bubbling bed tend to ”float” in the top of the bed.

With a CFBC design the heat transfer coefficients are the same order of magnitude at all locations within the combustion chamber (7). With volatile and char combustion throughout the combustion chamber in a CFBC system, the necessary heat transfer will be accommodated with less heat transfer area. Fuel can be fed to the combustion chamber at any location, but the optimum location will probably be near the bottom of the chamber.

2.3 Heat Transfer Studies

A first step in developing a CFBC for use with Mississippi lignite will require an understanding of local heat transfer coefficients to surfaces as a function of gas velocities and solid circulation mass rates. This is not only important for the design of the system but also for load control since gas velocities and solids circulation rates are key control variables.

Various heat transfer processes occur in a CFBC. These include heat transfer between particles and gas, between particles and other particles, and between the solids suspension and heat transfer surfaces (either at the wall of the combustor or suspended in the combustor). However, the key processes for the design and operation of CFB are those between the gas-solids suspension and heat transfer surfaces. This study investigated the heat transfer coefficient from the bed to the combustor wall. It is understood that if sufficient heat transfer surface area is available, extracting heat from the wall is the best design. This is because suspended heat transfer surfaces are subject to erosion and impede radial mixing of gas and solids. They may also reduce solids hold-up in the combustor and promote attrition.

Although there are a large number of studies on heat transfer in fluidized beds, very few studies (9, 10, 11) are performed in the circulating fluidization regime. Grace (12) recently presented an excellent review on published studies related to heat transfer in circulating beds and are summarized in Table 1 (12). The range of variables covered and the amount of information supplied by the authors are limited. All reported measurements are for relatively small columns at atmospheric pressure. Most are for suspended heat transfer surfaces rather than surfaces attached to the wall. Most heat transfer coefficients lie in the range from about 100 to 350 W/m^2K. This range of values and the strong dependence on suspension density are consistent with values given by Reh et al. (13, 14) and the trends reported by Pell and Johnson (14). The limited data from explicitly circulating bed units are supplemented by the early study of Mickley and Trilling (15). Many of their data appear to correspond to what we now refer to as fast fluidization with superficial velocity as high as 4 m/s and solids volumetric concentration as low as 2%.

The strong role of suspension density is demonstrated in Figure 2 where data from Mickley and Trilling (15), Kiang et al. (11), Stromberg (17,18), and Fraley et al. (10) are plotted. Suspension densities in all cases have been deduced from measured pressure drops. All of the lines and points in Figure 2 are for air at or near atmospheric pressure and modest temperatures where radiation effects are minor and gas properties nearly constant. Heat transfer surfaces were on the axis of the column in each case, with the exception of Stromberg whose measurements pertain to the outer wall. Figure 2 presents a reasonably consistent pattern, showing that heat transfer increases with suspension density and with decreasing particle size in the ranges covered by the data. The recent data of Subbarao et al. (19) agree with these trends, although they show a broader scatter.

TABLE 1

Experimental details of published experimental studies where authors measured surface-to-suspension heat transfer in circulating beds. Only data where 0.01 <_C^_ 0.02 and U > 1.5 m/s have been included, other conditions almost certainly lying outside the "fast fluidization" regime. (12)

Authors:	Mlckley and Trilling (1949)	Mlckley and Trilling (1949	Kiang et al (1976)	Stromberg (1981) Kobro 6 Brereton (1985)	Fraley et al (1983)	Chrysostome et al (1984)	Subbarao et al (1985)
Column :							
Type	Steel pipe	Aluminum tubes	Plexiglass tube	Steel pipe	Cyl. tube	Refractory lined	Plastic tube
Diameter (m)	0.073	0.10 and 0.025	0.10	0.20	0.076	0.50	0.10
Length (m)	1.3	2.5 and 1.9	3.66	3.0	1.28	5.0	5
lient transf. surface:	Calrod tube	External vail	Short cylinders	Ht. flow meters	Thin cylinder	Finned tubes	End of steel rod
Type	Elect, heating	Elect, heating	Elect, heating	Calorimetric 8 conductivi ty	Elect, heating	Water cooled	Bolling vater
Number	1	1	4	2	1	1	2
Length (m)	0.88	0.85 and 0.91	0.057	N.S.	0.15	N.S.	0.10
Dianeter (mm)	0.012	0.10 and 0.025	19	N.S.	9.5 *	13 6 25	25
Height (m)	-0. 2	-0.9 and 0.5	0.53,1.3,2.40,3.1	Various	N.S.	N.S.	0.22 and 1.6
Radial position, r/R	0.17	1	0 (axis only)	1	0 & -0.8	0,0.33,0.6	1
Partirle*;							
Type	Glass beads	Glass beads	Cracking catalyst	Sand	Class beads	Silica sand	Sand
Diameter (um)	40 to 450	70 to 450	-53	170,250	37	-280	130, 260
Density (kg/mb	2420-2720	2440-2720	-1700	2630	-2650	-2600	2650
Operating conditions:							
Pressure (bar)	1	1	I	I	1	1	1
Bulk temperature (X)	-320-380	—'•20-510	N.S.	1123 4 298	294-304	1073-1173	-300
Surface temp. (X)	-420	-530	N.S.	N.S.	327-407	N.S.	-350
Sup. gas ve). (m/s)	up to 2.5	up to 4.1	up to 4.9	-6-12	up to 2.8	2-5	N.S.
Solids vol. fract. ,C	0.02 - 0.20	0.02 - 0.20	0.013-0.029	< 0.01 - 0.033	0.045-0.18	N.S.	<0.01 - 0.04
Measured heat transfer coefficients (W/m²K)	50-400	120-500	-120-260	70-280	240-770	170-380*	140-350
Vat talilen i	Cля velocity Particle size Susp. density	Gas velocity Particle size Tube diameter Susp. density	Cas velocity Recycle rate Vertical position Solids Inventory	Particle size Bulk temperature Susp. density	Cas velocity Solids flow Radial position	Cas velocity Solids inventory Radial position Fin thickness	Particle site Susp. density

* hasrd OП hnrn fuhn ven

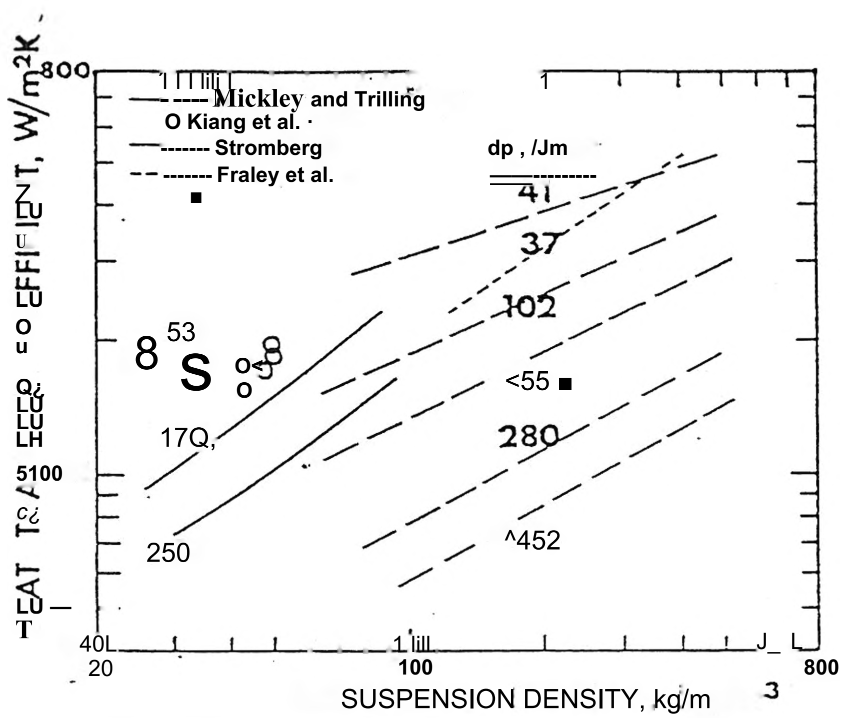


Figure 2 - Circulating bed to surface heat transfer coefficient as a function of the suspension density inferred from pressure drops from data presented in the open literature. (12)

Figure 3 (Grace (12)) plots data of Kiang et al. (11) showing that local heat transfer coefficients generally fall off with increasing height, at least when the measurements are taken along the axis of the column. This decrease is associated with a decrease in solids concentration with height as measured by Yerushalmi and Squires (20), Li and Kwauk (21), Weinstein et al. (22). At the lowest gas velocity, U = 0.6 m/s, the bed was probably operating in the bubbling or turbulent fluidization regime where the heat transfer coefficient falls off sharply with height, upon reaching the freeboard (George and Grace, (22)). As the gas velocity increases beyond about 1.5 m/s to produce "fast fluidization", the axial profiles become much less steep. The data of Genetti and Knudsen (7) show similar trends, but with an increase in h at the top of their column where particles were again present in greater concentration.

Data at Fraley et al. (10) plotted in Figure 4 (Grace (12)) indicate that heat transfer also varies with radial position. For high solids concentrations higher coefficients are found very near the containing wall of the circulating bed column than along the axis. This trend is consistent with hydrodynamic findings (e.g. Bierl et al. (24), Weinstein et al. (25)) showing that particles are more concentrated along the walls, while there is relatively dilute upflow in the interior of the column. Fraley et al. (10) fitted the equations

$$h = 10.2\,\rho_{susp}^{0.656} \quad \text{(SI units)} \qquad (2)$$

$$h = 2.87\,\rho_{susp}^{0.902} \quad \text{(SI units)} \qquad (2)$$

to their on-axis and near-wall data, respectively. These lines shown on Figure 4 cross at $\rho_{susp} = 173$ kg/m^3 (C ~ 0.065) indicating that the trend may reverse at low solids concentrations, with higher heat transfer then appearing near the axis. Data by Chrysostome et al. (26) for finned tubes indicate

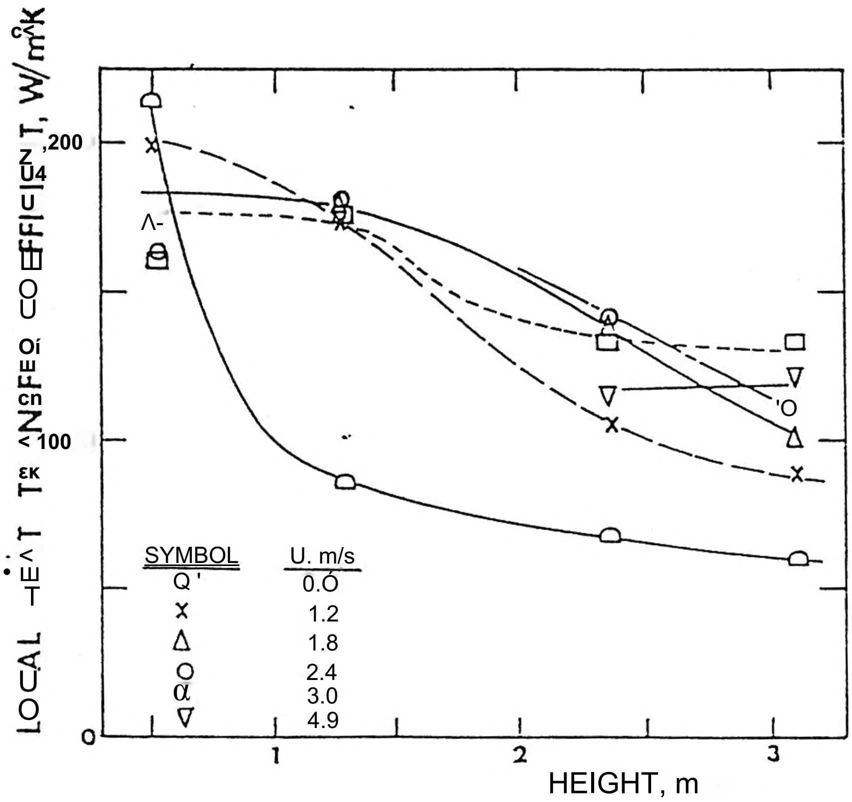


Figure 3 - Local circulating bed. to surface heat transfer coefficient as a function of height and superficial gas velocity from the data of Kiang et al (1975) for 53 μm cracking catalyst. (12)

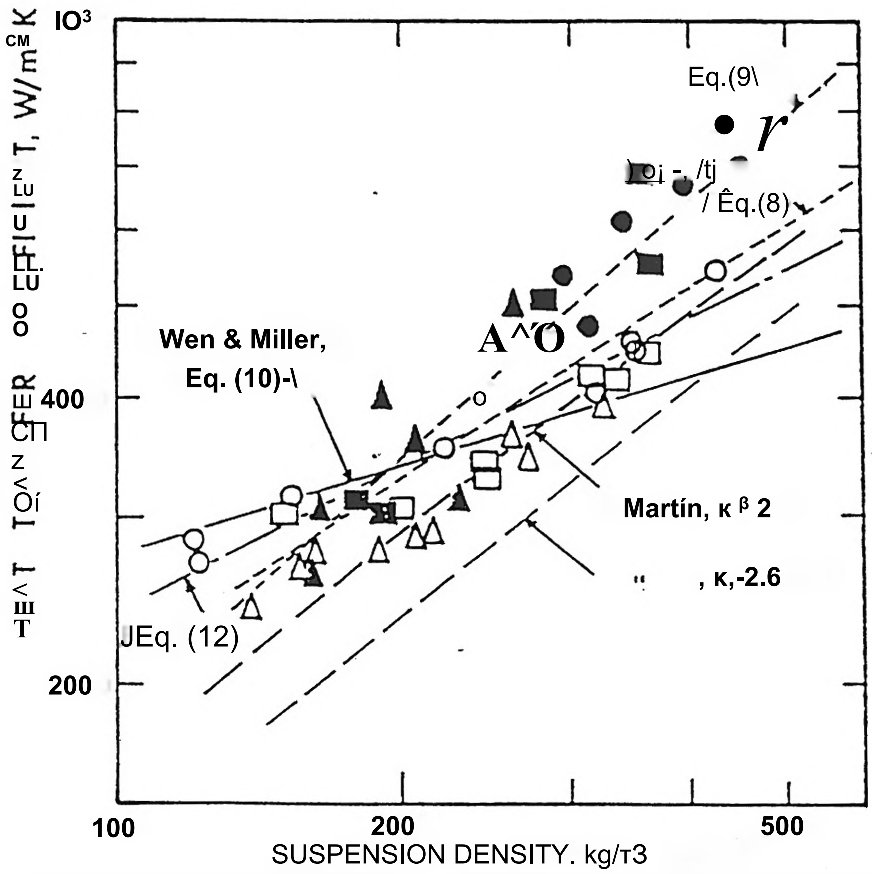


Figure 4- Heat transfer coefficients measured by Fraley et al (1983). Open symbols: cylinder along bed axis; blacked-in symbols: cylinder adjacent to column wall. Circles, squares and triangles correspond to U = 1.5-1.7 m/s, 1.8-2.0 m/s and > 2-0 m/s, respectively. (12)

coefficients to be somewhat higher along the axis than at r/R = 0.06, but generally slightly higher at r/R =0.33 than on the column axis; these measurements appear to be at lower solids concentrations than 173 kg/m^3 since the measured h values are in the range 170-38 $W/m^2 K$ based on bare tube areas.

figure 3 and 4 (12) and the results presented by Kobro and Brereton (27) indicate that there is little or not independent effect of superficial gas velocity on the suspension-to-surface heat transfer coefficients once the suspension density has been established. This suggests that the gas convective component of heat transfer is much less important than particle convective terms.

In this study, an experimental facility was constructed to study the heat transfer characteristics of a circulating fluidized bed. The experiment performed at the operating temperature range of a circulating fluidized bed combustor. This provided us with data that included the radiative heat transfer effects. Experimental equipment for Phase III, heat transfer from a vertical tube in a fluidized bed has been constructed and heat transfer coefficients will be reported in several months.

CHAPTER III

EXPERIMENTAL FACILITY FOR WALL-TO-BED HEAT TRANSFER

Experimental Set-Up And Program

An experimental facility was constructed to study the heat transfer characteristics in a circulating fluidized bed, as is shown in Figure 5.

The circulating fluidized bed unit is constructed of eight 4 inch high sections of 3" I.D. steel tubes with 6" I.D. water jackets. It is welded to a 4 foot solid accelerating section and a 4 foot extended freeboard section forming the fluidized bed unit. Heat transfer to the wall was measured by monitoring the water flow rate in the jacket and the inlet and outlet water temperatures. Bed temperatures are also measured at the inlet and outlet sections of each chamber. K-type thermocouples were used for the temperature measurement. Pressure taps were located between each section to measure the pressure drop across each section. These pressure taps are connected to a water manometer for measurement. A schematic diagram of the water-jacketed chamber showing the location of the thermocouples and pressure taps is shown in Figure 6. The pressure drop data are used to determine the solid suspension density in each section. Steady-state measurements are obtained when the pressure drop for each section is unchanged.

The recirculating behavior of the particles is achieved by connecting the top of the fluidized bed to a cyclone, a standpipe and a feeder device. The elutriated solid particles were collected by the cyclone. The design of the cyclone is shown in Figure 7. The collected particles move down a standpipe which was 1.5" in diameter. Two valves and a diverting tube are installed in the standpipe for measurement of solid flow rate. Solid particles are collected

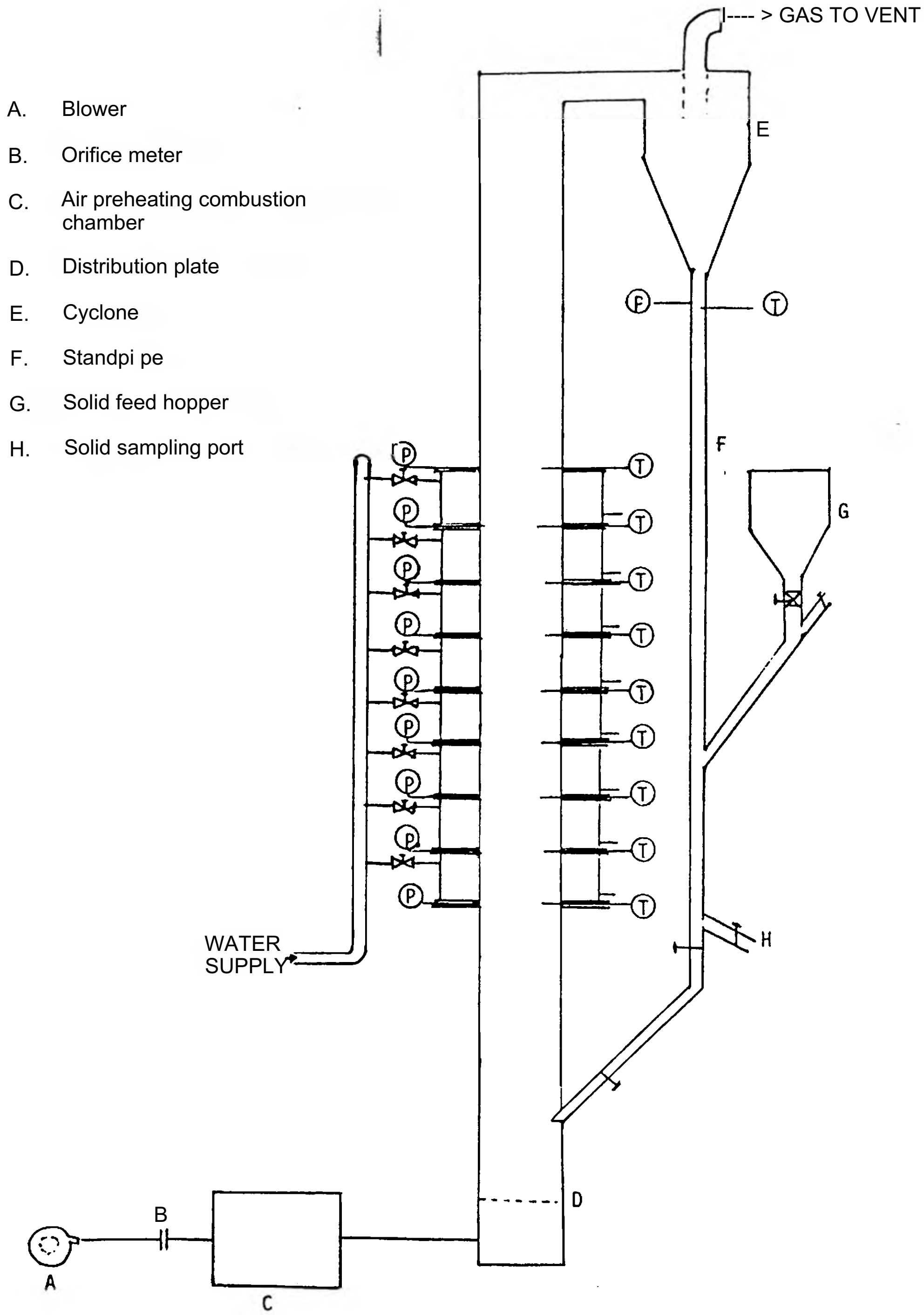


Figure 5. A schematic diagram of experimental set-up for heat transfer studies in circulating fluidized bed combustor (not to scale).

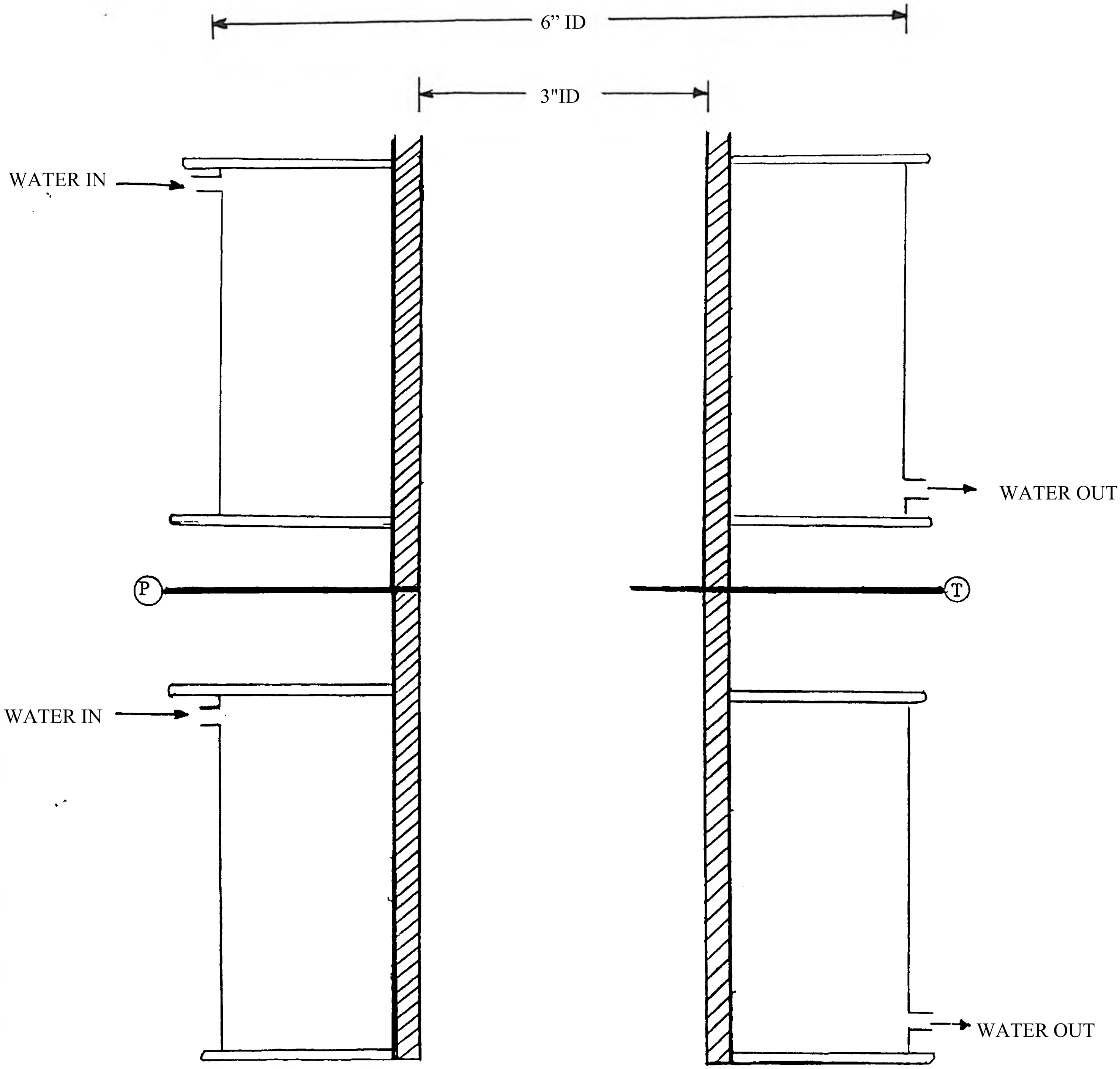
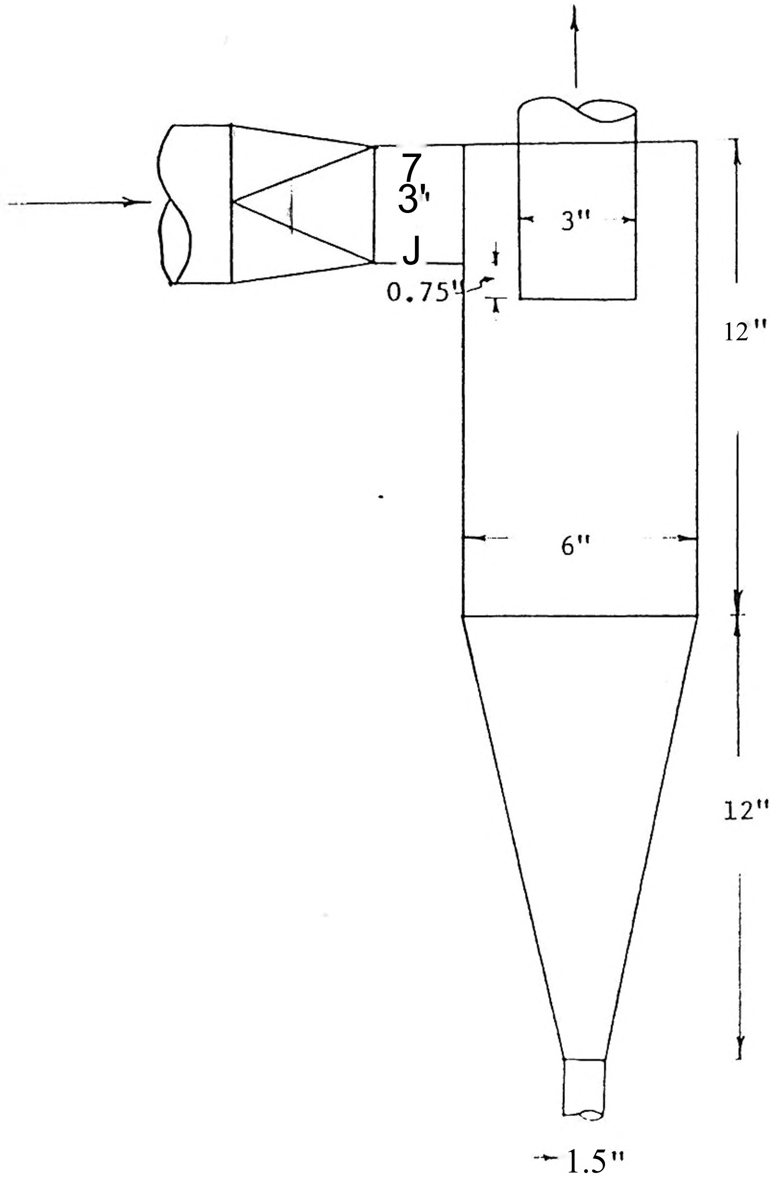


Figure 6. A schematic diagram of the water-jacketed chamber showing the location of thermocouple, T and pressure tap, P.



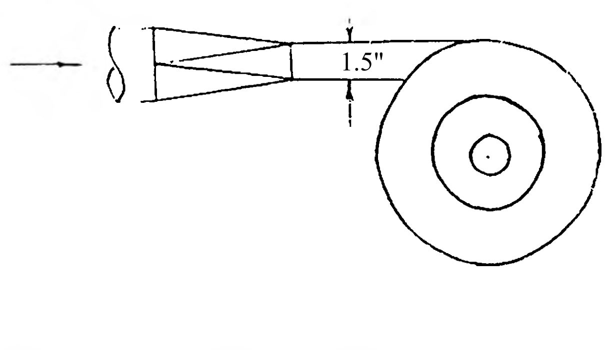


Figure 7. Design of Cyclone Separator

for a known short period of time and solid flow rate down the standpipe was determined. Bed material used were sand particles and are fed into the circulating bed via a solid feed hopper during start-up.

The solid particles moving down the standpipe are fed back into the fluidized bed unit by a pipe that leads into the bed forming a 45° with the horizonal. This orientation has shown to give best performance. Figure 8 illustrates the effect of the types of pipe exit on solid transfer rate (Trees, (28)). Small amounts of air are introduced into the 45° pipe to ensure smooth feed-back of solids into the fluidized bed. A sliding valve is used to regulate the solid recirculation flow rate. The solids are fed back to the bed about 3" above the distributor plate. The distribution plate is made up of a stainless steel perforated plate.

Fluidizing air was supplied by a rotary positive displacement blower. It is preheated to the desired temperature by passing through a gas-fired combustor. The burner design is a nozzle mix type which is a short flame high capacity unit designed for extremeley wide turndown ranges. Since gas and air are mixed only at the point of discharge, flashback is prevented. Exclusive stepped tunnel design creates very high turbulence and internal recirculation in the flame. This combustion tunnel shape produces pressures in the flame that are lower than the furnace pressure aiding flame retention. The design and dimension of the burner and the piping connection are shown in Figure 9 and 10 respectively. Propane is used as the fuel for the combustor. The air flow rate is measured by an orifice meter. The velocity and temperature of the air entering the circulating fluidized bed unit can be controlled at a predetermined value. Experimental runs were performed at different gas velocity, solid circulation flow rate bed temperature and particle type in order

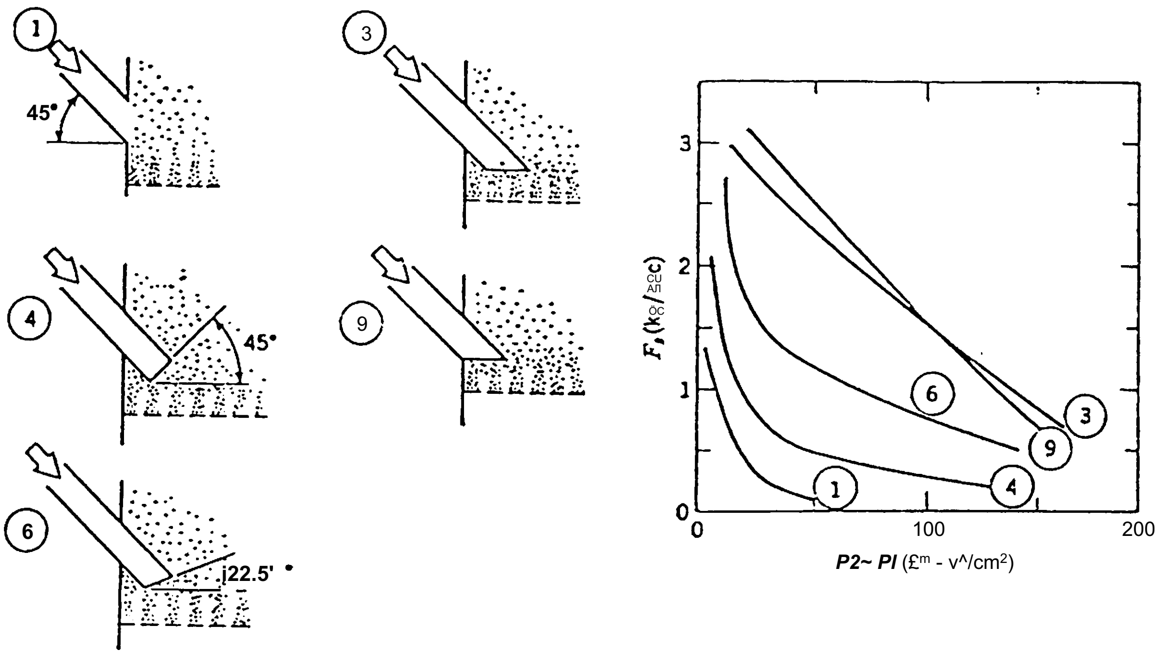


Figure 8. Effect of type of pipe exit on solid transfer rate (from Trees [28]). ¿ή = 10.5 cm, / = 305 cm, *B* = 45“.

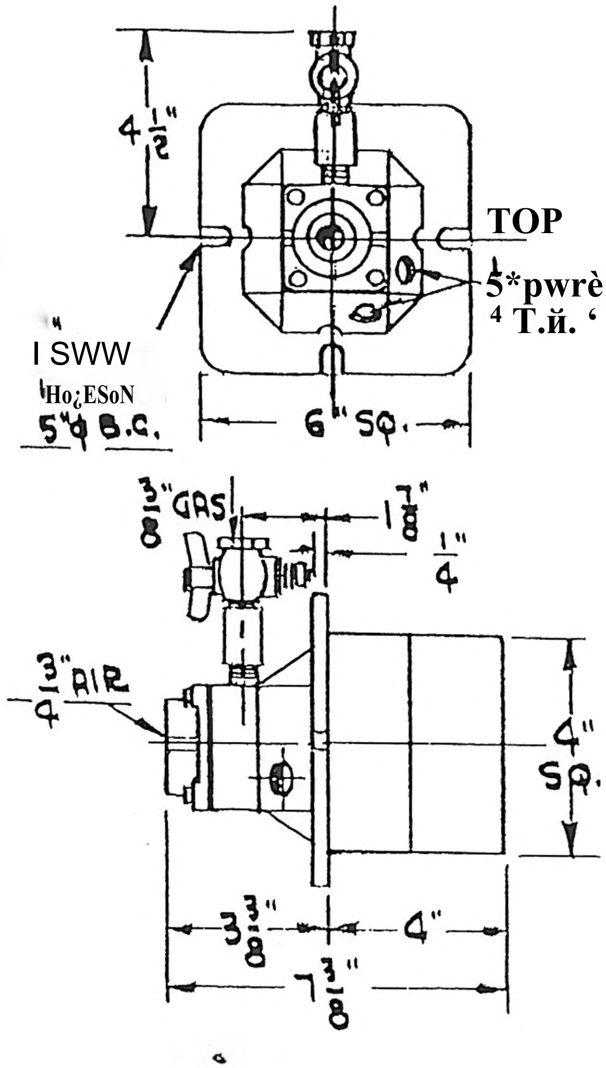


Figure 9. A schematic diagram of the nozzle mix burner.

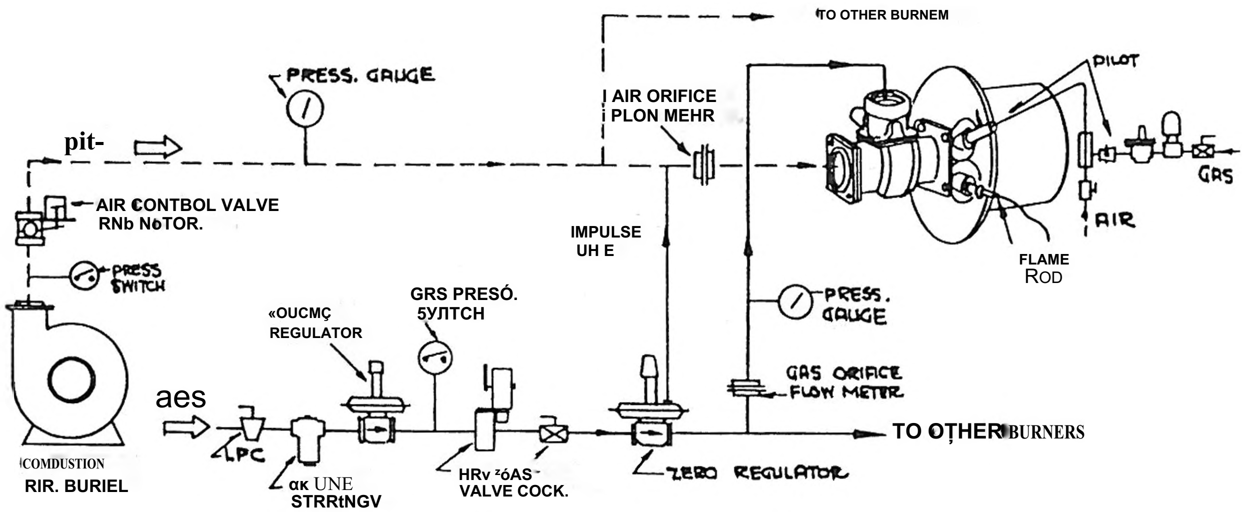


Figure 10. A schematic diagram of the piping for the burner system.

to provide enough data for the development of empirical correlations for heat transfer coefficient from the bed to the wall. The experimental operating conditions are listed in Table 2.

TABLE 2

Experimental Operating Conditions

Particle Type:	sand
Particle Diameter:	50 µm, 300 µm
Bed Temperature:	1000°K, 1173°K
Gas Velocity:	5 8 m/s

Solids Recirculation Rate: 0.3, 1, 3 times the base rate

CHAPTER IV

METHODOLOGY FOR DETERMINING THE BED-TO-WALL HEAT TRANSFER COEFFICIENT

4.1 Methodology

After experimental runs are performed, the measurements of the water inlet and outlet temperature as well as the bed temperature at the inlet and outlet of the water jacketed section can then be used to determine the heat transfer coefficient from the bed to the wall. The water flow rate and the operating conditions are also recorded. Assuming we have obtained the above information, a methodology is required to process the experimental information so as to obtain the heat transfer coefficient. This can be achieved by setting up energy balance over the combustion wall as shown in Figure 11, and described as follows.

Rate of heat transfer through the wall of inner pipe

$$q_a = (T_w - T_r)\, k \frac{2\pi}{\ln(\,)_i} \tag{1}$$

Rate of heat transfer from bed to wall

$$q = h\,(T_x - T)\,A + q_{dx} \tag{2}$$

then, for steady-state

$$(T_a - V)\frac{2\pi k l}{\ln(/)_i} = M\,(T_1 - V)\,A_i + q_r$$

where q_r = radiant heat transfer

$$= h_r\,(T_r - T_l)\,A_{ai}$$

$$= h_r\,(T_r - T_l)\,2\pi r_{ai} l$$

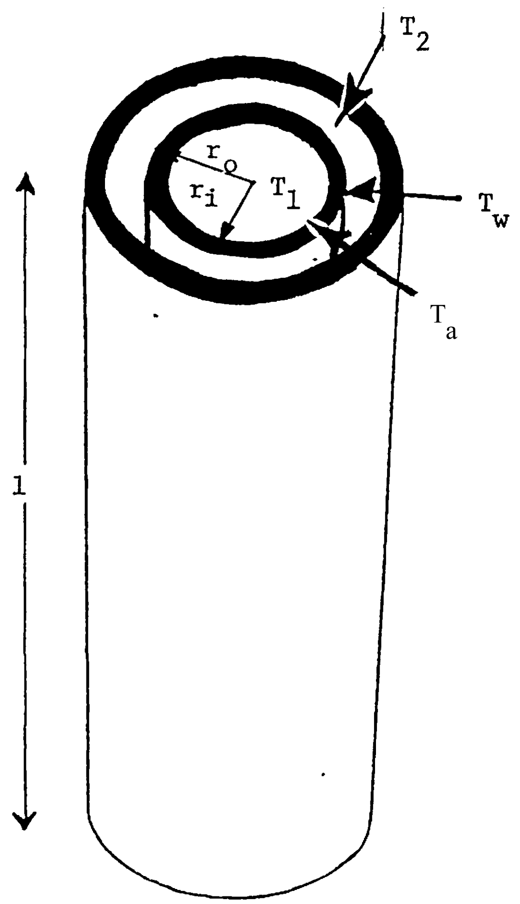


Figure 11. A schematic diagram of a water-jacketed chamber.

T_a = inner surface temperature of inner pipe

r^ = outer surface temperature of inner pipe

T_1 = bed temperature of CFB

r_i = inside radius of inner pipe

r_o = outside radius of inner pipe

h_i = inside heat transfer coefficient

For the rate of heat transfer from the outer surface of the inner pipe to water

$$q = h_o (T_w - T_2) A_o \tag{4}$$

combining with equation (1), we obtain

<Ta - V 2πk1 / ln(/) i , ho <TW - T2> Ao <5>

•

$A_o = 2\pi r_o 1$

T_2 = bulk temperature of water jacket

$$= \frac{T_{in,\,water} + T_{out,\,water}}{2}$$

h_o = water side heat transfer coefficient

The next step is to determine the water side heat transfer coefficient, h_o. Wiegard (1947) studied a large number of annular heat transfer data and recommended the following relationship for predicting h_o.

hd (—°Ê) k b = 0.023 (-M di°CpMd °.45 V b (-P-) k b (/-) 1 , b (6)

where $\dot{m}$ = mass flow rate

κ = thermal conductivity

Cp = heat capacity

μ = viscosity

these properties are evaluated at the bulk temperature also,

$$d_e = d_2 - d_1$$

d_2 = outside diameter of inner pipe

d_1 = inside diameter of outer pipe

This equation is valid for $Re > 10^4$.

Considering the heat transfer from the bed to the water in the jacket as a counter current heat exchanging operation, the overall heat transfer equation

$$q = U_o A_o \Delta T_{ln}$$

$$= \frac{\Delta T_{ln}}{\frac{1}{(h_i + h_r)A_i} + \frac{\ln(\frac{r_o}{r_i})}{2\pi k l} + \frac{1}{h_o A_o}} \qquad (7)$$

where $$\Delta T_{ln} = \frac{\Delta T_1 - \Delta T_2}{\ln(\frac{\Delta T_1}{\Delta T_2})}$$

The radiative heat transfer coefficient (h_r) can be calculated from

$$h_r = \frac{\varepsilon_a \varepsilon_d \sigma (T_i^4 - T_d^4) C}{[1 - (1 - \varepsilon_d)(1 - \varepsilon_a C)][T_i - T_d]} \qquad (8)$$

where $C = 1.5 \frac{r_i}{r_p}(1 - \varepsilon)$

r_p = radius of particle

ε = void fraction

ε_a = emissivity of inner pipe

ε_d = emissivity of particle

σ = Stefan-Boltzman constant

Heat transfer rate, q, can be determined from energy balance of water flowing into and out of the water jacket. The equation used is:

$$q = \dot{m} C_P (T_{out} - T_{in}) \qquad (9)$$

where T_{in} = inlet temperature of water

T_{Qut} = outlet temperature of water

4.2 Calculation Procedures

Using the above equations and the experimental data obtained, the heat transfer coefficient from the bed to the wall can be determined following the procedures listed below.

1. Use the inlet and the outlet temperature of water and water flow rate to calculate the total heat transfer rate, q, from equation (9).
2. Calculate h_Q from equation (6) at bulk temperature of water.
3. Guess the outer temperature (T_w) of inner pipe.
4. Calculate h^ from equation (8).
5. Using the values of h_Q, h^ and q, calculate h^ from equation (7).
6. From equation (1), calculate T_3
7. From equation (2), calculate T_{w_f}
8. Check the absolute value of $(T_w - T_{w_f})/T_w$. If the value is greater than the tolerance limit, repeat steps 4 to 8 by using the vàlues of T_{w_f}... Finally, we can obtain h_1.

A FORTRAN program has been written to accomplish the above trial-and-error procedures.

CHAPTER V

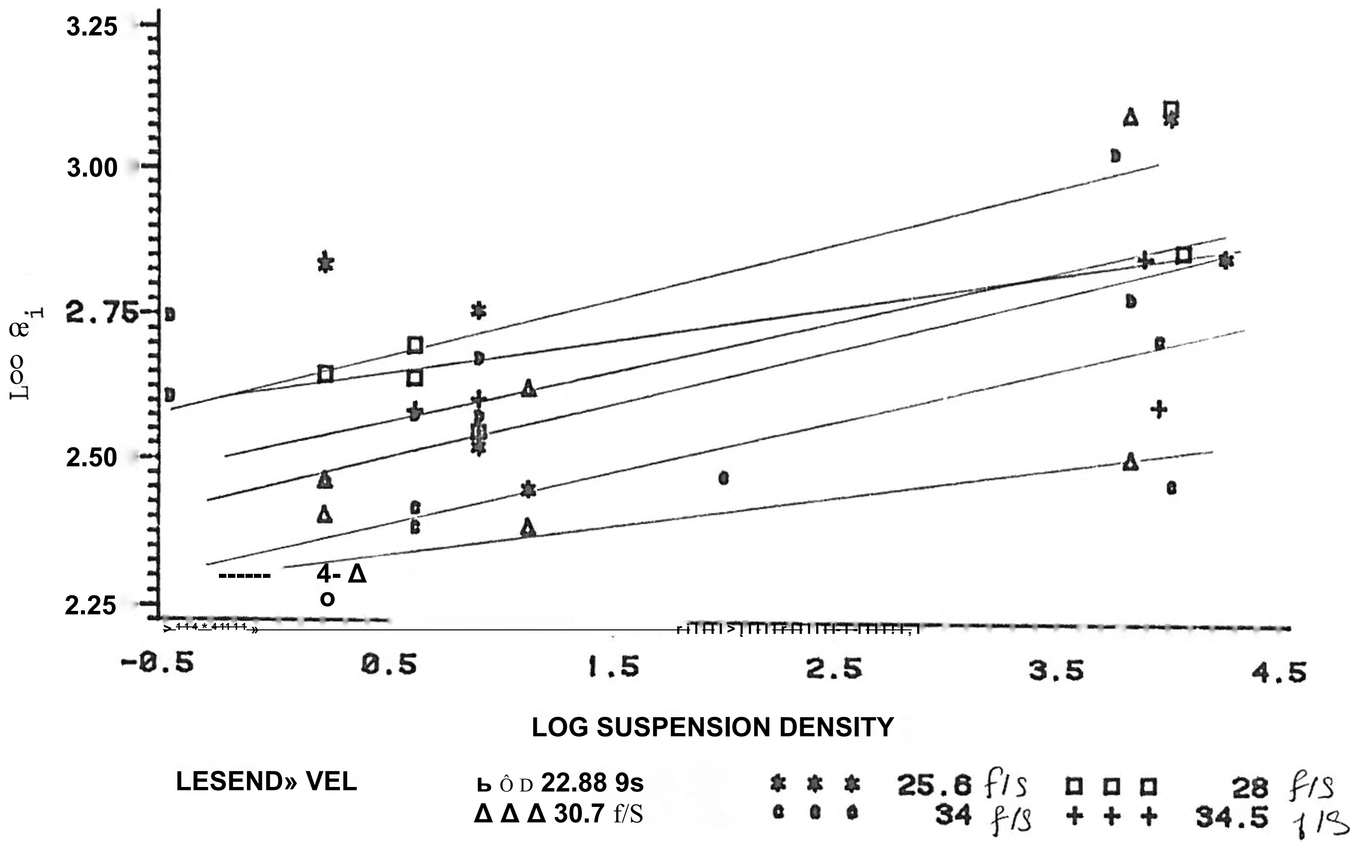
EXPERIMENTAL RESULTS FOR PHASE II

The bed side heat transfer coefficient (Btu/hr ft^2 °F) is shown versus suspension density on Figures 12-15. The heat transfer coefficient increases with suspension density. The heat transfer coefficient decreases as the particle size is increased as shown on Figure 15.

Figures 15-21 show the heat transfer coefficient versus the height from the distributor. The heat transfer coefficient decreases slightly with height from 4-5.5 feet from the distributor and increases from 5.5. to 7 feet from the distributor. The coefficient slightly increases with bed temperature indicating the radiation heat transfer is making a contribution.

The results are in general agreement with data presented in Chapter II.

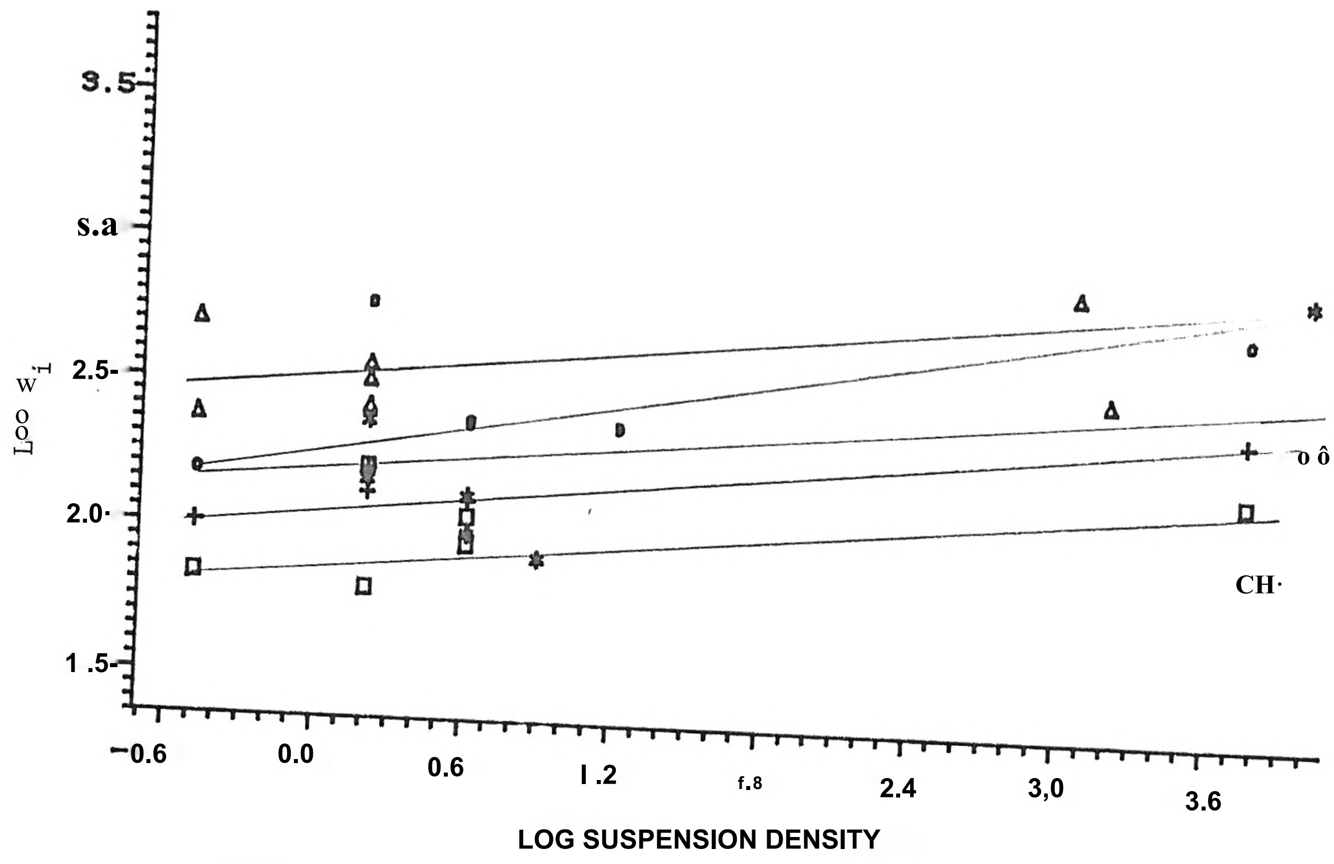
Fig. 12. PARTICLE SIZE 300 *jum* temp 600 F



H_i, $\frac{Btu}{hr\ ft\ °F^2}$

Suspension Density, lbm/ft^3

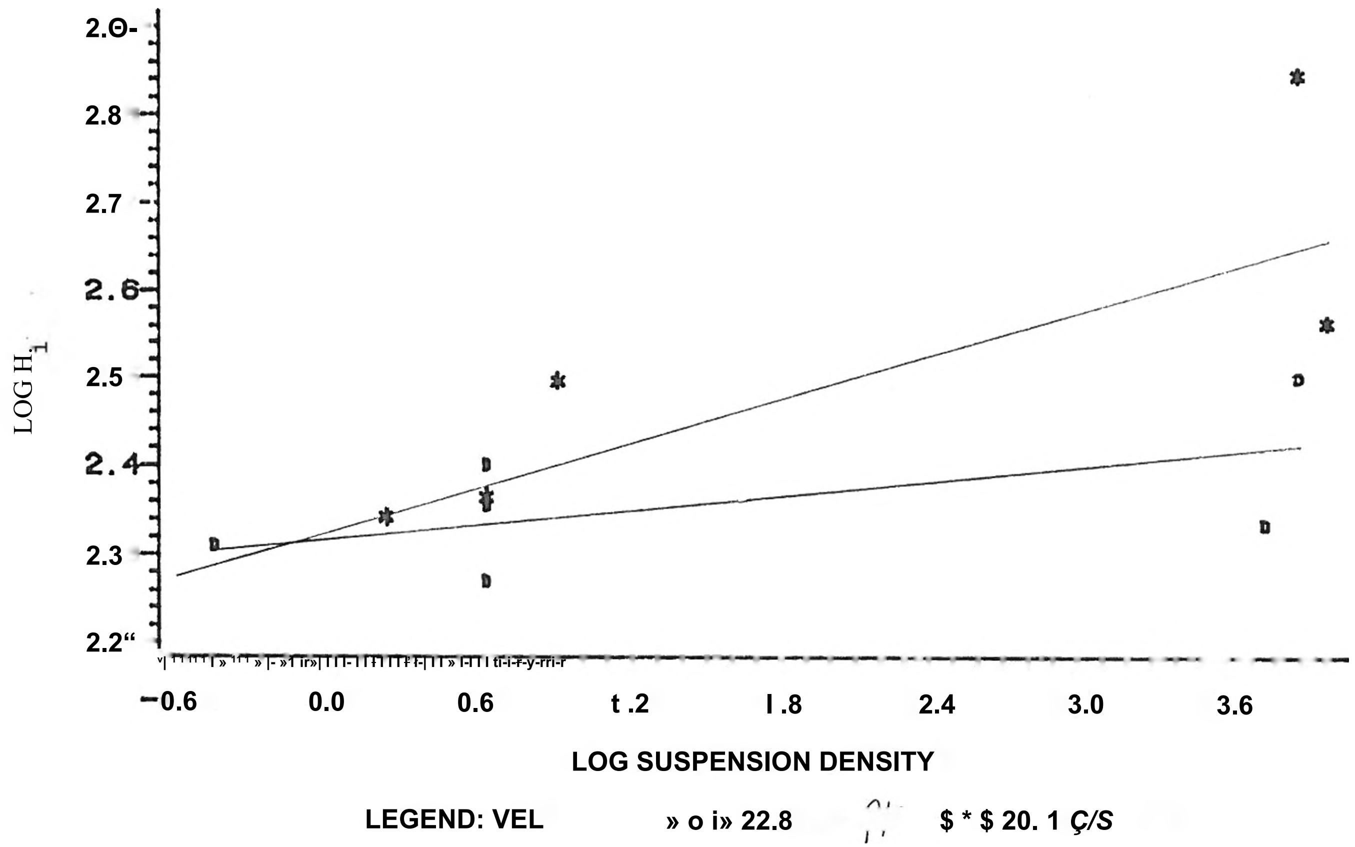
Fig. 13. PARTICLE SIZE 300yim temp 450 F



H_i , $\frac{\text{Btu}}{\text{hr ft °F}^2}$

Suspension Density, Ibm/ft^

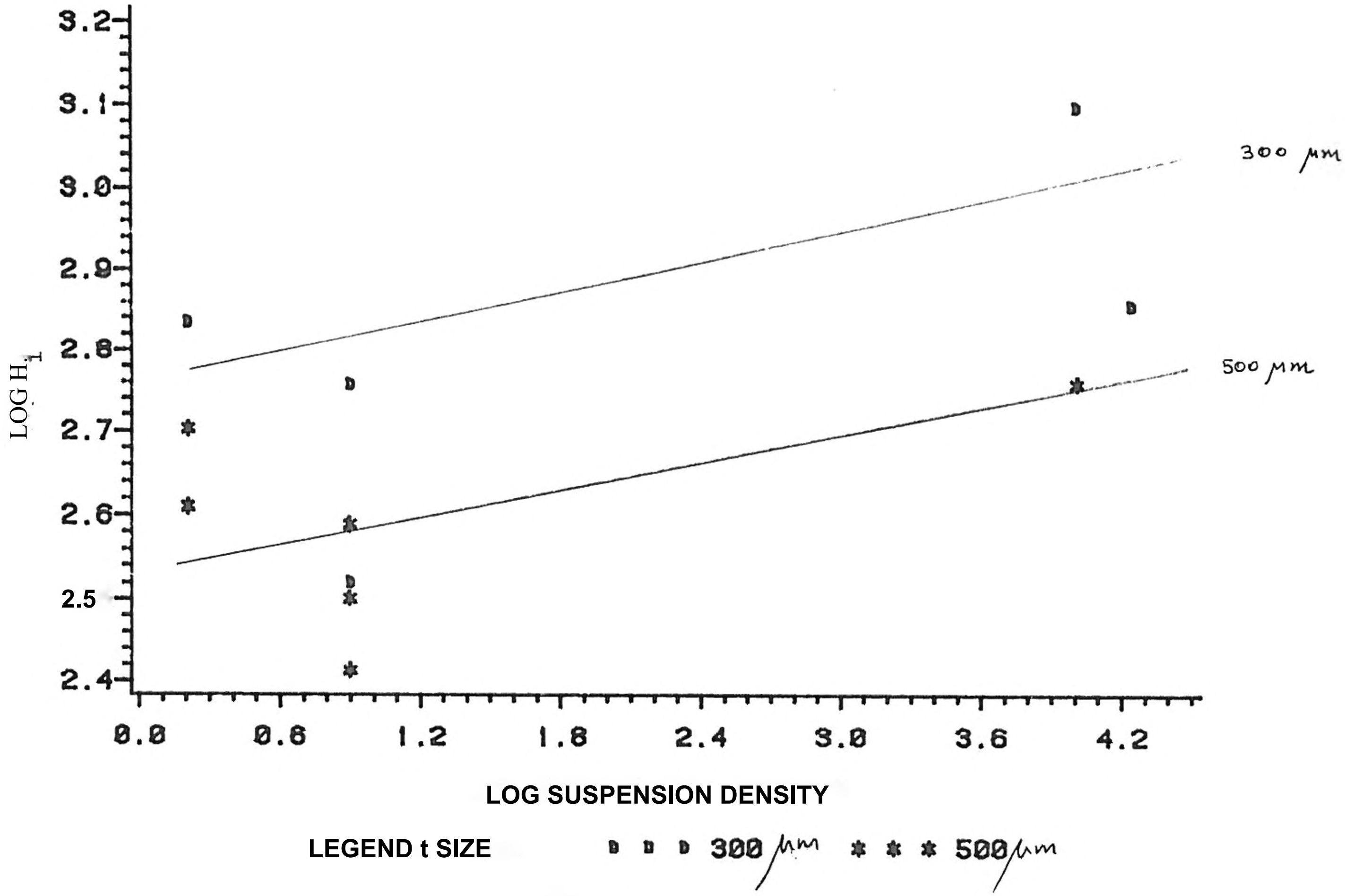
Fig. 14. PARTICLE SIZE 500 nm temp 430 F



H_i, $\frac{\text{Btu}}{\text{hr ft}^2\ °\text{F}}$

Suspension Density, lbm/ft^3

Fig. 15. PARTICLE SIZE 300 & 500 yjm temp 600 F



u Btu
hr ft^2°F

Suspension Density, 1bm/ft^3

Fig. 16. PARTICLE SIZE 300 nm temp 600 F

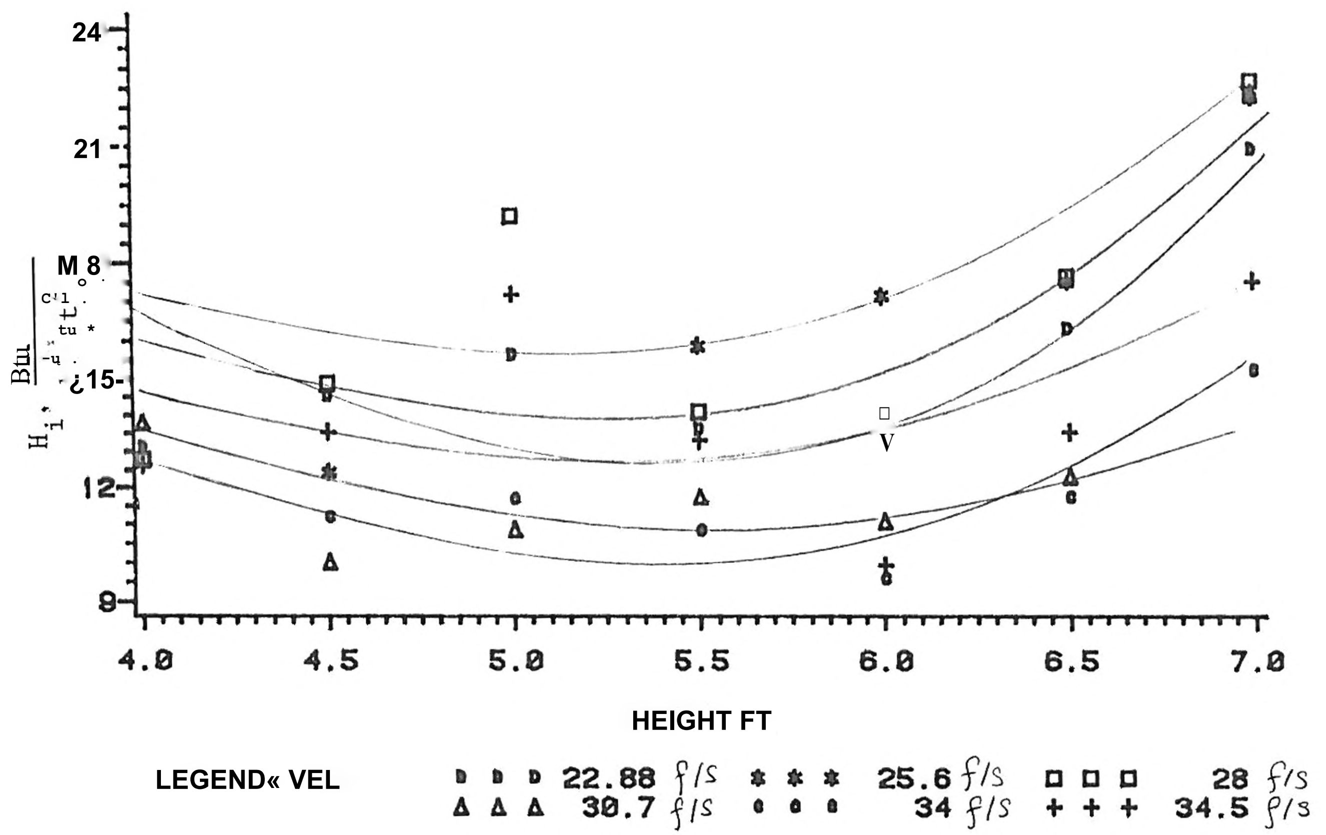


Fig. 17. PARTICLE SIZE 300 urn temp 550 F

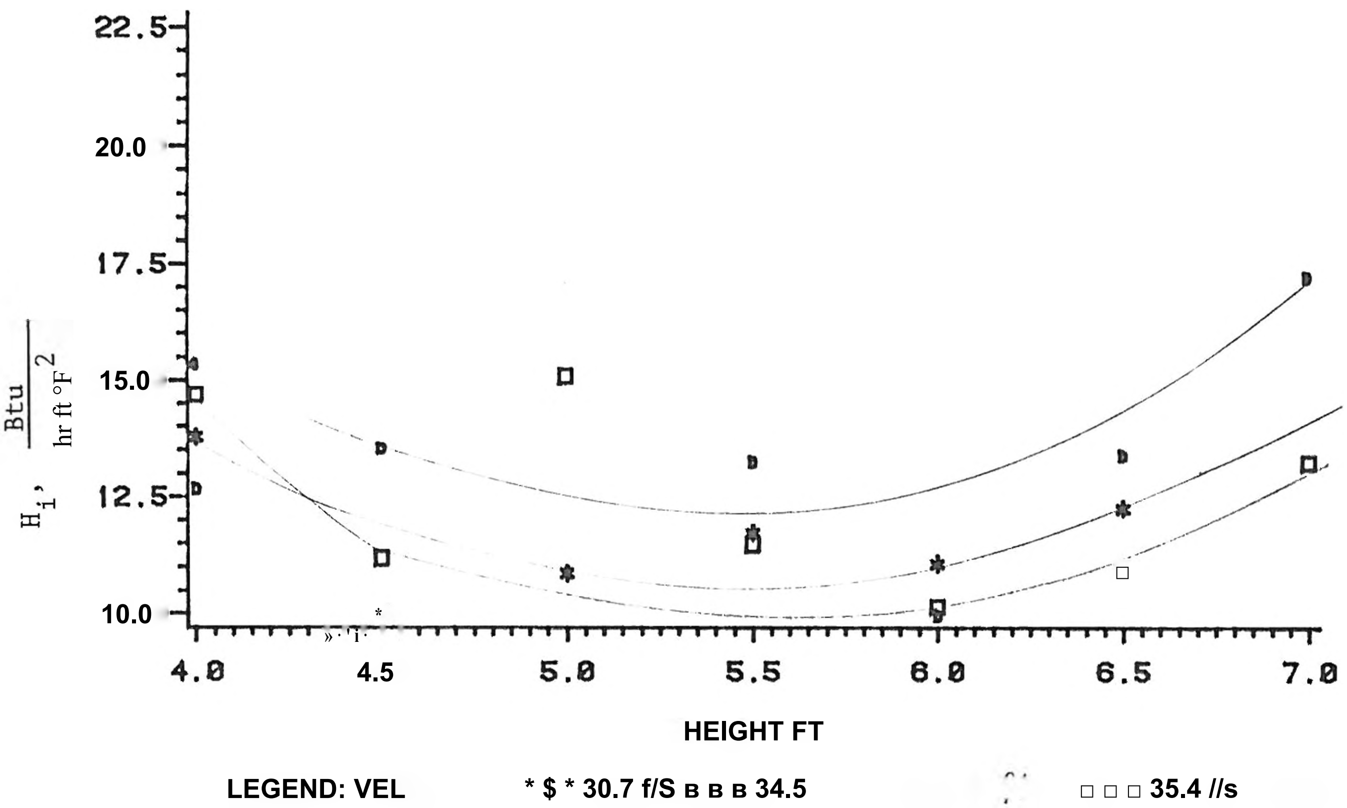


Fig. 18. PARTICLE SIZE 300 um temp 450 F

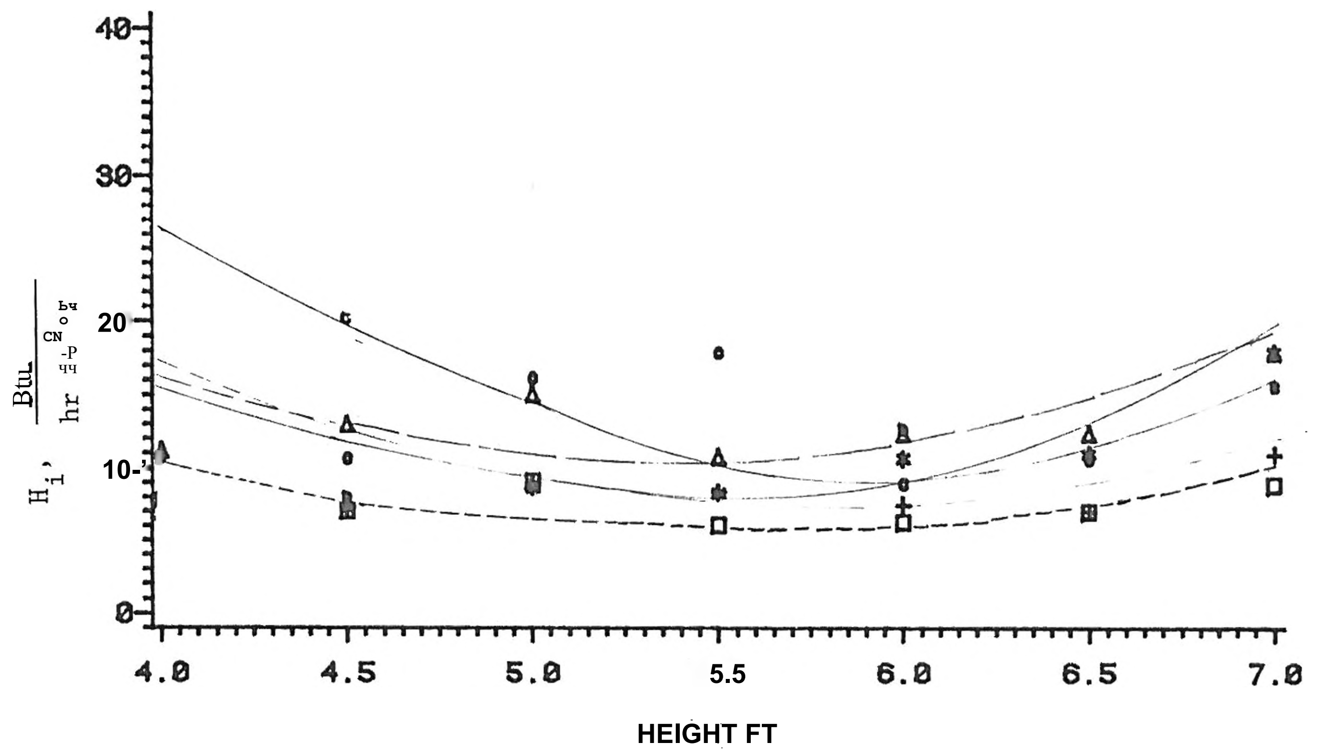


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Fig. 19. PARTICLE SIZE 50Ü шп temp 610 F

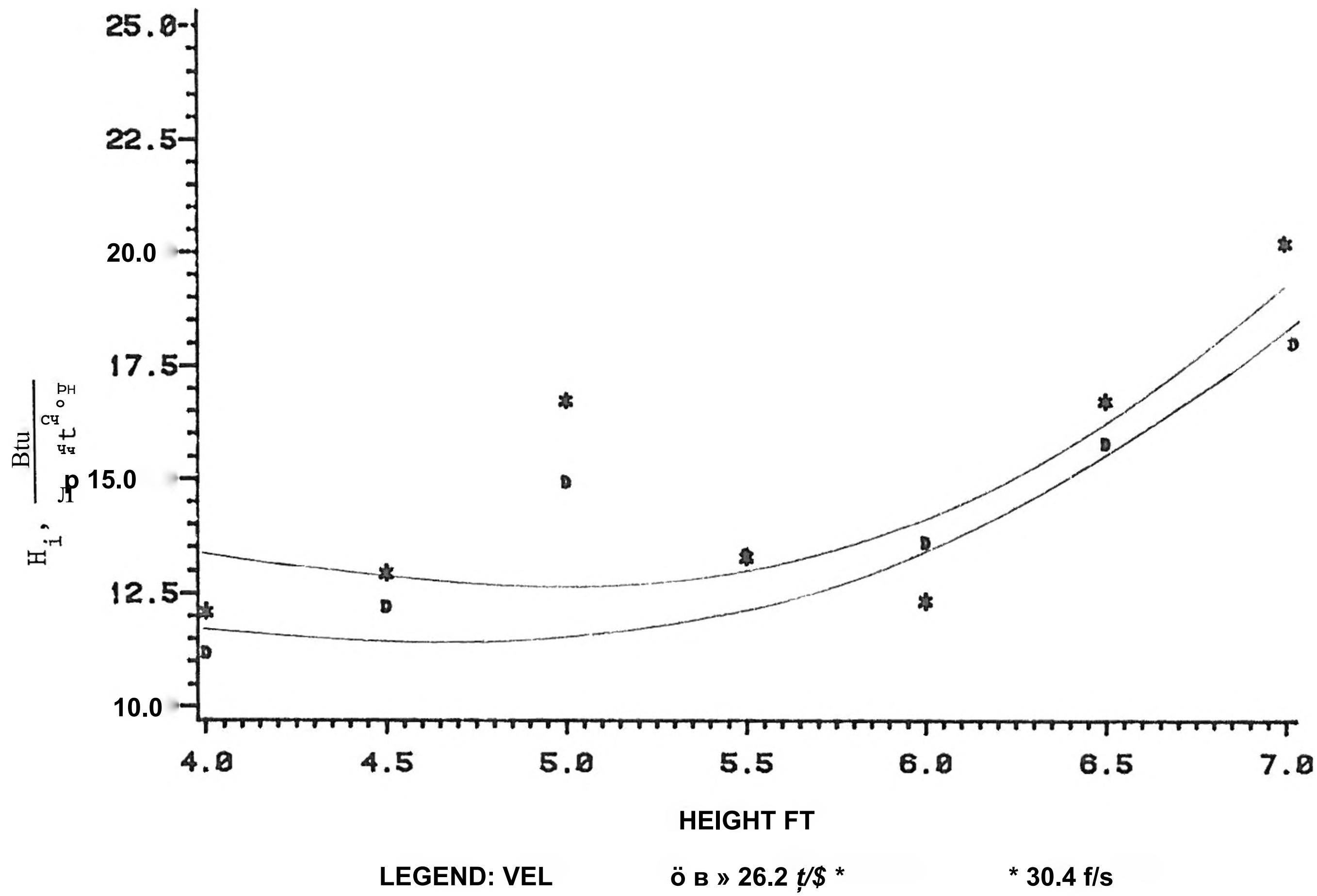

Fig. 20. PARTICLE SIZE 500 um temp 430 F

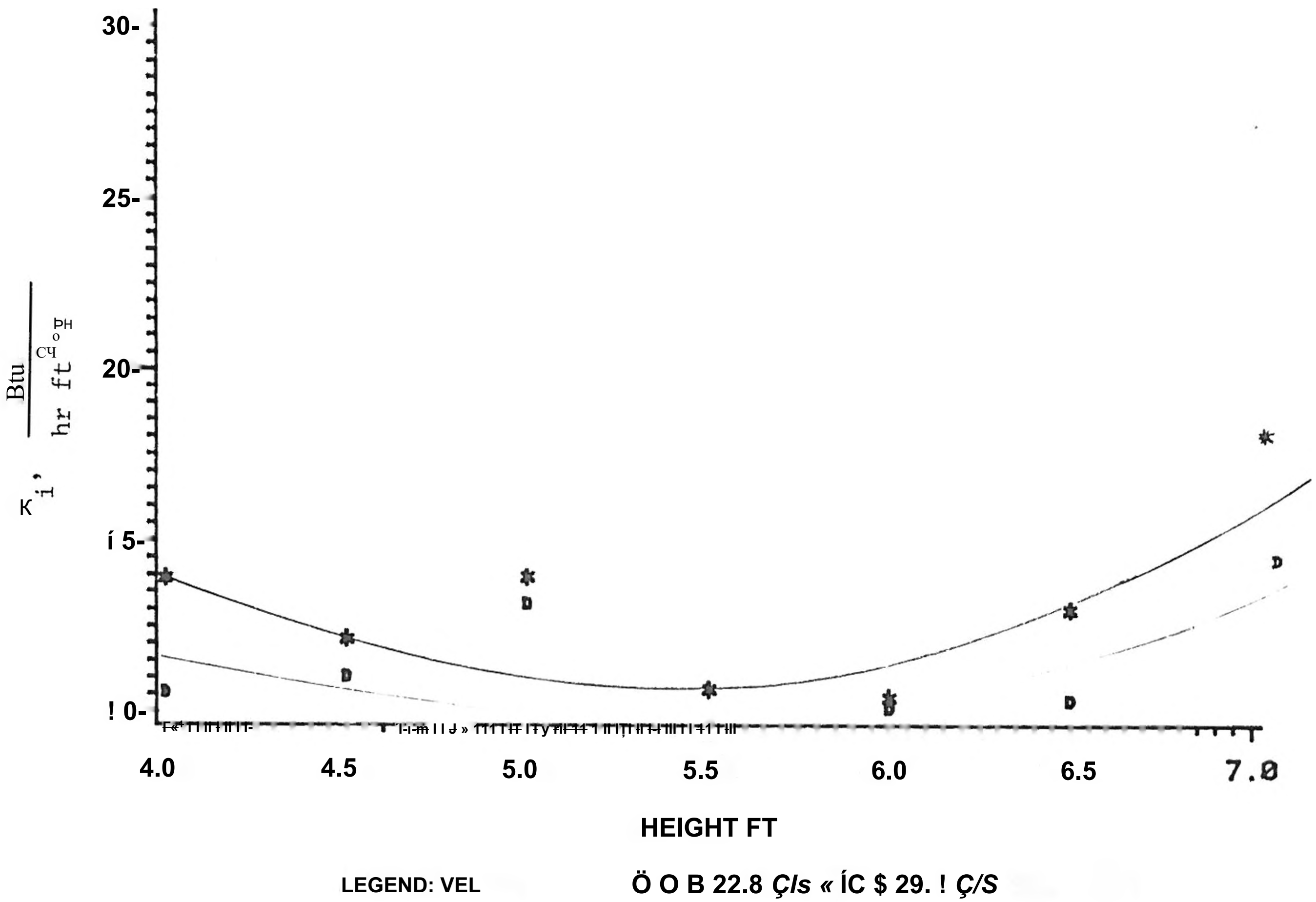

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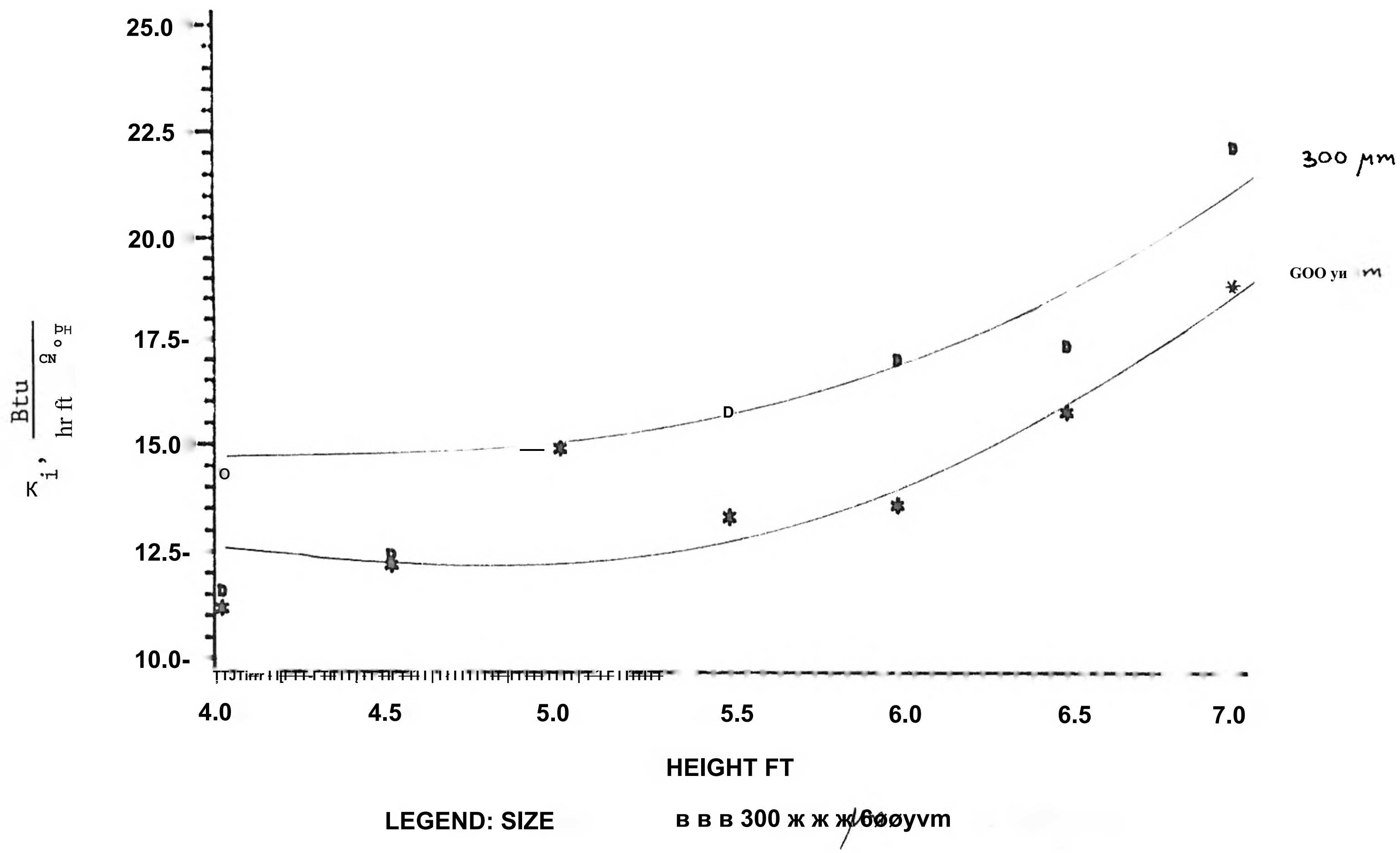
Fig. 21. PARTICLE SIZE 300 & 500 urn temp 600 F



CHAPTER VI

CONCLUSIONS

Circulating fluidized bed has the promise of becoming a viable option for burning Mississippi lignite both economically and in an environmentally acceptable manner. Since the design of heat transfer surfaces is critical in the overall equipment design and the operation and control of circulating fluidized beds, the present study investigates the heat transfer characteristics from the bed to the wall.

Specific objectives of this research project were:

1. To construct an experimental facility for studying the heat transfer characteristics in a circulating fluidized bed combustor.
2. To measure the bed-to-wall heat transfer coefficient as a function of gas velocity, solid concentration in the bed and solids recirculation rate in a circulating fluidized bed and to develop a correlation based on the experimental results.
3. To measure heat transfer coefficients for heat transfer from a vertical tube as a function of gas velocity, solid concentration in the bed and solids recirculation rate in a circulating fluidized bed.

Phase I of the project, which is the construction of the experimental facility, has been completed. Phase II of the project, which involves the collection of experimental data is complete. Phase III of this project is continuing this year without MMRI support. In this report, the design and construction of the experimental facility and the methodology used to extract the heat transfer coefficient from the data and experimental results of Phase II have been presented.

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