342 10-23-78
10

HCP/T2471-24

U.S. Department of Energy

September 1978

MASTER

Assistant Secretary for Energy Technology
Division of Fossil Fuel Utilization
Under Contract No. EX-76-C-01-2471

DR. 628

Industrial Application Fluidized Bed Combustion Category III Indirect Fired Heaters

Quarterly Technical Report No. 8 April 1—June 30, 1978

Available from:

National Technical Information Service (NTIS)
U.S. Department of Commerce
5285 Port Royal Road
Springfield, Virginia 22161

Price: Printed copy:
Microfiche: $3.00

HCP/T2471-24
Dist. Category UC-90e

U.S. Department of Energy **September 1978**

Assistant Secretary for Energy Technology
Division of Fossil Fuel Utilization
Washington, D.C. 20545

Under Contract No. EX-76-C-01-2471

Industrial Application Fluidized Bed Combustion Category III Indirect Field Heaters

Quarterly Technical Report No. 8 April 1—June 30, 1978

Prepared by
Exxon Research and Engineering Company
Engineering Technology Department
P.O. Box 101
Florham Park, N.J. 07932

NOTE: This document was prepared for the Energy Research and Development Administration (ERDA) prior to the activation of the Department of Energy (DOE) by the Energy Reorganization Act. Therefore, wherever ERDA is mentioned, its functions have been transferred to DOE.

TABLE OF CONTENTS

LIST OF FIGURES

ABSTRACT

A program is underway to evaluate the technical and economic potential for the application of fluidized bed combustion to refinery and petrochemical plant indirect fired process heaters. The strategy of the program is to build on 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 complementary 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, including bed mixing and density measurement, have been completed on six alternative tube bundle configurations ranging from 2-inch to 6-inch diameter tubes arranged 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. Finally, evaluations have been made on the effect of altering the tube-to-grid dimensions and of operating with limestone beds of different particle size distributions.

A Process Stream Coking Test Unit has been commissioned and is being used to study the parameters affecting coke laydown on the internal surfaces of hydrocarbon containing tubes under conditions of high temperature and heat transfer rate. Two test runs have been completed.

Design, procurement and early construction activities are underway for the third laboratory facility which will be a "hot" coal fired fluidized bed combustor. The component parts of this facility are described in some detail in this report. The 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 areas 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 complementary programs in the boiler area will be used in the overall technical and economic assessment of potential fluidized bed process heater applications.

The two principal 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 have on fluidization characteristics and the definition of the optimum or acceptable configuration for a tube bundle immersed within a fluidized bed must be determined.

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 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 revised Milestone Plan and Management Report included here as Figure 1.

The first major laboratory task, namely, the Two-Dimensional Flow Visualization Study, has been completed. This study evaluated the effect on fluidization and heat transfer when an array of relatively large diameter tubes was immersed in a fluidized bed of limestone. Alternative configurations of tubes up to 6 inches in diameter arranged on 2 to 4 tube diameter center-to-center spacing were investigated. The studies defined the range of acceptable tube bundle configurations that might be used in commercial process heater applications. Engineering data on fluidization parameters and conductive/convective heat transfer patterns were obtained.

The Process Stream Coking Test Unit is now in operation and two test runs have been completed. Both runs were made at relatively severe operating conditions of 45-60 kBtu/hr ft^2 heat flux and a nominal 650°F coil inlet temperature. The first run was aborted by a failure in the high pressure booster pump. The short test duration and aborptive shutdown resulted in the data from the run being of limited value and not suitable for definitive interpretation. However, the run did provide valuable guidance for the planning of subsequent runs.

Test Run No. 2 was successfully completed and a preliminary assessment of the results are discussed in this Report. The test very successfully bracketed the range of coke initiation conditions at the 60 kBtu/hr ft^2 heat flux level.

The third laboratory test unit, namely, the Heat Flux Test Unit, is well into the procurement and early construction phase. All major components are on order and some materials have begun arriving onsite. A first quarter 1979 construction completion is anticipated. Design information on the major facilities that will be included in this test unit are described in some detail in this Report.

3. Discussion of Technical Progress

3.1 Two-Dimensional Flow Visualization Studies

3.1.1 Background Information

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

Tube bundles were immersed in the bed and the effect on fluidization of these relatively large tubes was 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 location and bed pressure drop were also examined although these were of secondary interest since they are being investigated by boiler oriented programs.

3.1.2 Status of Work

All work planned under this Task of the Program has been completed and reported. Briefly summarized, the work successfully defined the configurations of 4" and 6" diameter tubes that would be suitable for use in Process Heaters. It defined several limiting parameters on fluidization velocity, tube spacing and combustor configuration that would have to be adhered to in order to assure satisfactory performance. Useful conductive/convective heat transfer data were obtained which are necessary to initiate full scale design studies. These same data also suggest effective means of operation for load following on commercial units by altering expanded bed depth and explosing more or less tubes to the fully fluidized zone of the bed. Other engineering parameters and general observations on fluidization and heat transfer performance in the presence of 4" and 6" diameter tubes were made which will be used in the later parts of this program.

All task data have been reported in Quarterly Technical Report Nos. 5, 6 and 7. The interested reader is referred to these reports for further details.

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, mass velocity and inside film temperature.

The test facility that has been built to carry out these studies has been installed at Exxon's Bayway Refinery, Linden, N.J. The unit consists of four heat exchangers, each heated by an electric radiant heater. Each exchanger is a single 0.6 inch I.D. x 9 ft. long (heated length) stainless steel tube. The basic scheme is to pass a stream of virgin crude oil through each of the four exchangers. Total unit throughput is approximately 900 Bbl/Day.

Each exchanger is exposed to a different combination of process conditions (mass flow, bulk temperature and heat flux) and each is 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 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 of Work

In addition to the hot calibration run discussed in Quarterly Technical Report No. 7, two test runs have been completed. The Test Run No. 1 was of relatively short duration and had to be aborted due to a pump failure. The results of this test are of limited value.

Run No. 2 was successfully completed on June 28th and the results are now being analyzed. A narrative description of each of these tests is given in the following paragraphs.

3.2.3 Test Run No. 1

Test Run No. 1 was initiated on March 21st. Test conditions were a nominal heat flux rate of 48 k Btu/hr ft^2 with a coil inlet temperature to the exchangers of 640°F. It was planned to operate the exchangers at the following mass velocities:

Exchanger	Mass Velocity lb m/ft^2 sec
A	600
B	450
C	300
D	150

However, during startup the safety valve on Heater C (300 lb m/ft^2 sec) released prematurely and failed to reseat so this exchanger had to be shut down. In order to obtain important data at this mass velocity it was decided to reduce the mass velocity on Heater B from 450 to 300. Heater A and D were run at their planned mass velocities of 600 and 150 lb m/ft^2 sec respectively.

About 42 hours into the run, the test had to be abruptly terminated when the leakage rate of the buffer fluid on the high pressure booster pump became excessive. The pump supplier, Sundyne, have determined that the cast impeller housing contains a probable manufacturing defect and have now provided a replacement casting under the pump warranty.

Meanwhile, the test sections from each of the exchangers were removed and examined for coke deposition. Very limited significance can be attached to the amount and pattern of coke laydown observed since the test was in progress only 42 hours (although the conditions of exposure were quite severe). However, some interesting observations were made.

Firstly, the coke laydown was clearly heavier in the exchanger tubes run at the lower mass velocities. Exchanger tube A which had operated at 600 lb m/ft^2 sec was essentially clean over its entire length. Exchanger tube B (300 lb m/ft^2 sec) had a light layer of relatively hard coke affixed to the tube wall that measured from nil to approximately .04 inches in thickness measured respectively from the inlet to the outlet end of the tube.

Exchanger tube D which had operated at the 150 lb m/ft^2 sec mass velocity had a much heavier layer of dense coke adhered to the tube wall that measured from about .01 inch to .07 inches from inlet to outlet of the tube.

In addition to the layer of relatively dense coke deposited on the tube walls, each of the exchangers also contained some deposits of very loose porous coke which spalled or fell from the tubes during cutting and examination. It is believed that this coke was formed during the shutdown from crude oil that remained in the tubes after flow was stopped and before the residual heat of the exchanger cabinet had been dissipated. This occurred primarily because of the unplanned and semi-emergency nature of the shutdown. On subsequent tests this problem will be avoided by prompt and complete nitrogen purging of individual exchangers as a part of the normal shutdown sequence.

Because of the short duration of this test run and the abortive nature of the shutdown, no definitive conclusions will be drawn from the data obtained on Test Run No. 1. This test was rescheduled to be repeated very early in the continuing test matrix. However, the run did serve a useful purpose in confirming that coke deposition on the inside of the tube walls could be initiated under these simulated process conditions and the rate of coke laydown appears to have been bracketed by the range of conditions imposed on the three exchangers. Moreover, the rate of coke deposition was clearly indicated to increase with increasing bulk temperature and with decreasing mass velocity.

3.2.4 Test Run No. 2

Following the completion of the aborted Test Run No. 1, the unit was out of service for five weeks because of a scheduled maintenance turnaround of the Bayway No. 7 Atmospheric Distillation Unit. This unit is the source of crude oil feed for the Coking Test Unit. The downtime period was used to replace the faulty high pressure pump impeller housing and to complete other miscellaneous maintenance on the unit.

Crude oil feed was again available to the unit in mid-May and testing was resumed. For Test Run No. 2, planned conditions were relatively severe with a target coil inlet temperature of 650°F and a heat flux of 60 k Btu/hr ft^2. Mass velocities for the four exchangers A through D were again to 600, 450, 300 and 150 lb m/ft^2 sec respectively.

However, several mechanical and instrument problems developed during the restart of the unit which delayed the initiation of Test Run No. 2. A circuit breaker on "D" heater burned out due to a faulty wiring connection and had to be replaced. Four safety valves on the unit were found to have broken springs. These failures were traced to the fact that the springs were incorrectly supplied with a material incompatible with hot crude oil containing sulfur compounds. Replacement was made with Inconel material to provide an added margin of performance assurance.

A number of erratic and extraneous alarm signals were also being generated by the Leeds and Northrup Transcan 1000 Data Logger and Alarm Processor. This caused nuisance shutdowns and erroneous data to be logged. The L&N service representative could not locate the problem so the alarm processor was removed and replaced with another unit.

After considerable time was spent to define and resolve each of these miscellaneous problems, Test Run No. 2 was finally initiated on June 20th. The firing mechanism for Robicon Power controller on "B" exchanger was inoperable at this time and it was expected to remain out of service for an additional two weeks before parts replacement. The test was therefore initiated with only exchangers "A", "C" and "D" in service. Target conditions for these three exchangers were as identified above and covered the full planned test range of 600 to 150 lb m/ft^2 sec mass velocity.

After preoxidizing the tubes to increased surface emissivity (see following section 3.2.5) the exchangers were lined out at near target test conditions. A typical set of running data plots for exchanger "D" are included (Figures 3 thru 8) to illustrate the time variance patterns of the various monitored parameters.

Figure 3 shows the heat flux vs. time for exchanger D. Target flux was 60 k Btu/hr ft^2. The actual heat flux varied from 54,495 to 60,473 Btu/hr ft^2 from hour four to forty. The average heat flux was 58,444 Btu/hr ft^2.

Figure 4 shows a plot of the mass velocity which was targeted at 150 lb m/ft^2 sec and actually averaged 143 lb m/ft^2 sec with a variation of $\pm$ 2 lb m/ft^2 sec.

Figure 5 is a plot of the change in thermal resistance at thermocouple position 7 located midway along the length of the exchanger tube. The thermal resistance is a calculated value based on measured tube wall temperature, flux rate and bulk fluid temperature (see Figure 6). Initially, the thermal resistance reflects the resistance of metal between the thermocouple node and the inside tube wall plus the fluid film resistance. As time progresses and coke is deposited on the tube wall the thermal resistance is increased by the coke layer. At constant heat flux and bulk temperature an increase in thermal resistance drives the thermocouple readings up. In actual practice tube wall temperatures are used to monitor internal coking rate. However, the apparent thermal resistance is a more indicative measure of coking since it accounts for the minor variations in mass velocity and bulk temperature. Similar thermal resistance plots were developed for all nine thermocouples along the length of the tube.

Figure 7 is a plot of measured pressure drop in the exchanger "D" tube and shows a small but perceptible increase as the run progressed.

As can be observed from these typical data, exchanger "D" showed a gradual increase in thermal resistance over the run period. After 40 hours of relatively steady conditions the tube metal temperature had increased as much as 250°F toward the outlet end of the tube D, indicating a significant amount of internal coke laydown. Therefore, this exchanger was shut down and prepared for internal inspection.

Meanwhile, during the same period of time, exchanger A and C which were operating at essentially identical bulk temperatures and heat flux but at mass velocities of 600 and 300 lb m/ft^2 sec respectively, were showing no indication of coking. So operation on these exchangers was continued.

After 93 hours of operation "A" heater lost electrical power. When the electric heater cooled, flow to heater A was stopped. It was later determined that a circuit breaker on heater A had burned out. This was the second such failure; the first having occurred on "D" heater on Test 1.

At the time of shutdown "A" heater exhibited absolutely no indication of coking (see Figure 8). In comparison to heater D, it could be concluded that "A" heater at 600 lb m/ft^2 sec was operating in a nil or minimal coking regime. Therefore the test had attained its primary objective even though it was prematurely terminated and would not be resumed. Heater "C" continued to operate and was showing definite indication of coke laydown.

Heater "C" was continued in operation for a total of 132 hours until it was intentionally shut down to terminate Test 2. At the end of run "C" heater maximum tube metal temperatures had increased 170°F over start of run condition. Therefore, on a relative scale, it was very conclusive that the mass velocity (and therefore presumably the inside film temperature) plays a dominate role in rate of coke laydown. "D" heater (150 lb m/ft^2 sec) showed significant coking in only 40 hours. "C" heater (300 lb m/ft^2 sec) produced a lesser but definite indication of coking in 130 hours. "A" heater (600 lb m/ft^2 sec) ran only 93 hours but displayed no indication of coking in this period.

The individual exchanger tubes are now being disected for internal inspection and physical examination. Initial observations confirm that the patterns and extent of internal coking are consistent with monitored temperature measurement taken during the test. A more detailed analyses and comparison of these and subsequent tests will be presented in later Program Reports.

3.2.5 Effects of Exchanger Tube Emissivity Defined

During Test Run No. 1 it was noted that the radiant heater wire temperatures in all four exchangers were running significantly higher than predicted or than observed during the earlier hot calibration run. At a nominal 48,000 Btu/hr ft^2 flux, the wire temperatures were in excess of 2000°F which is about 400°F higher than expected. Wire life is severely reduced as the wire temperature increases. For example, predicted life at 2280°F is about one fourth that at 2000°F. Therefore, operating the unit at the high fluxes scheduled later in the program would have threatened to cause premature failure of the heaters unless the wire temperature could be reduced.

From an analysis of the Test No. 1 data it was calculated that the tube emissivity must have been about .45. This compares to an emissivity of .8 used in the pretest design calculations and confirmed during the hot calibration run.

A check of available data showed that the emissivity of stainless steel depends greatly on surface conditions. A bright polished surface can have an emissivity as low as .3 but by forming an oxide layer the emissivity can be increased to .8 or .9.

A review of past unit operations strongly suggested this difference in surface emissivity to be the cause of the high wire temperature during Test Run No. 1. This was a relatively short run with new tubes recently installed in all heaters. In contrast, the tubes used during the hot calibration run had gone through many temperature cycles and had been operated at relatively high temperatures during the initial checkout and pre oil-in commissioning of the unit. These tubes, therefore, had a discolored and well oxidized surface.

Therefore, the startup procedure has been modified to include a "burn-in" step in which new tubes are heated in the exchanger to 1500°F for 4 hours to form a dark oxide layer. The effectiveness of this treatment was substantiated during Run No. 2.

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, orientatation and fluidization velocity.

The data to satisfy the requirements of this task are being 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 complementary ambient temperature studies, which are now completed, have defined the conductive/convective component of the heat transfer mechanism. By comparing results from the high temperature and ambient tests the radiation component will 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 Ambient Temperature Heat Flux Tests

The conductive/convective heat transfer measurements on single tubes and tube bundles were completed in conjunction with the Two-Dimensional Flow Visualization tests completed earlier in this program. Results have been reported in Quarterly Technical Report Nos. 5, 6 and 7.

Briefly summarized, the tests defined the peripheral patterns of heat input to horizontal tubes fully immersed in a fluidized bed of limestone as well as to tubes in the splash and freeboard zones above the bed. Three distinct regions of heat transfer were identified around the tubes; namely, the predominately gas shrouded underside of the tube, the dense layer or "cap" on top of the tube and the "surfline" or rapidly oscillating region of solids-to-tube contacting at the 45° and 335° region (0° or 360° region defined as the top of the tube).

It was also determined that for a wide blend particle size bed material, heat transfer performance is predominately determined by the finer particles in the blend and is relatively unaffected by the coarser particles.

3.3.3 High Temperature Heat Flux Tests

A laboratory facility is being constructed at the Exxon Engineering Center, Florham Park, N.J. in which high temperature heat flux tests will be carried out. A schematic flow sheet showing the principle components of the system is included as Figure 9. Elevation and Plan views indicating general equipment arrangement are shown in Figure 10 and 11, respectively.

The fluidized bed test chamber will be approximately 2 ft x 4 1/2 ft in plan for a total bed cross-sectional area slightly in excess of 9 ft^2. While the unit will be designed with the capability of burning coal, initial tests will be conducted without combustion in the bed. The hot fluidization gas will be produced by burning vaporized propane fuel in the 12 M Btu/hr. refractory lined precombustor. The hot flue gas will flow through a refractory distributor grid into the test chamber. In this way fuel distribution and combustion rate variables will be eliminated as an element of concern in the initial studies. The initial studies are intended to define the overall and peripheral heat transfer rates to tubes immersed in a "hot" fluidized bed of limestone. In subsequent tests coal will be burned directly in the bed and other factors such as unit operability, startup and load following will be demonstrated.

For all tests water cooled tubes will be used to remove the heat from the fluidized bed. The tube bundle configuration to be tested will be made of 4 inch diameter tubes on 2-diameter center-to-center equilateral triangular spacing. This configuration was selected based on the results of testing alternative tube arrangements during the Two-Dimensional Flow Visualization Studies completed as an earlier task in this program.

The precombustor and FBC test chamber is being fabricated by Process Combustion Corp. in Pittsburgh, Penna. The unit is scheduled for delivery to the Florham Park site during July.

Details of the Flue Gas Cooler are shown in Figure 12. This piece of equipment required some special design consideration. The function of the cooler is to reduce the temperature of the flue gas coming off the primary cyclones from about 1600°F (FBC bed operating temperature) to a temperature range of 325°F to 425°F. This temperature range is rather critical since the gas must be maintained above its acid dew point to prevent serious acid corrosion and below the maximum working temperature for the Nomex bags in the downstream bag filter.

It was originally planned to control this temperature by using a direct water spray system. However, based primarily on the experience at the Exxon Miniplant FBC unit, it was concluded that the use of direct water sprays could cause the limestone particulate entrained in the flue gas to deposit out in a cement-like coating on the inside surfaces of the ducting downstream of the quench section or on the bag filter cold surfaces. Consequently, the decision was made to provide a closed circuit tubular cooling water system in this service.

In addition to the critical and relatively narrow permissible temperature range for the flue gas, the design requirements were further complicated by the wide range of bed temperatures (1100°F to 1600°F) and flue gas flow rates (1000-5000 SCFM) in the planned test matrix. It was obvious that all these requirements could not be met with a fixed surface water cooler.

The cooler is, therefore, designed with six cooler segments manifolded in series flow. Flue gas temperature will be maintained within the 325-425°F range by flowing water through more or fewer of the coil segments as the heat removal demands change.

A brief description, suppliers and anticipated delivery schedules for all major equipment components that will make up this Heat Flux Test Unit are shown on the attached Equipment List, Figure 13. Based on these deliveries a First Quarter 1979 construction completion for this unit is anticipated.

FORM ERDA 535
(2/1/77)

U. S. ENERGY RESEARCH AND DEVELOPMENT ADMINISTRATION

MILESTONE PLAN AND MANAGEMENT REPORT

1. Contract Identification: Industrial Application of Fluidized Bed Combustion - Category III Indirect Fired Heaters	2. Reporting Period through 6/30/78	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

9. Fiscal Years and Months: FY '78 | FY'79 | FY'80 | FY'81

CY'78 | CY'79 | CY'80 | CY'81

D J F M A M J J A S O N D J F M A M J J A S O N D J F M A M J J A S O N D J F M A M J J A S

7. Identification Number	8. Reporting Category (e.g., contract line item or work breakdown structure element)	9. Start	10. Percent Complete
1	Project Management	7/76	
2	2-Dimensional Flow	8/76	100
3	Process Stream Coking Studies	8/76	
4	Fluidized Bed Flux Studies	1/77	
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. Construction complete 2. Preliminary economics

12. Signature of Contractor's Project Manager and Date D. C. Cherrington 7/25/78	13. Signature of Government Technical Representative and Date

FIGURE 1

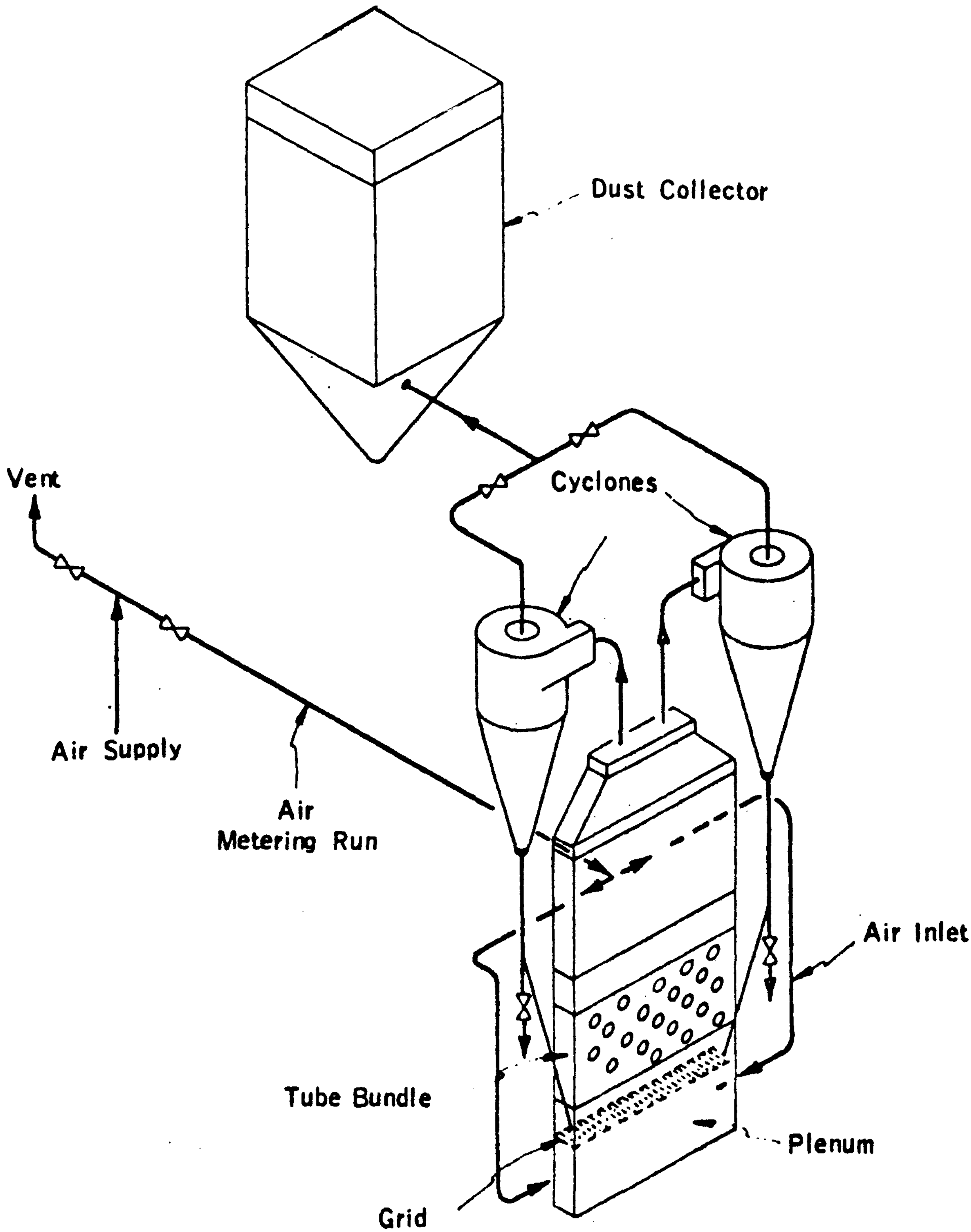
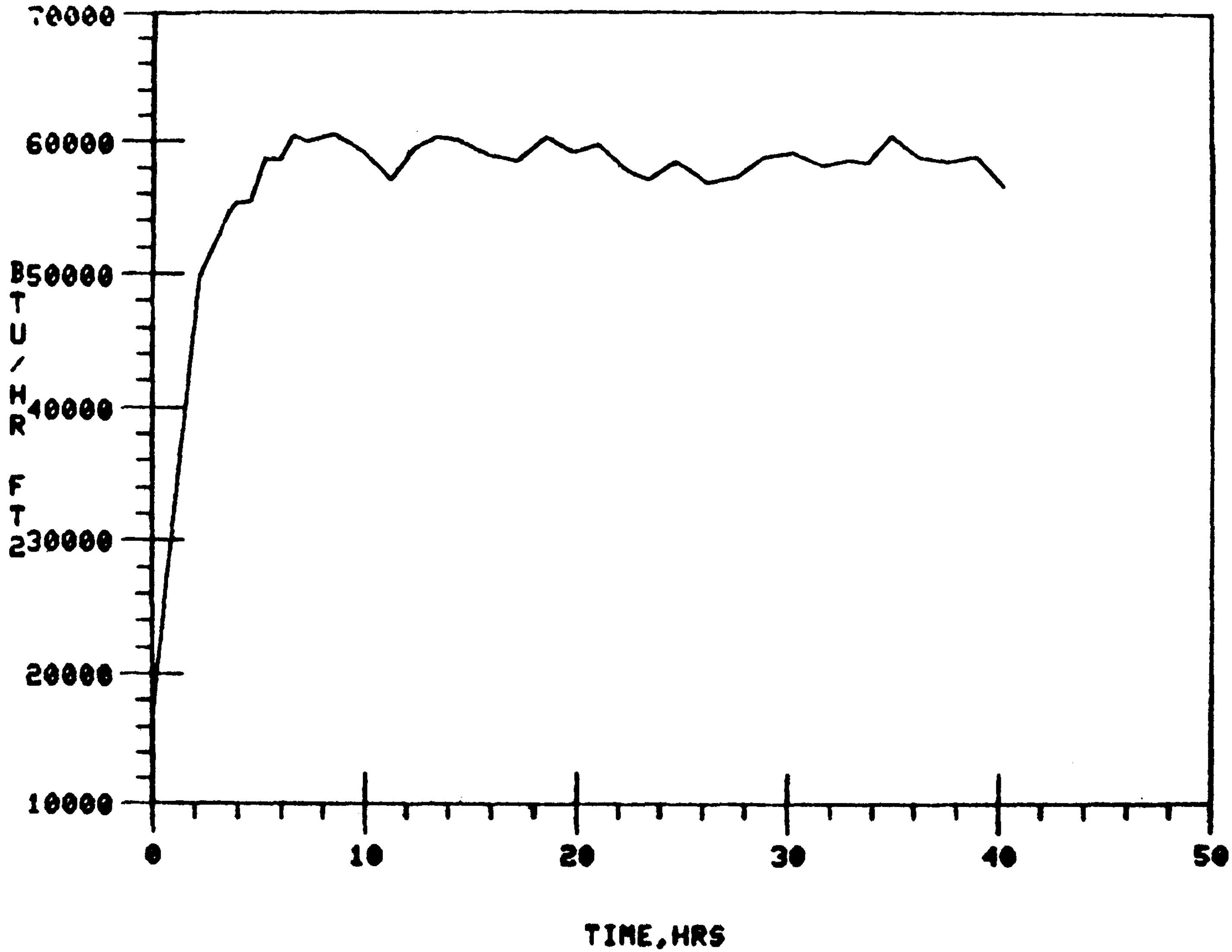


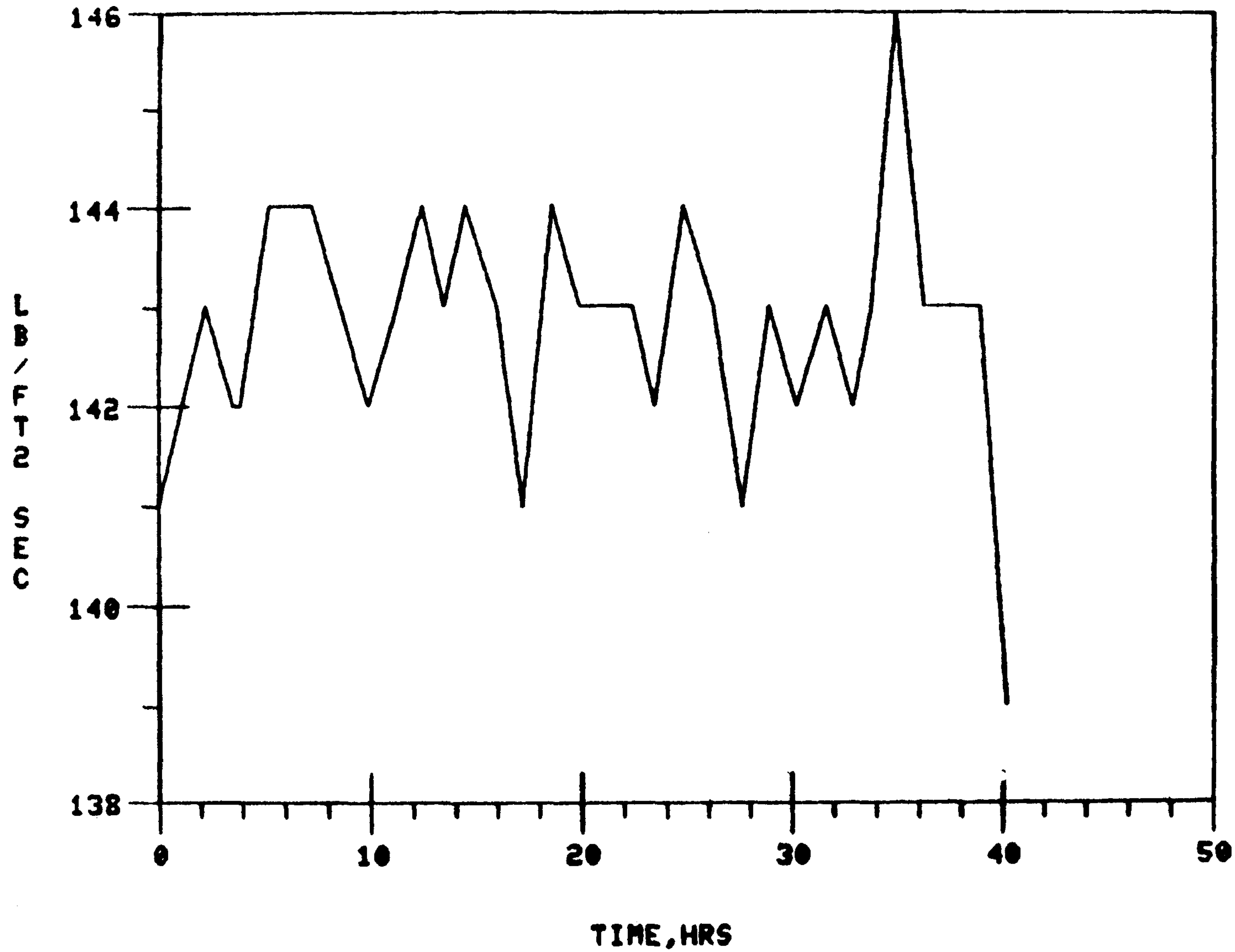
Figure 2

SCHEMATIC-TWO DIMENSIONAL FLOW VISUALIZATION UNIT



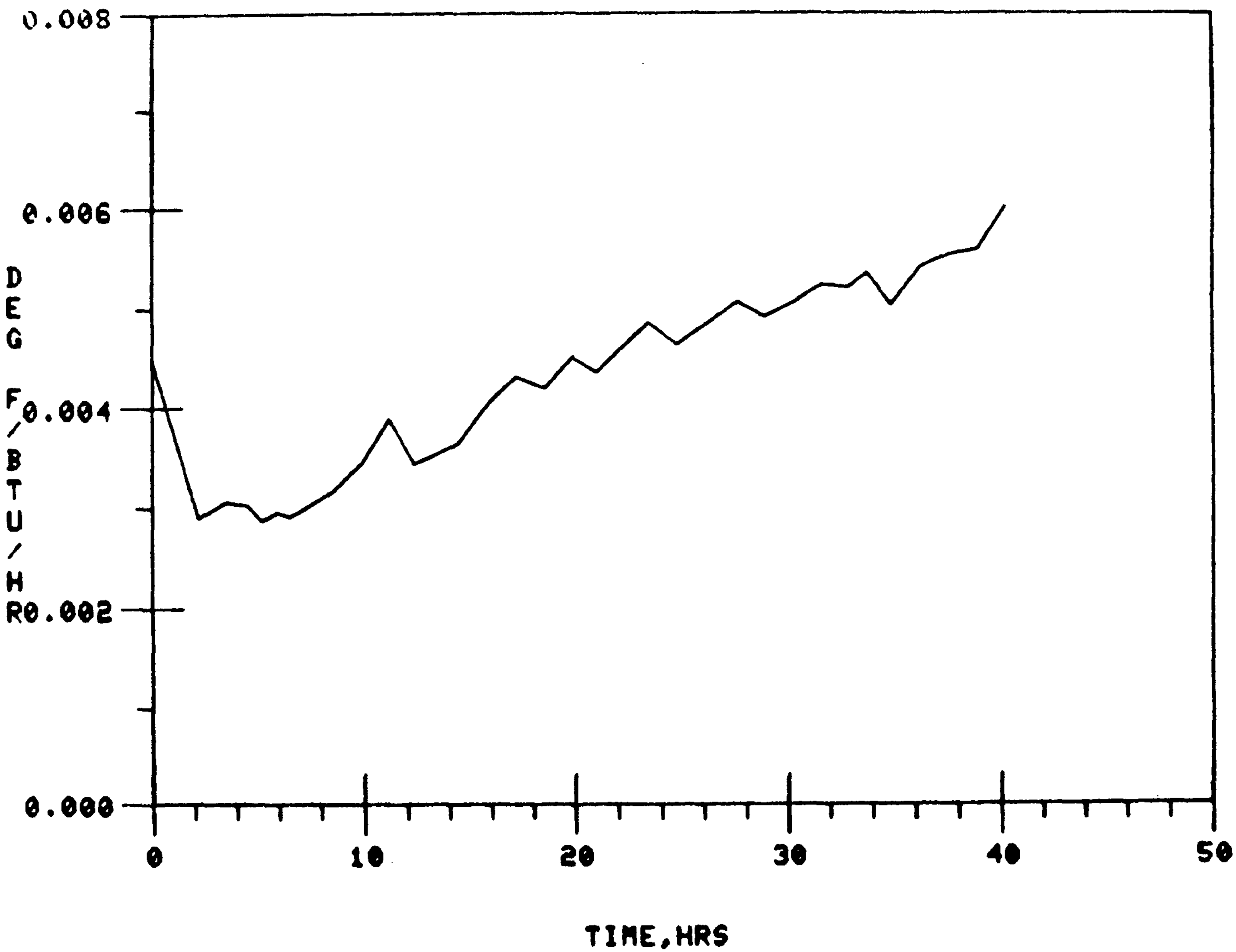
HEATER D HEAT FLUX PROFILE - TEST 2

FIGURE 3



HEATER D MASS VELOCITY PROFILE - TEST 2

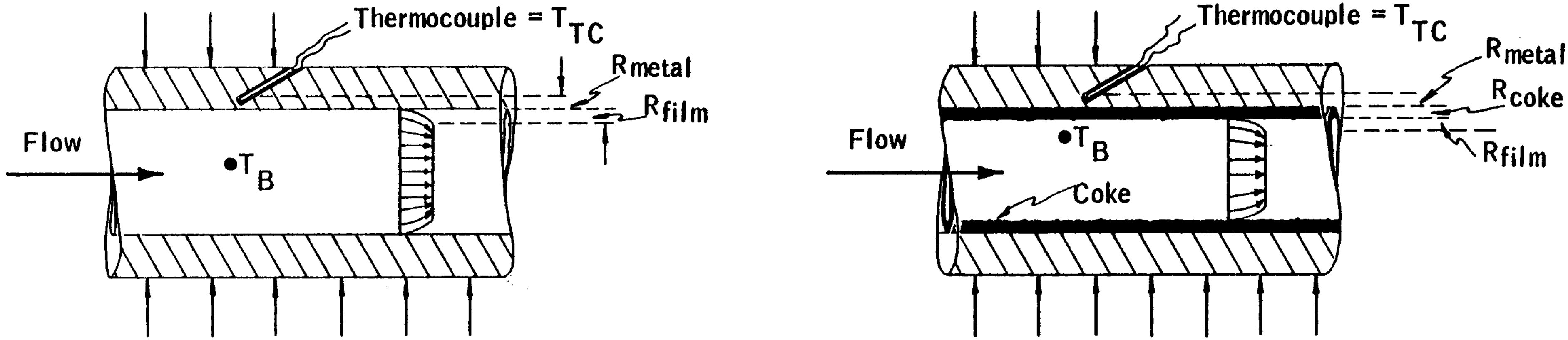
FIGURE 4



THERMAL RESISTANCE AT THERMOCOUPLE POSITION 7 HEATER D - TEST 2

FIGURE 5

INITIAL CONDITION — COKED CONDITION



Uniform Heat Flux — Uniform Heat Flux

$$\text{Thermal Resistance} = \frac{\Delta T}{\text{Heat Input}} = \frac{T_{TC} - T_B}{(\text{Heat Flux}) \times A}$$

- T_{TC} - Measured Wall Temperature
- T_B - Measured Fluid Bulk Temperature
- Heat Flux - Measured
- A - Inside Tube Surface Area

Initial Resistance, R_i

$$R_i = R_{metal} + R_{film}$$

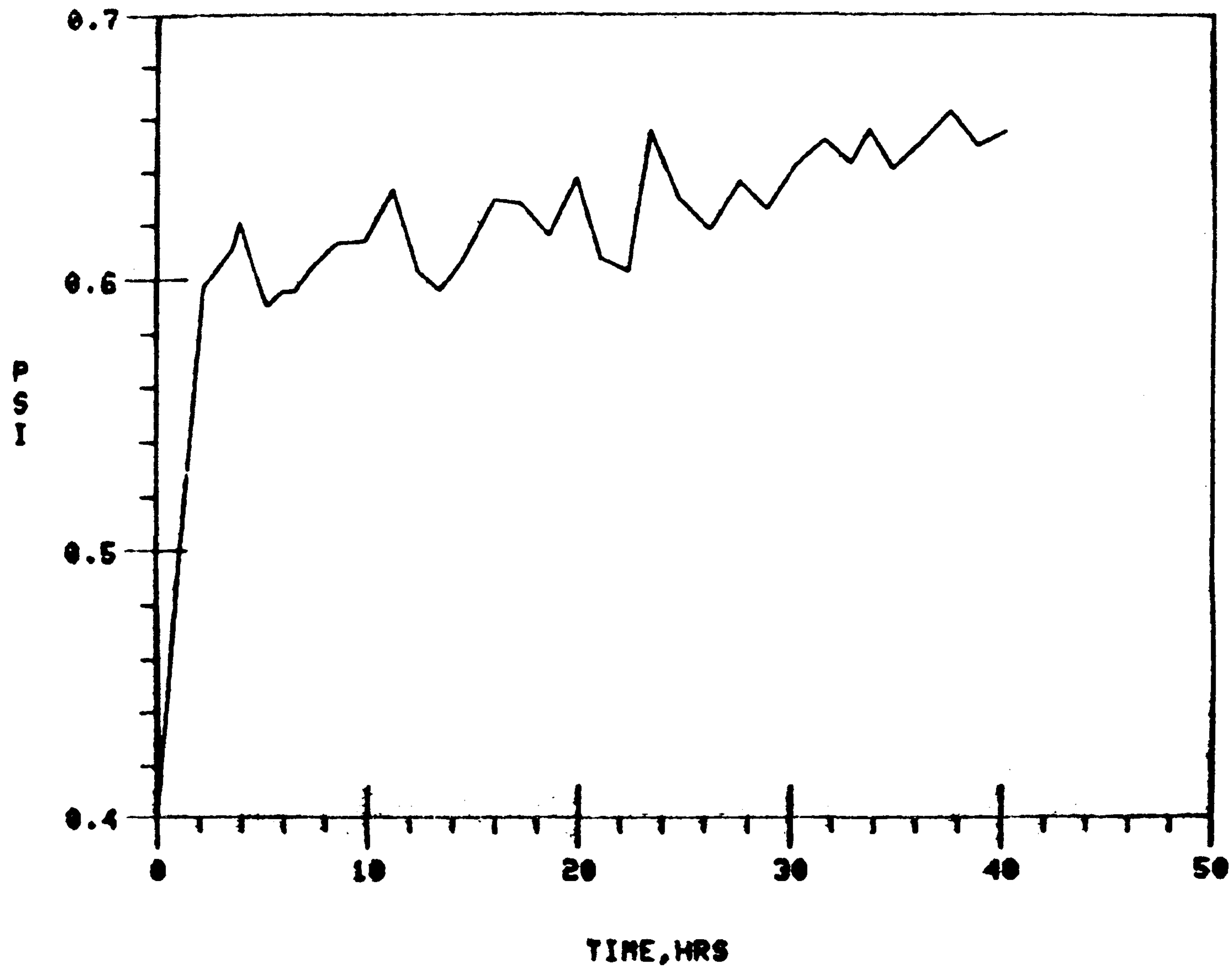
Coked Resistance, R

$$R = R_{metal} + R_{coke} + R_{film}$$

Effect of Coking: $R - R_i = R_{coke}$

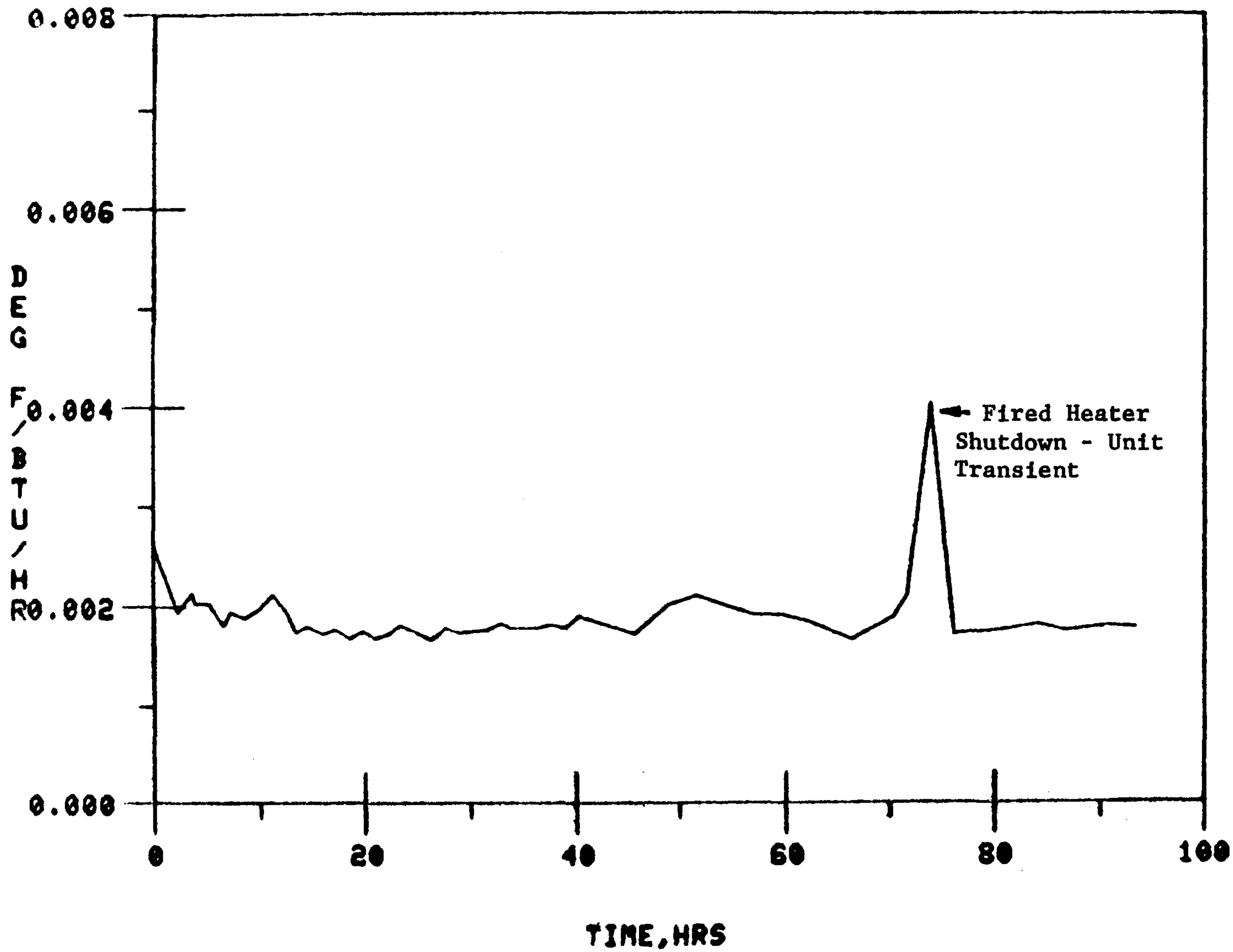
Figure 6

EFFECT OF COKE LAYER ON HEAT EXCHANGER TUBE THERMAL RESISTANCE



HEATER D PRESSURE DROP PROFILE - TEST 2

FIGURE 7



HEATER A THERMAL RESISTANCE AT THERMOCOUPLE POSITION 8 - TEST 2

FIGURE 8

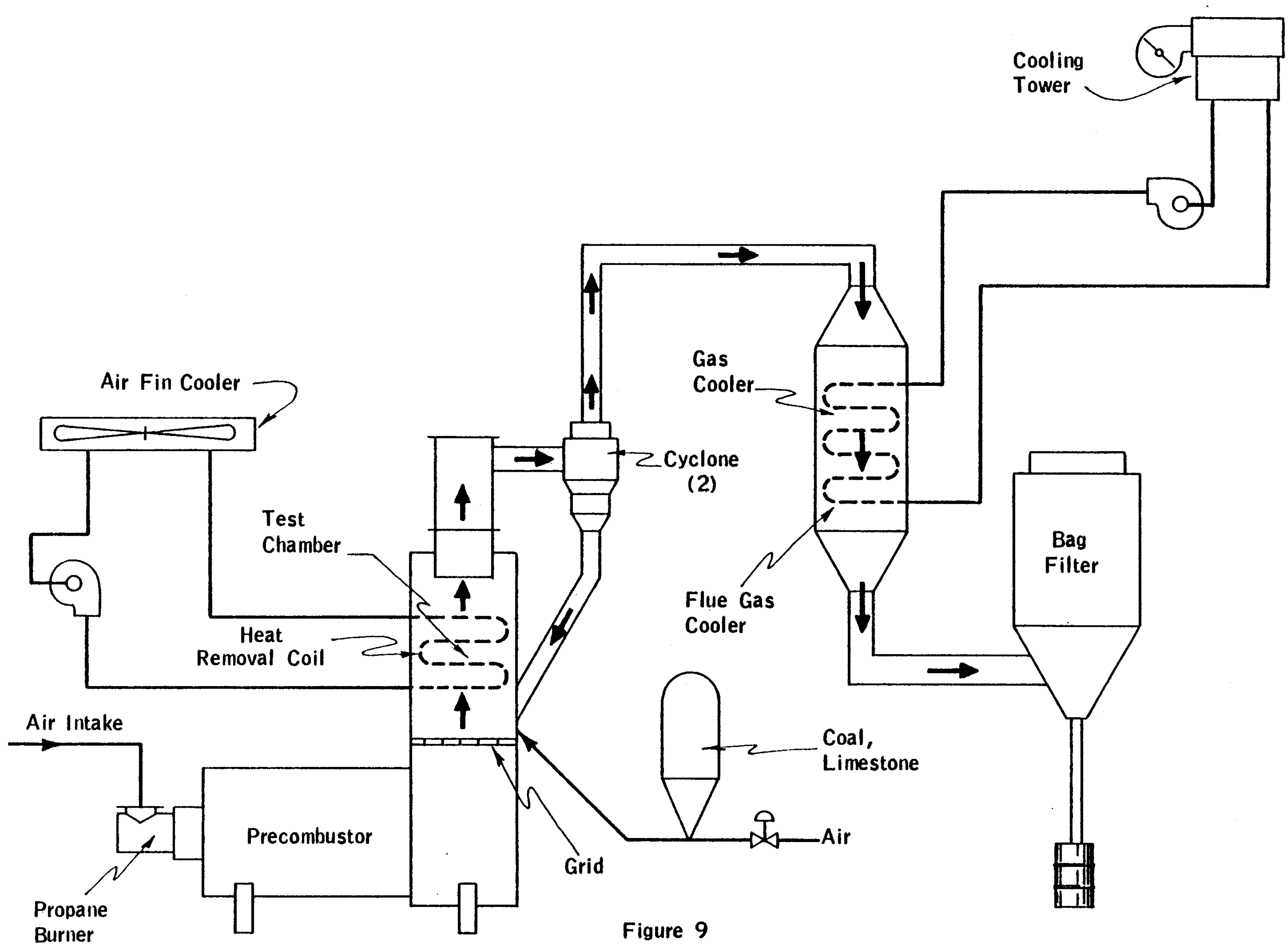
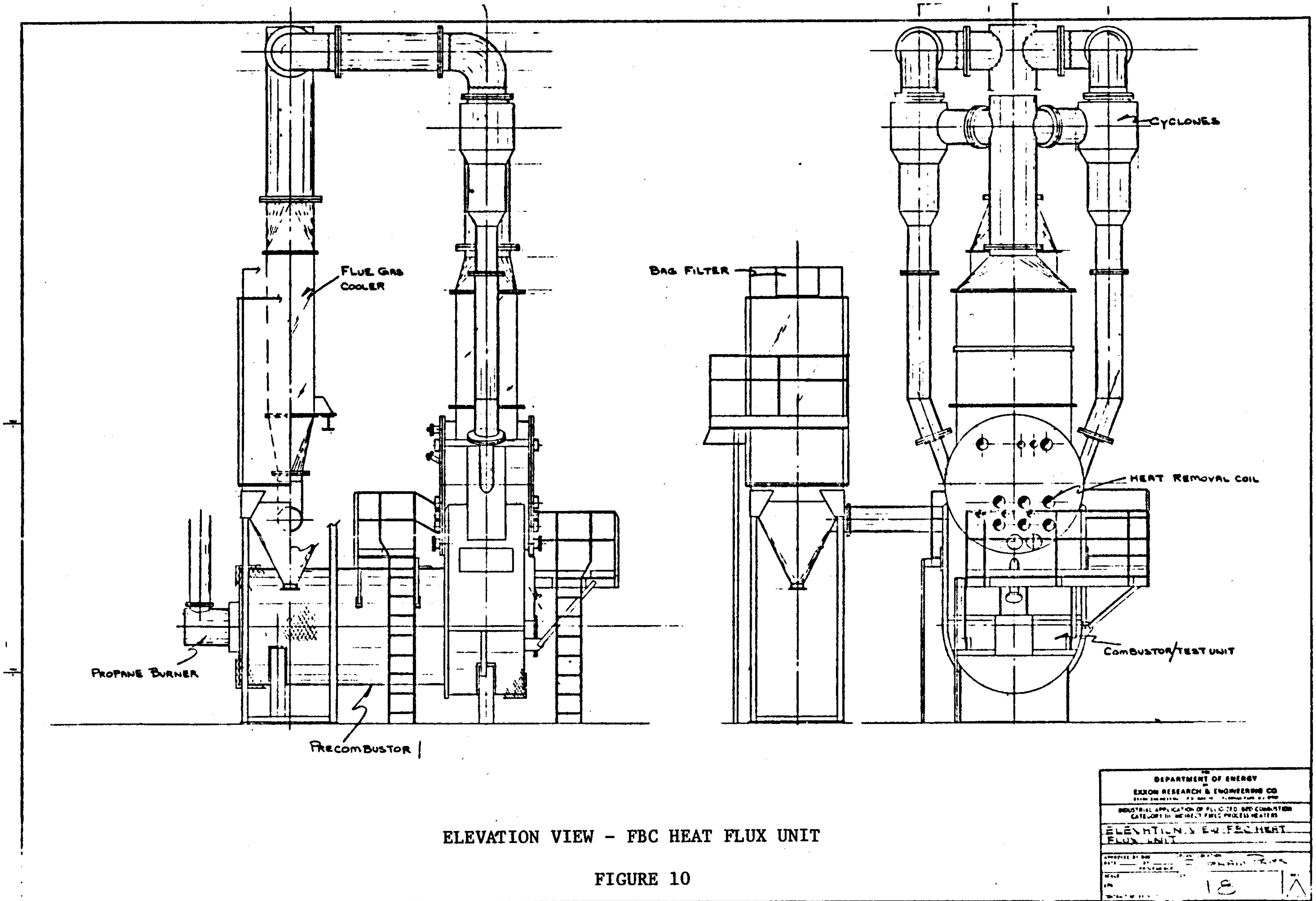


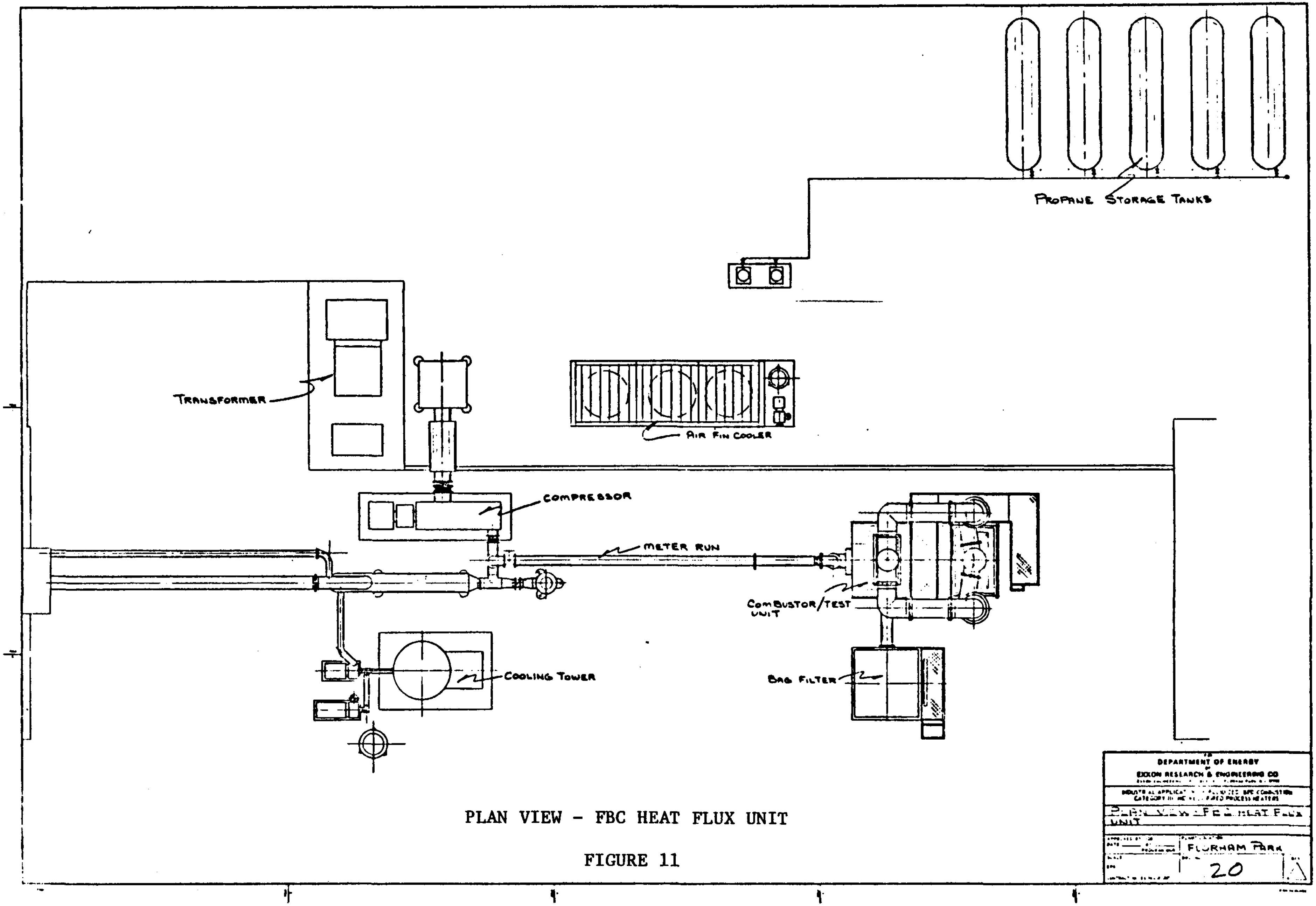
Figure 9

FLOW PLAN
FBC HEAT FLUX UNIT



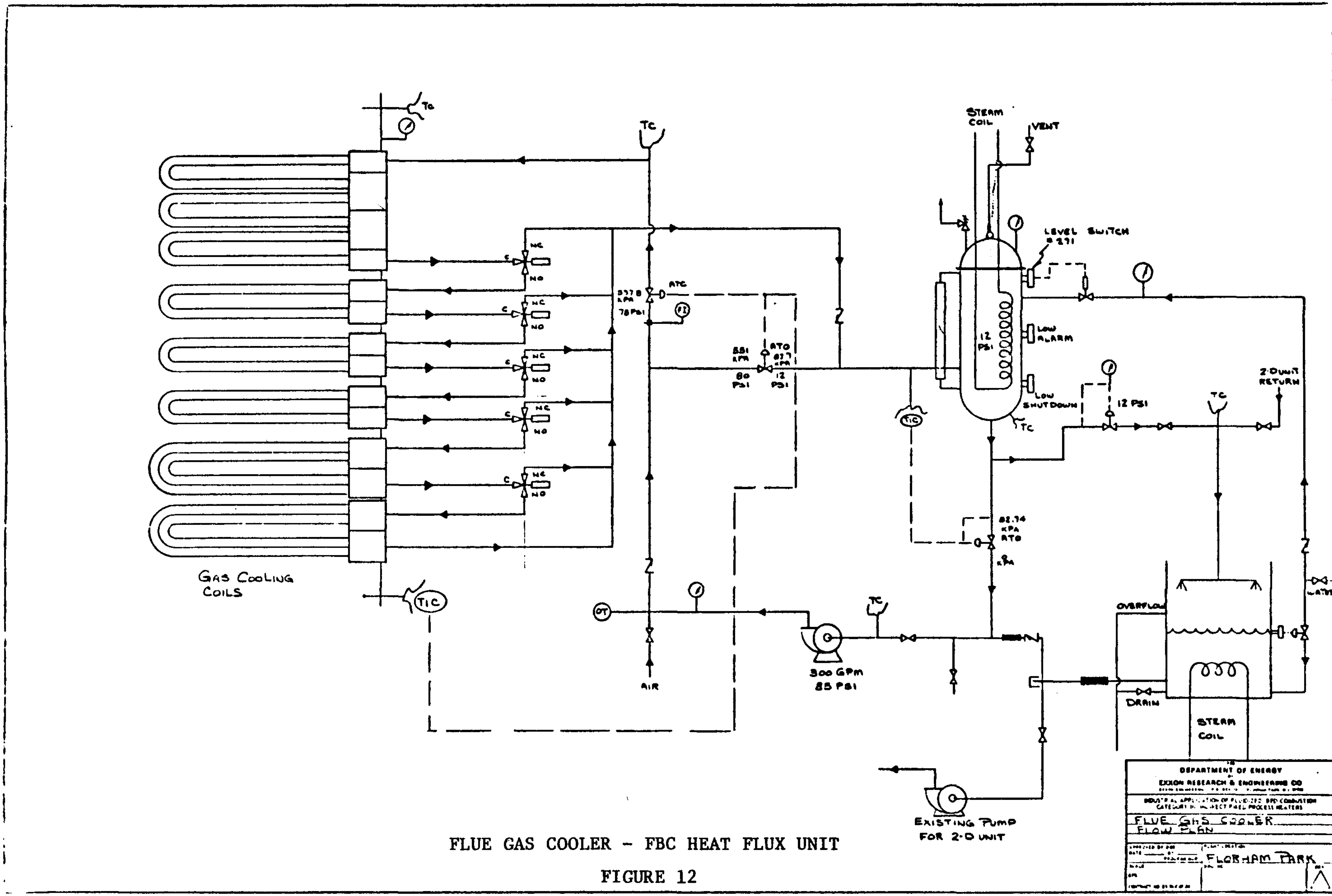
ELEVATION VIEW - FBC HEAT FLUX UNIT

FIGURE 10



PLAN VIEW - FBC HEAT FLUX UNIT

FIGURE 11



FLUE GAS COOLER - FBC HEAT FLUX UNIT

FIGURE 12

EQUIPMENT LIST
FBC HEAT FLUX TEST UNIT

Item	Description	Supplier	Anticipated Delivery
Air meter run	10" sch. 20 x 31' - 0' meter run for 3467 cfm at 14.7 psia, 60°F	Daniel	On-site
Air fin cooler	Capacity: 5 MBtu/hr, Size: 7 ft by 20 ft; To cool water passing through coils in test chamber	GEA	July 10
Precombustor/ FBC Test Chamber	Designed for operation up to 1600°F, Capacity: 12 MBtu/hr	Process Comb. Corp.	July 17
Cyclones (2)	Refractory-lined (1600°F max) in parallel; max gas rate, 80 acfs; max dust loading, 0.04 lb/acfs	Ducon	Sept. 19
Flue Gas Cooler	Capacity: 4.5 MBtu/hr; cooling flue gas from 1000-1600°F to 300-400°F	------ to be developed ------	
Bag Filter	5820 ACFM at 300-400°F; max uncaptured effluent less than 1 lb/hr solids	MikroPul	Oct. 17
Cooling Tower	For water used in flue gas cooler; cooling 150°F to 120°F	Delta	On-site(1)
Coal and Limestone Handling System	To be added at a later date.		

(1) Used during 2-D Flow Visualization Studies

FIGURE 13