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60 MWt METHANATION PLANT DESIGN FOR HTGR PROCESS HEAT

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July 1982

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Advanced Reactor Systems Department
Sunnyvale, California

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ADVANCED REACTOR SYSTEMS DEPARTMENT • GENERAL ELECTRIC COMPANY
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ABSTRACT

This report describes a 60 MW_t Methanation Plant for generating steam for industrial applications. The plant consists of four 15 MW_t methanation trains. Each train is connected to a pipeline and receives synthesis gas (syngas) from a High Temperature Gas-Cooled Reactor Reforming (HTGR-R) plant. Conversion of the syngas to methane and water releases exothermic heat which is used to generate steam.

Syngas is received at the Methanation Plant at a temperature of 80° F and 900 psia. One adiabatic catalytic reactor and one isothermal catalytic reactor, in each methanation train, converts the syngas to 92.2% (dry bases) methane. Methane and condensate are returned at temperatures of 100-125°F and at pressures of 860-870 psia to the HTGR-R plant for the reproduction of syngas.

Feedwater, mostly condensate returned from the industrial users, is received at a temperature of 100° F and a pressure of 50 psia. The water is pumped through a series of heat exchangers to the isothermal methanator which also functions as a steam generator. Saturated steam is superheated to obtain a final temperature of 730° F at a pressure of 900 psia. A steam user five miles from the Methanation Plant receives steam at about 530° F and 700 psia.

Each methanation train includes, in addition to the methanators, desulfurizers, gas-to-gas heat exchangers, gas-to-water heat exchangers, steam superheater, steam drum, and deaerator as well as appropriate pumps and compressors.

The methanation train size of 15 MW_t and the use of an advanced monolithic (honeycomb) catalyst reduces the size of the methanators and heat exchangers so that all components can be factory assembled into modules for railroad transport. A capital cost savings of 31%, over a single 40 MW_t methanation train, is estimated.

ACKNOWLEDGEMENT

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1. INTRODUCTION

This final report describes a 60 MWt Methanation Plant used to supply steam to several, diverse industrial users. The plant consists of four 15 MWt Methanation Trains, a single Methanation Process Support System, a control room and a maintenance building. The methanation plant receives syngas from a High Temperature Gas-Cooled Reactor Reforming (HTGR-R) plant via a pipeline at 80°F and 900 psia. Steam leaves the methanation plant via distribution pipes at 730°F and 900 psia. Feedwater is about 90% condensate returned from the steam users and 10% make-up with temperature of 100°F at 50 psia.

The 60 MWt Methanation Plant is a "stand-alone" facility located as far as 100 miles or more away from the HTGR-R plant. All the steam users are assumed to be within a 5 mile radius of the methanation plant. Distribution piping connects the methanation plant with the users and parallel piping is provided for return of condensate to the methanation plant.

1.1 Background

The 60 MWt Methanation Plant is an effort to improve the efficiency, reliability and availability and to decrease the capital costs estimated for the 40 MWt Methanation plant designed last year and reported in Reference 1. Unless specifically addressed in this report the information presented in Reference 1 is used in the 60 MWt Methanation Plant design. This includes such specifics as material of construction, appropriate codes and standards, safety requirements, and operating procedures.

1.2 Objectives

The objectives of the 60 MWt Methanation Plant were addressed as follows:

1.2.1 Improved Efficiency

Too much heat from the methane gas and condensate streams of the 40 MWt methanation plant was given up to atmosphere. The resultant efficiency for that plant was 89%. Two heat exchangers have been added to the 15 MWt methanation trains; one transfers heat from the methane gas stream to incoming feedwater and the other transfers heat from the condensate to the incoming feedwater. The temperature of the feedwater is raised from 100°F to 220°F and the resulting plant efficiency is raised to 98.2%

1.2.2 Improved Reliability

Reliability is dependent upon syngas supply, outages for maintenance and outages for catalyst replacement. By using four-15 MWt methanation trains, one train is out of operation 16 hours every day. Anticipated steam demand is 10-15 MWt for 8 hours (1 train), 40-45 MWt for 8 hours (3 trains) and 50-60 MWt for 8 hours (4 trains). This virtually eliminates outages for maintenance and the overall reliability is improved by 3-4% over the 40 MW_t Methanation Plant design.

1.2.3 Improved Availability

The use of four-15 MWt methanation trains improves the availability of the plant by 9-10% over the single-train 40 MW_t Methanation Plant design.

1.2.4 Decreased Capital Cost

Many factors contribute to reducing capital cost in spite of the plant size being increased from 40 MWt to 60 MWt and the number of methanation trains being increased from one to four. The significant factors are:

a. The use of monolithic catalyst, decreases the total volume of catalyst required from 1,530 ft^3 to 256 ft^3. This decreases the methanator sizes by a factor of 6.

b. Decrease in output steam temperature from 900°F to 730°F eliminates one adiabatic methanator in the 15 MWt methanation train. The decrease in steam temperature also decreases the thickness of tube sheets and vessel walls and flanges.

c. Increase in the L/D ratio of the desulfurizers, methanator vessels and the steam drum results in decreased wall thickness and the use of seamless or centrifugal cast tubes which decreases fabrication costs.

d. Modular Construction - Each of the four 15 MWt methanation trains are to be assembled into 5 modules per train at the fabricators. This should reduce the total plant cost by about 15% (Reference 2) since field labor for component installation, electrical and instrumentation wiring are greatly reduced and field piping is almost eliminated.

Table 1-1 presents a comparison of all the key features of the 40 MWt single train and the 60 MWt four train methanation plants.

1.3 Marketing Requirements

Marketing research, completed late in FY 1981, indicates that a methanation plant supplying steam to many diverse industrial users would need to have variable output to meet the demands. The number of users operating on one shift, two shifts and three shifts per day was approximately equal and the suggested demand to be satisfied was about 10 MW_t, 30 MW_t and 45 MW_t, respectively. (See Appendix A.)

In order to satisfy the market demand, the decision was made to use four 15 MW_t methanation trains which would also provide additional margin in terms of reliability and availability. The use of four 10 MW_t trains would not satisfy the demand and the use of three 20 MW_t trains introduced problems of operating flexibility and when one train was down, the peak-load demand could not be satisfied. The design characteristics of the Advanced Methanation System were decided in the first quarter of FY 1982, as reported in Appendix B. Later review by Marketing indicated that the choice of four 15 MW_t methanation trains was appropriate (see Appendix C).

TABLE 1-1

COMPARISON
40 MWt (1981) vs. 60 MWt (1982)
METHANATION PLANTS

Item	1981	1982
Capacity (MWt)		
Peak Power	40.0	60.0
Average Power	26.7	40.0
Minimum Power	20.0	10.0
Methanator Trains (No. x MWt)	1x40	4x15
Catalyst Type	Tablet	Monolithic
Catalyst Volume (Ft^3/Train)	1530	64
Syngas In (lbm-hr/°F/psia)	72,144/100/900	119,246/80/900
CH_4 Out (lbm-hr/°F/psia)	37,448/100/870	61,576/100/860
Condensate Out (lbm-hr/°F/psia)	34,696/280/870	57,670/125/865
Feedwater In (lbm-hr/°F/psia)	100,000/100/40	159,600/100/50
Steam Out (lbm-hr/°F/psia)	100,000/900/900	159,600/730/900
CH_4 Conversion (%-dry)	94.1	92.2
Efficiency (%)	89.0	98.2
Reliablity (%)	$>$ 95	$>$ 98
Availability (%)		
Peak Power	$>$ 90	$>$ 98
Average Power	$>$ 90	$>$ 99
Minimum Power	$>$ 90	$>$ 99

1.4 References

1. Davis, C. R., et al., GEFR-00568, "Gas-Cooled Reactor Program Application, Methanation Plant Design for HTGR Process Heat," September 1981.

2. Glaser, L. B., Kramer, J., and Siegelman, G. A., "Modular and Barge - Mounted Plants - An Update," Chemical Engineering Progress, September 1981.

2. SUMMARY

2.1 Significant Accomplishments

The design of a 60 MWt Methanation Plant using four-15 MWt Methanation Trains shows significant improvements over the 40 MWt, single train, methanation plant designed in 1981. The improvements are:

a. Efficiency. Efficiency of the plant is increased from 89% to 98.2% by use of two additional feedwater heaters.

b. Reliability. Reliability is improved by 3-4% since 16 hours/day is available for maintenance. Only planned maintenance outages required are one week every 2.5 years to replace methanator catalyst.

c. Availability. Availability is increased to over 99% for the average demand (40 MWt) and to at least 98% for peak demand (60 MWt).

d. Capital Costs. Capital cost reduction contributions include using monolithic catalyst, smaller vessels and heat exchangers, reducing the steam temperature from 900°F to 730°F and by using modular construction which minimizes the field labor costs. The final capital cost reduction is estimated at 31%; from $\$16 \times 10^6$ to $\$11 \times 10^6$ per plant as presented in Appendix G.

Plant sizes can be any capacity between 15 MWt to 90 MWt in increments of 15 MWt. For plants of 100 MWt and above the Methanation Train size can be increased to 25 MWt and the improvements noted above would still apply.

2.2 Recommendations

The paramount need of the Methanator Design and Development Task is the development of the monolithic (honeycomb) nickel impregnated catalyst used in the current Advanced Methanation System Design. The catalyst used in the current design has received limited laboratory testing and examination on only a very small scale.

It is recommended that the monolithic catalyst used in the current design (size and shape) be "bench-tested" in the coming fiscal year. The adequacy of pure nickel must be proven or the need for a nickel alloy determined. An economical manufacturing process of the catalyst in production quantities must be developed. Performance of the catalyst at intended operating conditions must be measured. The METH computer code must be validated by actual test data. This portion of the catalyst development is applicable to both the adiabatic methanator and the reformer in the HTGR-R plant.

The appropriateness of using the monolithic catalyst in the isothermal methanator will require additional bench-tests, followed by testing in tubes, submerged in water at the intended operating conditions.

It is recommended that a working arrangement be established with a catalyst manufacturer(s) for procurement of sample catalyst. A contract should also be negotiated for the performance of the catalyst tests by a knowledgeable expert with adequate testing facilities. It is recommended that "proof-of-performance" testing of the catalyst for the isothermal methanator application be accomplished by testing a "few tubes" with monolithic catalyst in the FRG's Adam-Eva facility. Initial discussions with the operators of Adam-Eva indicate a strong desire on their part to participate in a joint effort of catalyst testing and evaluation.

3. 60 MWt METHANATION PLANT DESCRIPTION

Syngas generated by a HTGR-R plant is shipped to the methanation plant by a pipeline. Each of the four - 15 MWt methanation trains are connected to the pipeline. The connecting pipe includes pressure control, flow control and emergency shutoff valves. At maximum capacity of 15 MWt, each train receives syngas at the rate of 497 lbm/min., at a temperature of 80°F and a pressure of 900 psia.

Feedwater is mostly condensate received from local steam users, mixed with makeup water. At full capacity operations each methanation train receives 665 lbm/min. at a temperature of 100°F and a pressure of 50 psia.

Each methanation train produces 665 lbm/min. of steam at a temperature of 730°F and a pressure of 900 psia for distribution to the steam users. Calculations indicate that a user at the end of a 5-mile distribution pipe will receive steam at about 530°F and at about 700 psia.

During the methanation process the syngas is converted to 92.2% (dry bases) methane. The methane, 256 lbm/min. per train, at a temperature of 100°F and a pressure of 865 psia is piped backed to the HTGR-R plant. Condensate is also returned to the HTGR-R plant in a parallel pipe at the rate of 244 lbm/min. per train. The returned condensate may not require clean-up or clean-up will be accomplished at the HTGR-R plant.

3.1 Methanation Process System

The methanation process system is identical for each of the four trains. The system converts syngas (H_2 + CO_2 + CO) to methane and water (CH_4 + H_2O) by passing the gas through a monolithic nickel catalyst which produces exothermic heat. Part of the exothermic heat is used to preheat the syngas for desulfurization and to sustain the catalytic reaction. The remaining heat is used to convert 100°F water to 730°F superheated steam.

3.1.1 Normal Operation

During normal operation the methanation systems (4 trains) are self-sustaining except for electrical power used to drive the feedwater pumps and two recycle-gas compressors. (Even these components could be driven by steam turbines fed by steam generated by the methanation process but electrical motor drives would still be needed for start-up operations.)

3.1.1.1 Syngas to Methane Conversion

(NOTE: The reader will be able to follow the flow of syngas/methane and condensate if he uses Figure 3-1, "Flow Diagram" as he reads the following description)

Syngas at 80°F and 900 psia flows from the pipeline to the No. 1 Syngas Heater where its temperature is raised to 250°F, then through No. 2 Syngas Heater where the temperature is raised to 425°F and then through No. 3 Syngas Heater where the temperature is increased to 651°F. From No. 3 Syngas Heater it flows through the desulfurizers where the sulfur content is reduced to less than 0.05 ppmv. From the desulfurizers the syngas flows to a separation point where part of the flow is routed to the No. 1 Adiabatic Methanator and the remainder is routed to the No. 2 Isothermal Methanator/-Evaporator. At the entrance of No. 1 Methanator the fresh syngas is mixed with recycle gas taken from the exit of the No. 1 Methanator. At the exit of No. 1 Methanator the gas stream is split; a portion is routed through the Steam Superheater and No. 3 Syngas Heater and then returned to the entrance of No. 1 Methanator, the remainder of the gas flows to the entrance of No. 2 Methanator/Evaporator. At the entrance of No. 2 Methanator/Evaporator there is a mixing of fresh syngas, gas from No. 1 Methanator and recycle-gas from the exit of No. 2 Methanator. Gas not recycled from the exit of No. 2 Methanator flows sequentially to the No. 3 Feedwater Heater, No. 2 Syngas Heater, No. 2 Feedwater Heater, No. 1 Syngas Heater and No. 1G Feedwater Heater. The temperature of the gas leaving the No. 2 Methanator is 734°F which is lowered by steps as it flows through the 5 heat exchangers until it becomes 130°F upon leaving the No. 1G Feedwater Heater. From No. 1G Feedwater Heater the methane flows through an Air Cooler, K. O. Drum and Dryer after which it is piped back to the HTGR-R plant at a temperature of 100°F and a pressure of 865 psia.

As the temperature of the methane gas and steam mixture decreases the steam condenses. The condensate formed is drained from the heat exchangers and routed through the No. 1C Feedwater Heater where heat of the condensate is transferred to a portion of the incoming feedwater. The condensate is then collected in a drum, after which it is piped back to the HTGR-R plant at a temperature of 125°F and a pressure of 870 psia.

3.1.1.2 Water to Steam Conversion

Feedwater for the methanation plant is primarily condensate, returned from the local steam users. The condensate is "cleaned" as required and mixed with purified makeup water. It is assumed that feedwater flows into No. 1G and No. 1C Feedwater Heaters at a temperature of 100°F and a pressure of 50 psia. From these two heat exchangers the water flows into the Deaerator where it is mixed with a small quantity of saturated steam taken from the Steam Drum. The resulting mixture in the deaerator is 240°F and 30 psia and all entrained gases are vented to atmosphere. The flow of feedwater is 665 lbm/min. and the flow of steam is 14 lbm/min. The feedwater pump pumps the water through the No. 2 and No. 3 Feedwater Heaters at temperatures of 240°F increasing to 492°F and pressures of 915 psia decreasing to 909 psia. From No. 3 Feedwater Heater the water flows to the Steam Drum where it is mixed with saturated water and steam to obtain a stable temperature of 532°F at 905 psia. The Steam Drum is located above the No. 2 Methanator/Evaporator and the two are connected by four-4 inch downcomers and four 6-inch raisers. Water flows, by natural circulation down through the methanator/evaporator where boiling occurs and saturated water and steam raises back to the steam drum. In the drum the water and steam are separated and the steam flows to the Steam Superheater. Upon leaving the Steam Superheater the steam temperature is 730°F at a pressure of 900 psia which is delivered to the distribution pipelines at the rate of 665 lbm/min.

3.1.2 Partial Capacity Operation

The 60 MWt Methanation Plant can be operated at almost any desired capacity; from a low of 67% of one 15 MWt Train (10 MWt) up to 110% of all four - 15 MWt Trains (66 MWt) of its nominal rating. For operations between 15 MWt and 60 MWt it is best to operate as many trains as possible at their nominal rating with the other trains on hot standby.

The 110% operation requires a 10% increase in syngas feed flow and 12% increase in feedwater flow. This requires an increase in pressure of about 6 psia for both. The end results would be no change in the temperatures of the syngas to methane stream, but the superheated steam would be delivered to the distribution pipeline at about 720°F and 906 psia.

For decreased capacity operation the syngas feed and feedwater flow would be reduced proportionally, with the pressure held at the nominal 900 psia. For 67% rated capacity operation the ratio of recycle gas to feed gas for the adiabatic methanator must remain the same as for rated capacity. The ratio of recycle gas to feed gas for the Isothermal Methanator/Evaporator must be increased to maintain the same outlet temperatures (gas and steam) as for rated capacity operation. At reduced capacity operation, all heat exchangers will be oversized, so to maintain desired mass and energy transfer in the heat exchangers, they all have "by-pass" water lines with flow control valves. This arrangement permits maintaining all temperatures and pressures, of both syngas to methane and water to steam streams, the same as for rated capacity operation and only the quantities of methane and steam produced per unit of time will be reduced.

3.2 Methanator Process Support System

The purpose of the Support System is to supply heat to start-up the methanation systems from ambient temperature, to reduce new catalyst and to supply supplementary heat for shutdown and hot standby operations. One Dryer, Condensate Drum and Odorizer are also included in the support system. These components are designed to handle the methane and condensate from one to four 15 MWt Methanation Trains. This is shown in Figure 3-3. The other components comprising the support system are:

a. Gas-fired nitrogen heater for desulfurizers.

b. Gas-fired nitrogen and hydrogen heater for the two methanators.

c. Gas-fired feedwater heater.

d. Liquified nitrogen supply tank.

e. Hydrogen supply (used only for catalyst reduction).

In addition to the above equipment, the feedwater pump and the two recycle-gas compressors of each train are used for startup, shutdown and hot standby operation.

Gas to fuel the three heaters is taken from either the syngas supply or methane return pipeline. If these pipelines are uncharged, locally available natural gas or propane will be used.

3.2.1 Startup Procedure - See Note

3.2.2 Normal Shutdown - See Note

3.2.3 Hot Standby Operation - See Note

3.2.4 Emergency Shutdown - See Note

NOTE: Procedures and operations for 3.2.1 through 3.2.4 are the same for one 15 MWt Methanator Train as for the 40 MWt Methanator Train described in Sections 6.2.1 through 6.2.4 of Reference 1.1. These Sections are included as Appendix H to this report.

3.3 Methanation System Components

The 60 MWt Methanation Plant includes four-15 MWt Methanation Trains. Each train includes one adiabatic methanator, one isothermal methanator/evaporator, two desulfurizers, three syngas heaters, four feedwater heaters, steam drum, steam superheater, one feedwater pump, two recycle-gas compressors, one air cooler and one K. O. Drum with Flare Stack. A complete list of equipment, quantities and figure numbers are given in Table 3-1. Components not having Figure numbers are described by their capacity or functional requirements. These are commercial components which were selected by Bechtel.

In addition to the component drawings the following Tables and Figures are included:

Table 3-2: "Mass & Volumes, Syngas & Methane"
(Includes pipe and nozzle sizes)
Table 3-3: "Mass & Volumes, Water and Steam"
(Includes pipe and nozzle sizes)
Table 3-4: "Mass & Volume, Condensate"
(Includes pipe and nozzle sizes)
Figure 3-1: "Flow Diagram, 15 MWt Train"
Figure 3-2: "Piping Diagram, 15 MWt Train"
Figure 3-3: "Support System, 60 MWt Plant"

3.3.1 Additional Component Information

All heat exchangers, except the combination isothermal methanator/evaporator, are to be single-pass, tube and shell, of conventional TEMA-R design. The isothermal methanator/evaporator is two-pass tube side, single-pass shell side with removable tube bundle of the Patterson-Kellery Company design. All vessels, heat exchangers and steam generating equipment are designed to the ASME Code, Section VIII, Division 1. All piping is to be

TABLE 3-1

15 MWt Methanation Train
Equipment List

ITEM	QTY	DESCRIPTION
1. Desulfurizer	2	Fig. 3-4
2. Methanator #1	1	Fig. 3-5
3. Methanator #2/Evaporator	1	Fig. 3-6 & 3-7
4. Syngas Heater #1	1	Fig. 3-8
5. Syngas Heater #2	1	Fig. 3-9
6. Syngas Heater #3	1	Fig. 3-10
7. Steam Drum	1	Fig. 3-11 & 3-6
8. Steam Superheater	1	Fig. 3-12
9. Feedwater Heater #1C	1	Fig. 3-13
10. Feedwater Heater #1G	1	Fig. 3-14
11. Feedwater Heater #2	1	Fig. 3-15
12. Feedwater Heater #3	1	Fig. 3-16
13. Deaerator	1	
Capacity/Pressure		1,200 gallons/30 psia
Dimensions		3'-9" Dia x 13'-3" V. H. with 1/4" wall thickness
14. Feedwater Pump	3	2-50% capacity + 1 spare
Capacity/HP		40 gpm/30 HP (each)
Pressure In/Out		30/915 psia
Dimensions (ea unit)		18" x 58"

Table 3-1 (Con't)

15.	Recycle Gas Compressor for Methanator #1	3	2-50% Capacity + 1 spare
	Capacity		904 lbm (802 ft^3)/min.
	Temp & Press. In		651°F/886 psia
	Temp. & Press. Out		651°F/892 psia
	No. Stages		1
	Power		15 HP
	Dimensions		36" W x 48" L x 60" H
16.	Recycle Gas Compressor for Meth. #2/Evaporator	3	2-50% Capacity + 1 Spare
	Capacity		390.5 lbm (367 ft^3)/min
	Temp. & Press. In		734°F/875 psia
	Temp. & Press. Out		734°F/890 psia
	No. Stages		1
	Power		20 HP
	Dimensions		36" W x 48" L x 60" H
17.	Pressure Relief Valve	1	14 lbm (6.84 ft^3)/
	Capacity H_2O/Steam		221.9 ft^3/min
	Temp. & Press. In		532°F/905 psia
	Temp. & Press. Out		240°F/30 psia
18.	Air cooler	2	1-100% + 1 Spare
	Heat Removed		2.3 x 10^5 Btu/hr-max.
	Dimensions		120" W x 120" L x 144" H
19.	K. O. Drum with Flare Stack	1	
	Type		Package unit self-supporting with 2 pilot ignitors, pilot gas - +62 scfn.
	Capacity		135 cfm
	Temp/Press.		750°F/1000 psia
	Dimensions		48" Dia. x 108" H (Drum)

TABLE 3-2

MASS & VOLUMES
SYNGAS & METHANE

Line No.	Total Flow (lbm/Min)	Lbm/Mole	Pressure (psia)	Temperature (°F)	Flow (Ft³/Min-STP)	Flow (Ft³/Min-P-°F)	Pipe Size (Dia-Sch.)	Velocity (FPM)
1	496.9	10.3874	900	90	17,175	328	2.5-10	8,667
2	496.9	10.3874	898	250	17,175	434	2.5-10	11,470
3	496.9	10.3874	896	425	17,175	542	3.0-40	10,295
4	496.9	10.3874	894	651	17,175	682	3.5-40	9,931
5	496.9	10.3874	892	651	17,175	684	3.5-40	9,952
6	155.8	10.3874	892	651	5,385	215	2.0-40	9,211
7	1167.3	15.0934	892	651	27,764	1105	4.0-40	12,504
8	1167.3	16.2252	890	941	25,827	1300	5.0-80	10,293
9	1011.4	16.2252	890	941	22,369	1126	5.0-80	8,914
10	1011.4	16.2252	888	778	22,369	998	4.0-80	12,496
11	1011.4	16.2252	886	651	23,369	897	3.5-40	13,060
12	1011.4	16.2252	892	651	23,369	891	3.5-40	12,973
13	341.1	10.3874	892	651	11,789	470	3.0-40	9,158
14	156.7	16.2252	890	941	3,468	740	3.5-80	11,986
15	939.2	16.2252	890	734	20,215	882	4.0-80	11,044
16	944.8	16.6795	875	734	20,336	887	4.0-80	11,107
17	442.0	16.6795	875	734	9,513	415	2.5-40	12,476
18	442.0	16.6795	890	734	9,513	407	2.5-40	12,236
19	502.9	16.6795	875	734	10,825	472	3.0-40	9,198
20	502.9	16.6795	873	476	10,825	371	2.5-40	11,154
21	442.4	16.5116	871	426	9,619	312	2.0-40	13,388
22	319.2	15.9945	869	354	7,164	215	2.0-40	9,226
23	274.7	15.7034	867	265	6,264	168	1.5-40	11,883
24	260.5	15.5996	865	130	5,994	130	1.5-40	9,195
25	256.6	15.6089	862	100	6,066	127	1.4-40	9,103
*26	1026.4	15.6089	862	100	23,606	490	3.0-40	9,548
*27	1026.4	15.6089	860	100	23,606	490	3.0-40	9,548

*4-15 MWt Trains

TABLE 3-3

MASS & VOLUMES
WATER & STEAM

*f = Fluid
sf = Saturated Fluid
sfg = Saturated Fluid & Steam
sg = Saturated Steam
SHg = Superheated Steam

Line No.	Total Flow (lbm/Min)	Condition*	Pressure (psia)	Temperature °F	Density (lbm/Ft^3)	Flow Ft^3/Min)	Pipe Size Dia-Sch.	Velocity (FPM)
A	395.2	f	50	100	61.99	6.38	1.5-40	413
B	269.8	f	50	100	61.99	4.35	1.25-10	384
C	395.2	f	47	220	59.63	6.63	1.5-10	429
D	269.8	f	47	220	59.63	4.52	1.25-10	399
E	665.0	f	47	220	59.63	11.15	2.0-10	440
F	678.6	f	30	240	59.10	11.48	2.5-40	345
G	678.6	f	915	240	59.10	11.48	2.5-40	345
H	678.6	f	912	395	53.85	12.60	2.5-40	379
I	678.6	f	909	492	49.41	13.73	2.5-40	413
J	6,786.0	sf	905	532	46.95	144.5	4x4-40	409
K	6,786.0	sfg	905	532	14.41	470.9	4x6-40	587
L	678.6	sg	905	532	1.989	341.2	3.0-40	6,661
M	13.6	sg	905	532	1.989	6.84	.375-40	5,155
N	13.6	SHg	30	240	0.061	221.9	1.5-10	14,378
O	665.0	sg	905	532	1.989	334.3	3.0-40	6,515
P	665.0	SHg	900	730	1.405	473.4	3.0-80	10,330

TABLE 3-4

MASS & VOLUMES
CONDENSATE

Line No.	Total Flow (lbm/Min)	Pressure (psia)	Temperature °F	Density (lbm/Ft³)	Flow Ft³/Min)	Pipe Size Dia-Sch.	Velocity (FPM)
C1	59.9	871	426	52.53	1.14	0.75-40	493
C2	122.2	869	354	55.44	2.20	1.0-40	367
C3	182.1	869	378	54.54	3.34	1.25-40	322
C4	44.8	867	265	58.38	0.76	0.5-40	360
C5	227.0	867	355	55.40	4.10	1.5-40	292
C6	227.0	865	130	61.54	3.69	1.5-40	264
C7	13.4	865	130	61.54	0.21	0.25-40	291
C8	240.3	865	130	61.54	3.90	1.5-40	278
C9	243.4	865	130	61.54	3.96	1.5-40	280
*C10	973.8	865	130	61.54	15.82	2.5-40	476
*C11	977.1	865	125	61.65	15.85	2.5-40	477

*Condensate Drum to Pipeline

stainless steel, type 304, except for syngas piping from the pipeline to the No. 1 Methanator, which is carbon steel. Material of construction for the components are as follows:

a. Desulfurizers - Constructed of SA-335 (2-1/4 Cr-1Mo) steel. The perforated metal grate is made of the same material or SS-304L.

b. No. 1 Adiabatic Methanator - Constructed of SA-335 steel with internal SS-316 cladding, 0.188 in. thick.

c. No. 2 Isothermal Methanator/Evaporator - Constructed of SA-335 steel for the shell, end caps and tube sheets with 0.188 in. thick internal SS-316 cladding. Tubes are SS-304L.

d. No. 1, 2 and 3 Syngas Heaters - End caps constructed of SA-335 steel with 0.108 in. thick SS-316 cladding. Shell is SA-335 steel with no cladding. Tubes and tubesheets are SS-304L.

e. No. 1C, 1G, 2 and 3 Feedwater Heaters - Constructed in total of SS-304L.

f. Steam Drum - Constructed of SA-335 steel with 0.188 in. thick SS-316 internal cladding. Internals are made of SS-304L.

g. Steam Superheater - Constructed of SA-335 steel with 0.188 in. thick SS-316 cladding for the shell, end-caps and tube sheets. Tubes are SS-304L.

h. All other components in contact with water or steam are constructed of SS-304L.

3.4 Support System Components

The components comprising the methanation process support system are used to individually startup and shutdown the four-15 MWt Methanation Trains. The support system is also used to maintain from one to all four trains in hot standby operation. A complete list of all components and their quantities are given in Table 3-5 and they are identified by their capacity or functional requirements. These are commercial components selected by Bechtel. Figure 3-3 shows the support system components and their connections to the four methanation trains.

3.5 Modular Construction

Each 15 MWt Methanation Train is to be assembled into five modules which are RR transportable. A module includes steel structures, components, component supports, electrical and instrument cable trays and piping. See Figure 3-17.) Modular construction permits a total methanation train to be assembled and operated (cold checkout) at the fabrication plant. For transporting, only uncoupling of pipes and cable which connect one module to another is required. The connecting pipes will be shipped with the modules for reassembly of the train at the methanation plant site. Conceptual design of the module structures by Bechtel show sufficient detail to permit a reasonably accurate (±20%) estimation of its cost.

3.6 Plant Arrangement Design

The plant arrangement design is guided by appropriate codes, standards and safety guidelines. The control room and maintenance building are located in the center of the four methanation trains. The support system is centrally located in relation to the methanation trains, which minimizes the length of pipe and cable runs from the support system to the trains. The 60 MWt Methanator Plant design is shown in Figures 3-18, 3-19 and 3-20.

TABLE 3-5

60 MWt Methanation Plant Support System
Equipment List

ITEM	QTY	DESCRIPTION
1. Gas-fired Heater for Desulfurizers	1	
Heat Duty		2.5×10^6 Btu/hr
Dimensions		72" Sq. x 264" H
2. Gas-fired Heater for Methanators	1	
Heat Duty		7.5×10^6 Btu/hr
Dimensions		96" Sq. x 360" H
3. Gas-fired Boiler for Feedwater	1	
Heat Duty		5.0×10^6 Btu/hr
Dimensions		96" Sq. x 360 " H
4. Dryer	2	1-100% + 1 spare
Type		TEG-Scrubbing package unit with absorber & regenerator
Operating Temp. & Press.		100°F/862 psia
Outlet Moisture		7 lbs $H_2O/10^6$ scf
Reboiler Duty		150,000 Btu/hr-max.
Dimensions		
Absorber		36" Dia. x 180" H
Regenerator		30" Dia. x 84" H

TABLE 3-5 (Con't)

5.	Condensate Drum	1	
	Capacity		300 gallons
	Temp./Press. (Design)		150°F/900 psia
	Dimensions		36" Dia. x 72" L w/1.0" thick shell
6.	Nitrogen Storage	1	Liquid storage tank with necessary vaporizers and controls.
	Holding Capacity		7.5 x 10^5 scf storage tank
	Dimensions		108" Dia. x 360" H
7.	Hydrogen Supply	-	Valves and controls as required to unload RR tank car. Required for about 4 weeks every 2.5 years.
*8.	Water Treatment Plant	1	
	Capacity (max)		35 gal/hr
9.	Odorizer	1	Mixture of mecaptans, hydrogen sulfide and/or carbon sulfide.
	Capacity (max)		12.5 scf/hr
*10.	Emergency Power Supply	1	
	Type		Diesel generator, package unit, fuel used - 450 gal/day
	Power output		350 kW
	Dimensions		40" W x 120" L x 60" H

TABLE 3-5 (Con't)

*11.	Control Room Panels, controls and site electrical distribution system, including lighting	As required.

*Not shown on Flow Diagrams

4. CAPITAL AND OPERATING COST ESTIMATES

4.1 Capital Costs for the 60 MWt Methanation Plant

Where possible the capital cost for the 15 MWt Methanation Train equipment is scaled from the 40 MWt Methanation System of last year, reported in Reference 1. The basic reference for equipment cost estimates is dollars/pound of material, fabricated and assembled into modules. The material values used are:

a. SA-335 steel at $1.55/lb.

b. SS-304L at $1.79/lb.

c. SS-316 at $2.65/lb. (used only for cladding).

d. Thermal insulation at $9.35/ft^3.

e. Desulfurizer catalyst:

   1) Cobalt-Molybdenum at $72.00/ft^3

   2) Zinc-Oxide at $105.00/ft^3

f. Methanation catalyst at $900.00/ft^3.

Table 4.1 presents the capital costs of material, labor and modular assembly of the major components of one 15 MWt Methanation Train. Table 4.2 gives the commerical components needed for one 15 MWt Methanation Train. These items were cost estimated by Bechtel. Table 4.3 lists all the components of the Support System which were also cost estimated by Bechtel. The Dryer, Condensate Drum and Odorizer may be assembled into a common module, but the remaining components of the support system are too large for modular assembly.

4.1.1 Other Capital Cost Information

All capital costs are based on building 14-60 MWt Methanation Plants which requires 56-15 MWt Methanation Trains and 14 Support Systems. The construction time is based on installation and startup or one plant or more per month (total time $\leq$ fourteen months). The following capital costs were estimated by Bechtel:

TABLE 4.1

CAPITAL COSTS
MATERIAL, LABOR & MODULAR ASSEMBLY
15 MWt METHANATION TRAIN COMPONENTS

Component		Qty	Costs x $000 (By GE)
1.	Desulfurizer	2	80.2*
2.	Adiabatic Methanator	1	57.0*
3.	Isothermal Meth./Evap.	1	66.4*
4.	Steam Drum	1	21.9
5.	No. 1 Syngas Heater	1	12.0
6.	No. 2 Syngas Heater	1	12.3
7.	No. 3 Syngas Heater	1	10.7
8.	Steam Superheater	1	35.2
9.	No. 1C Feedwater Heater	1	27.1
10.	No. 1G Feedwater Heater	1	53.0
11.	No. 2 Feedwater Heater	1	47.6
12.	No. 3 Feedwater Heater	1	12.4
13.	Deaerator	1	5.4**
14.	Air Cooler	1	11.3**
15.	K.O. Drum w/Flare Stack	1	23.2**
		Subtotal	$475.7

* Includes cost of catalyst.

**Scaled from 40 MWt Methanation Train.

TABLE 4.2

CAPITAL COSTS

COMMERICAL COMPONENTS & MODULAR ASSEMBLY

15 MWt METHANATION TRAIN

Component		Qty	Costs x $000 (By Bechtel)
1.	No. 1 Recycle Gas Compressor Capacity: 54,420 lbm/hr Temp/Press In: 651°F/886 psia Temp/Press Out: 651°F/892 psia Units: 2+1 (50% spare)	3	125.0
2.	No. 2 Recycle Gas Compressor Capacity: 23,400 lbm/hr Temp/Press In: 734°F/875 psia Temp/Press Out: 734°F/890 psia Units: 2+1 (50% spare)	3	95.0
3.	Feedwater Pump Capacity: 50 gpm-ea. (max) Temp/Press In: 240°F/30 psia Temp/Press Out: 240°F/915 psia Units: 2+1 (50% spare)	3	16.0
4.	Steam Pressure Relief Valve Capacity: 840 lbm/hr Temp/Press In: 532°F/905 psia Temp/Press Out: 240°F/30 psia	1	12.0
5.	Allowance for Misc. Equipment		67.0

TABLE 4.3

CAPITAL COSTS
MATERIAL, LABOR AND INSTALLATION
METHANATION PROCESS SUPPORT SYSTEM COMPONENTS

Component	Qty	Costs x $000 (By Bechtel)
1. No. 1 Startup Gas Heater Heat Duty: 2.5×10^6 Btu/hr, gas fired No. of Units: 1	1	132.0
2. No. 2 Startup Gas Heater Heat Duty: 7.5×10^6 Btu/hr, gas fired No. of Units: 1	1	212.0
3. Startup Feedwater Heater Heat Duty: 5.0×10^6 Btu/hr, gas fired No. of Units: 1	1	164.0
4. Product Gas Dryer Type: Sivalls Model DHT-30-250-2101 Operating Temp. & Press: 100°F/865 psia Outlet Moisture: 7 lbs $H_2O/10^6$ scf Reboiler Duty: 150,000 Btu/hr No. of Units: 1+1 (100% spare)	2	132.0

TABLE 4.3 (Continued)

CAPITAL COSTS
MATERIAL, LABOR AND INSTALLATION
METHANATION PROCESS SUPPORT SYSTEM COMPONENTS

Component		Qty	Costs x $000 (By Bechtel)
5.	Product Gas Odorizing Unit Metering pumps: 2 x 100% Rate: 1.07 lb/hr mercaptans Storage: 90 days 2,400 lbs (liquid) 43 ft^3 Note: 1 unit serves combined output of 4 Methanation Trains	2	14.0
6.	Condensate Drum Capacity: 300 gallons Temp/Press: 150°F/900 psia No. of Units: 1	1	10.0
7.	Startup Nitrogen Storage Package unit with recycle pump and controls Holding capacity: 7.5×10^5 scf No. of Units: 1	1	263.0
8.	Hydrogen Supply Station Tanker connections with valves and controls	1	13.0

TABLE 4.3 (Continued)

CAPITAL COSTS

MATERIAL, LABOR AND INSTALLATION

METHANATION PROCESS SUPPORT SYSTEM COMPONENTS

Component	Qty	Costs x $000 (By Bechtel)
9. Water Treatment Plant Capacity: 35 gal/hr	1	20.0
10. Emergency Power Supply System Type: diesel generator Output: 350 kW	1	36.0

a. Direct Field Costs. Since each methanation train is to be assembled into modules and operated (cold check-out) at the fabrication plant a minimum of field labor is required.*

b. Indirect Field Costs. A percentage of the direct field costs.

c. Engineering & Fee. Since only one standard plant is to be built the cost of engineering and fee is equally amortized over the 14 plants and 56 trains.

d. Contingency. With modular construction and factory assembly and checkout the contingency is estimated to be 16% of the total direct costs.

*Field labor is needed for the following:

1. Site preparation.
2. Install modules and very large components.
3. Connect (pipe, electrical and instrument cables) modules and support system to modules.
4. Construct control room.
5. Construct maintenance and storage building.
6. Acceptance testing of site hookups.

4.1.2 Capital Cost Summary

The total capital cost estimated by Bechtel is \$11 x 10^6 per 60 MW_t methanation plant, when 14 plants are built in a short time period using modular construction. The detailed Bechtel estimate is presented in Appendix G.

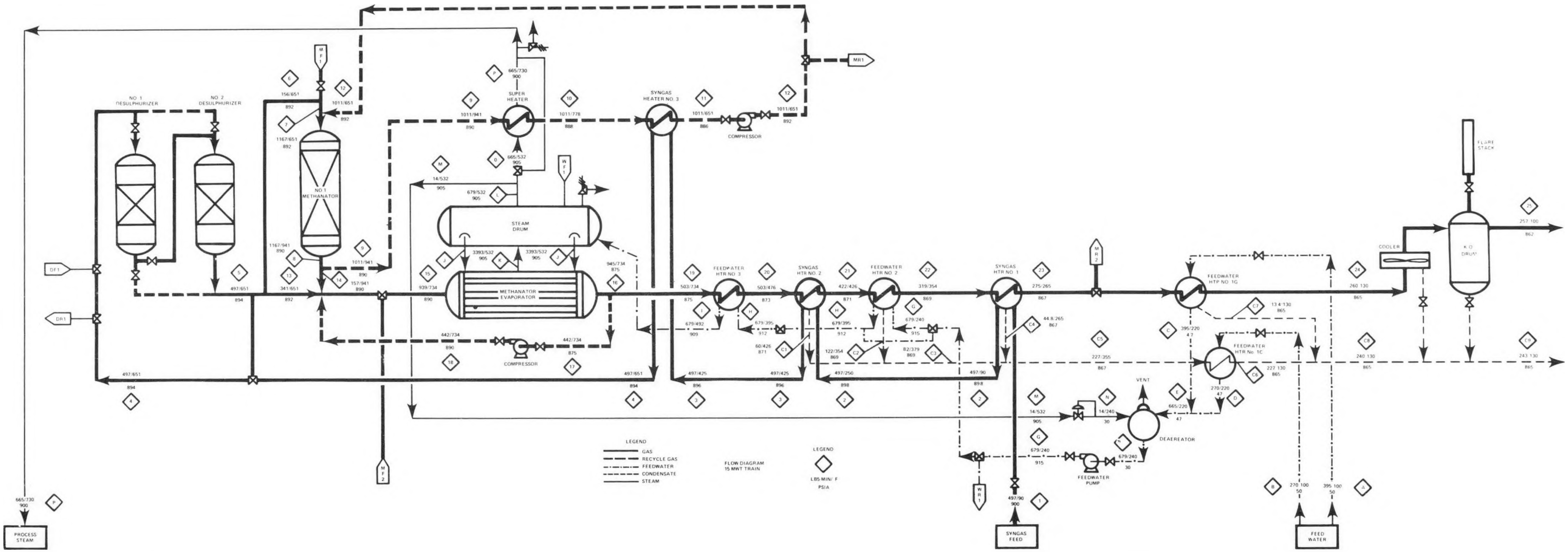
4.2 Operating and Maintenance Costs

The operating and maintenance cost estimates for the 60 MW_t methanation plant is an update of the operating cost for the 40 MW_t methanation plant described in Reference 1. Since there are four trains instead of one, it is estimated that the operating labor needs to be increased by 50%. See Table 4.4.

TABLE 4.4

OPERATING & MAINTENANCE COSTS - 60 MWt METHANATION PLANT
($/yr)

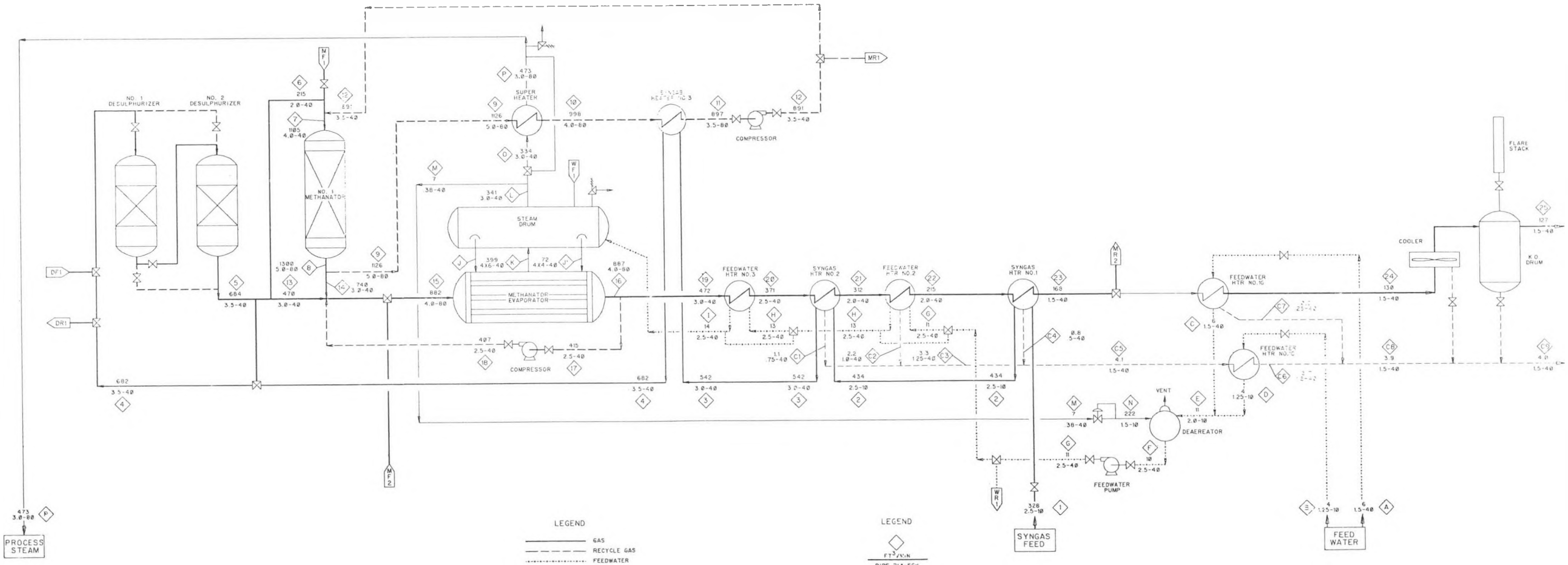
		Costs
	Direct Costs	
1.	Electrical Power ($0.04/kWH)	$216,430
2.	Make-up Water ($0.40/1,000 gal)	2,500
3.	Catalyst Replacement:	
	Desulfurizers:	210,000
	Methanators:	92,160
4.	Operating Labor (Supt. +12 operators)	331,000
5.	Maintenance Labor (3% Const. Cost)	239,200
	Indirect Costs	
6.	Administrative and Support Labor (30% of 4 & 5)	171,060
7.	General and Administration Expense (60% of 4 & 5)	342,120
8.	Property Taxes and Insurance (2.5% of Total Plant Investment)	275,000
	TOTAL O & M Cost	$1,870,470



FLOW DIAGRAM 15 MWt TRAIN

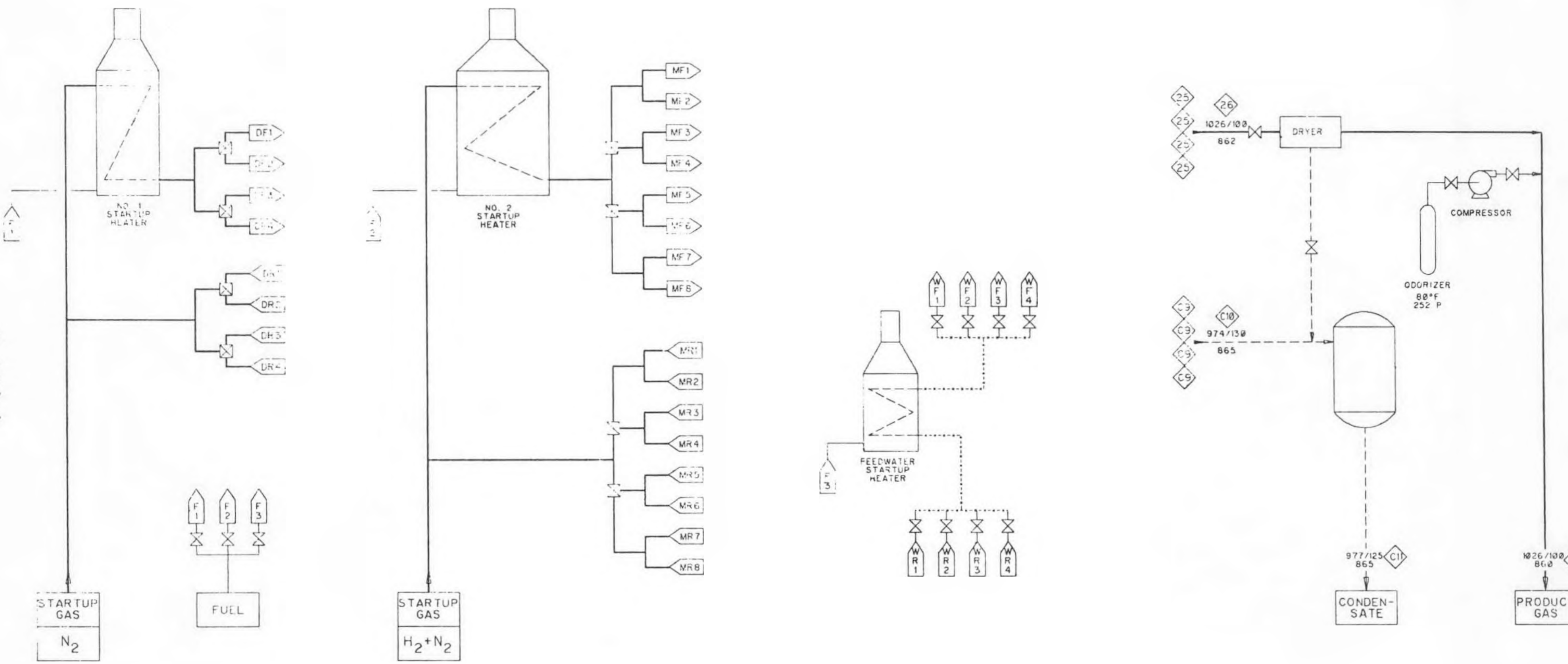
82-429-13

FIGURE 3-1



PIPING DIAGRAM
15MWT TRAIN

FIGURE 3-2



SUPPORT SYSTEM
60MWT PLANT

FIGURE 3-3

DE-SULFURIZER 1 & 2 15 MWt TRAIN

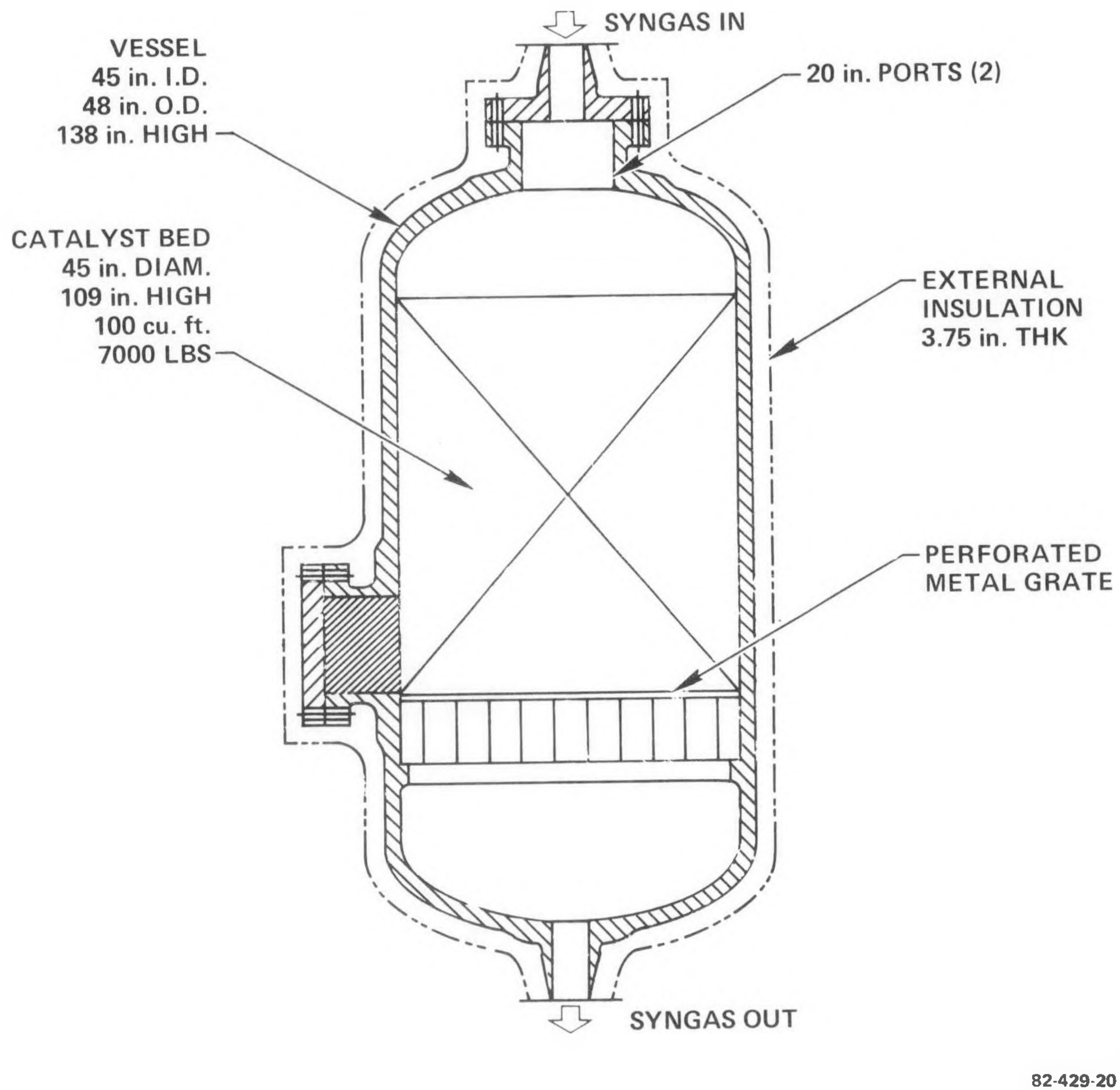
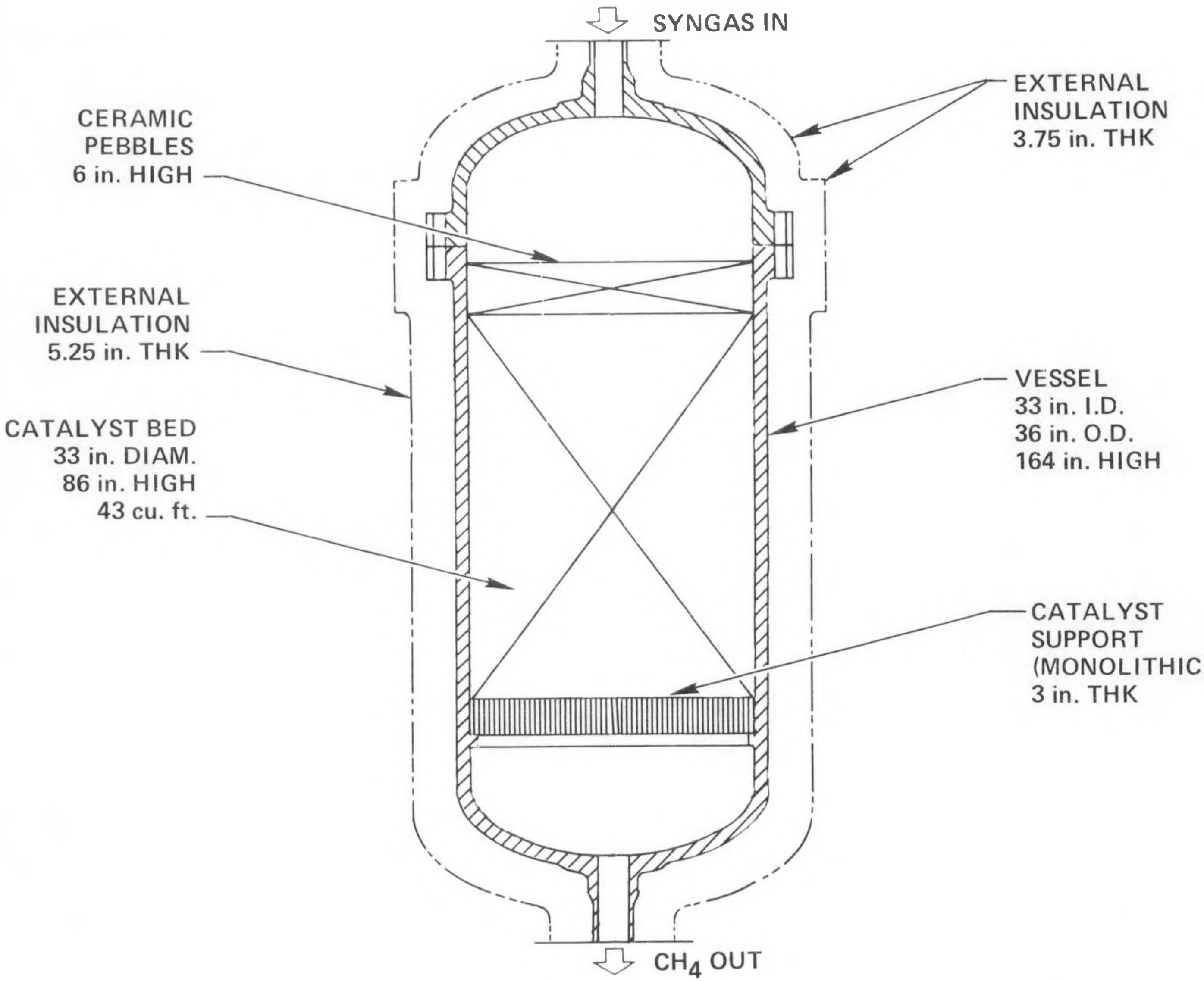


FIGURE 3-4

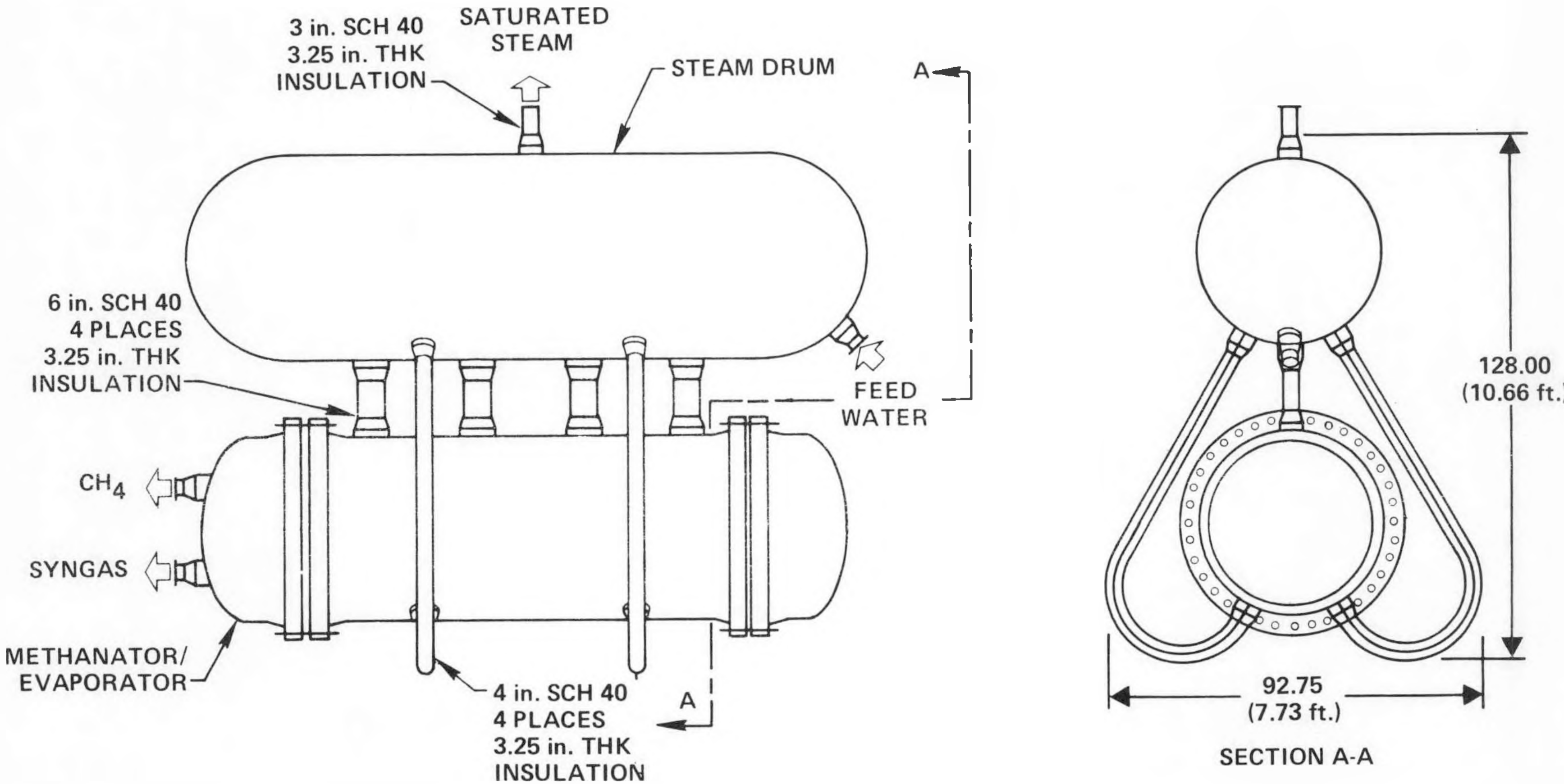
ADIABATIC METHANATOR 15 MWt TRAIN



82-429-15

FIGURE 3-5

METHANATOR/EVAPORATOR STEAM DRUM ASSEMBLY 15 MWt TRAIN



82-429-16

FIGURE 3-6

ISOTHERMAL METHANATOR/EVAPORATOR 15 MWt TRAIN

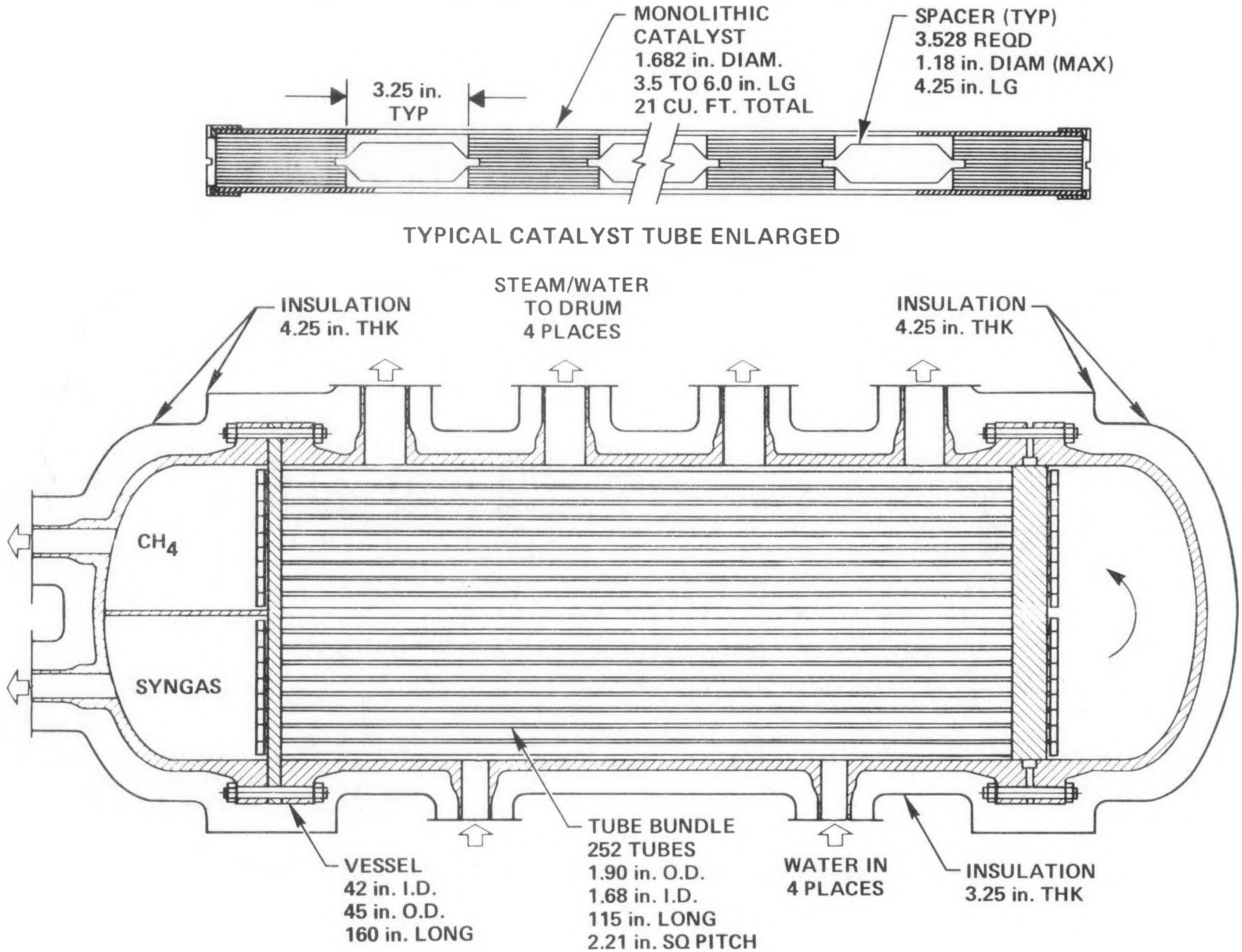


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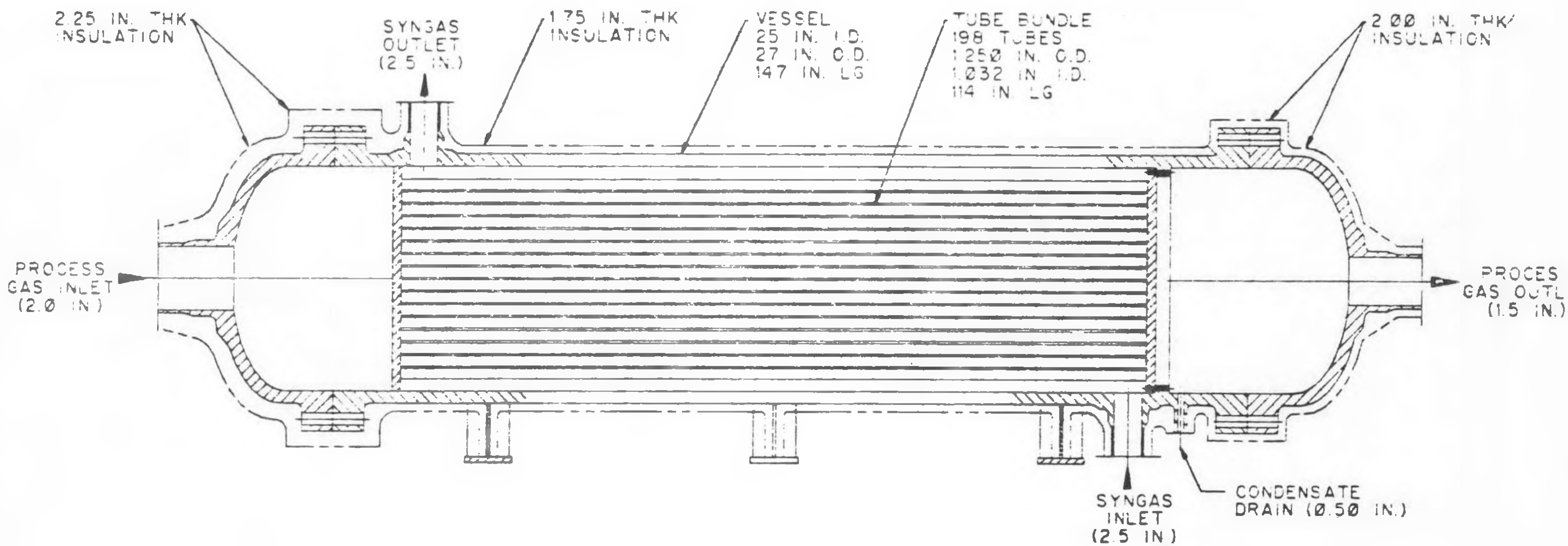
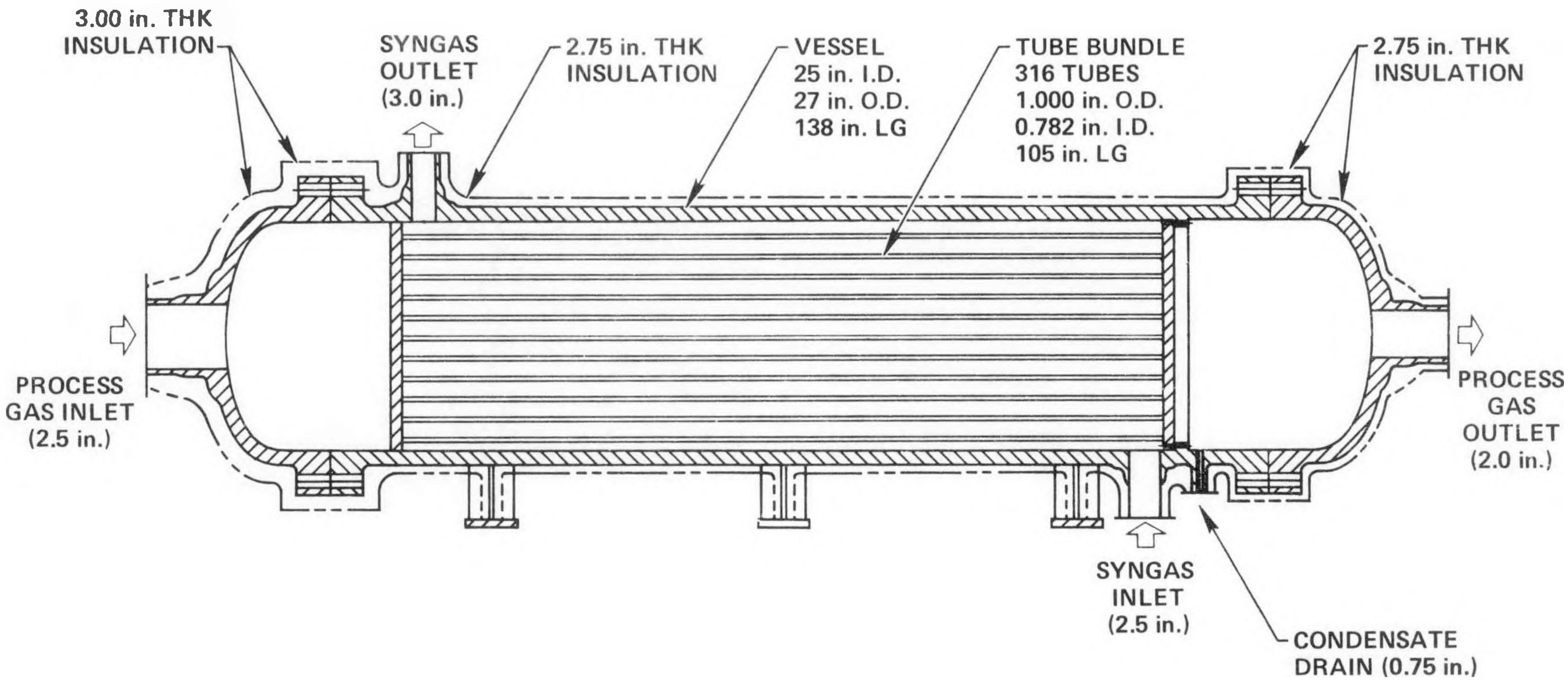
FIGURE 3-7



SYNGAS HEATER NO.1
15MWT TRAIN

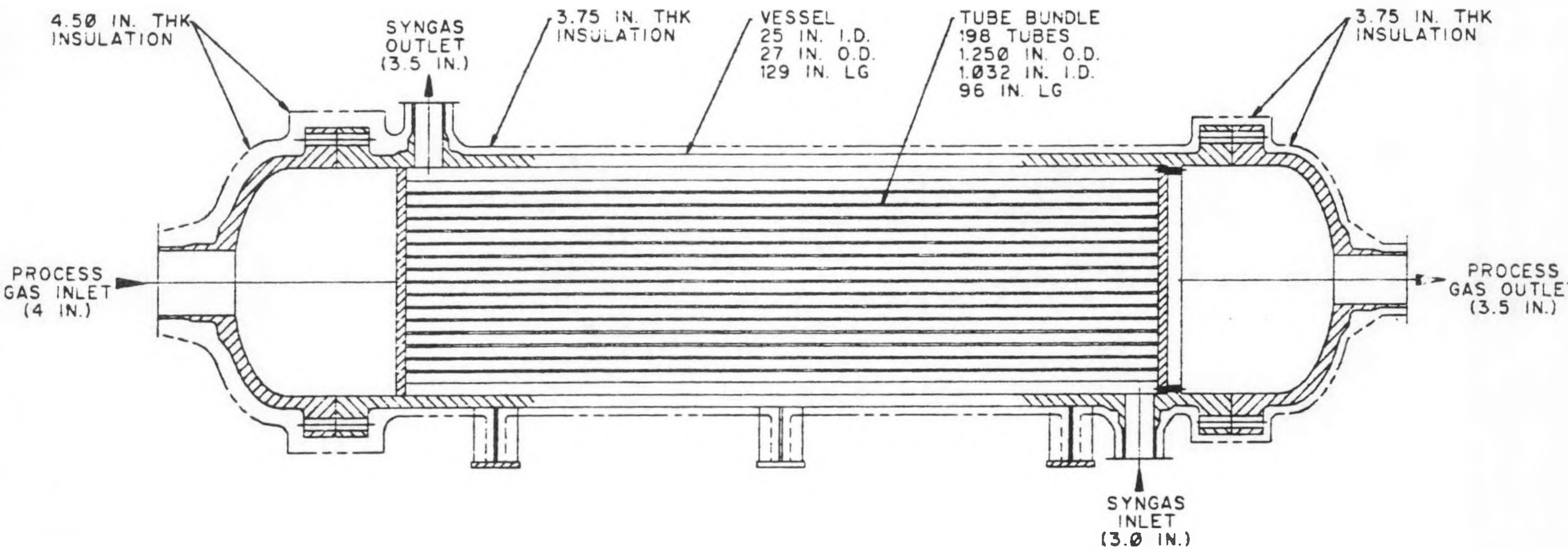
FIGURE 3-8

SYNGAS HEATER NO. 2 15 MWt TRAIN



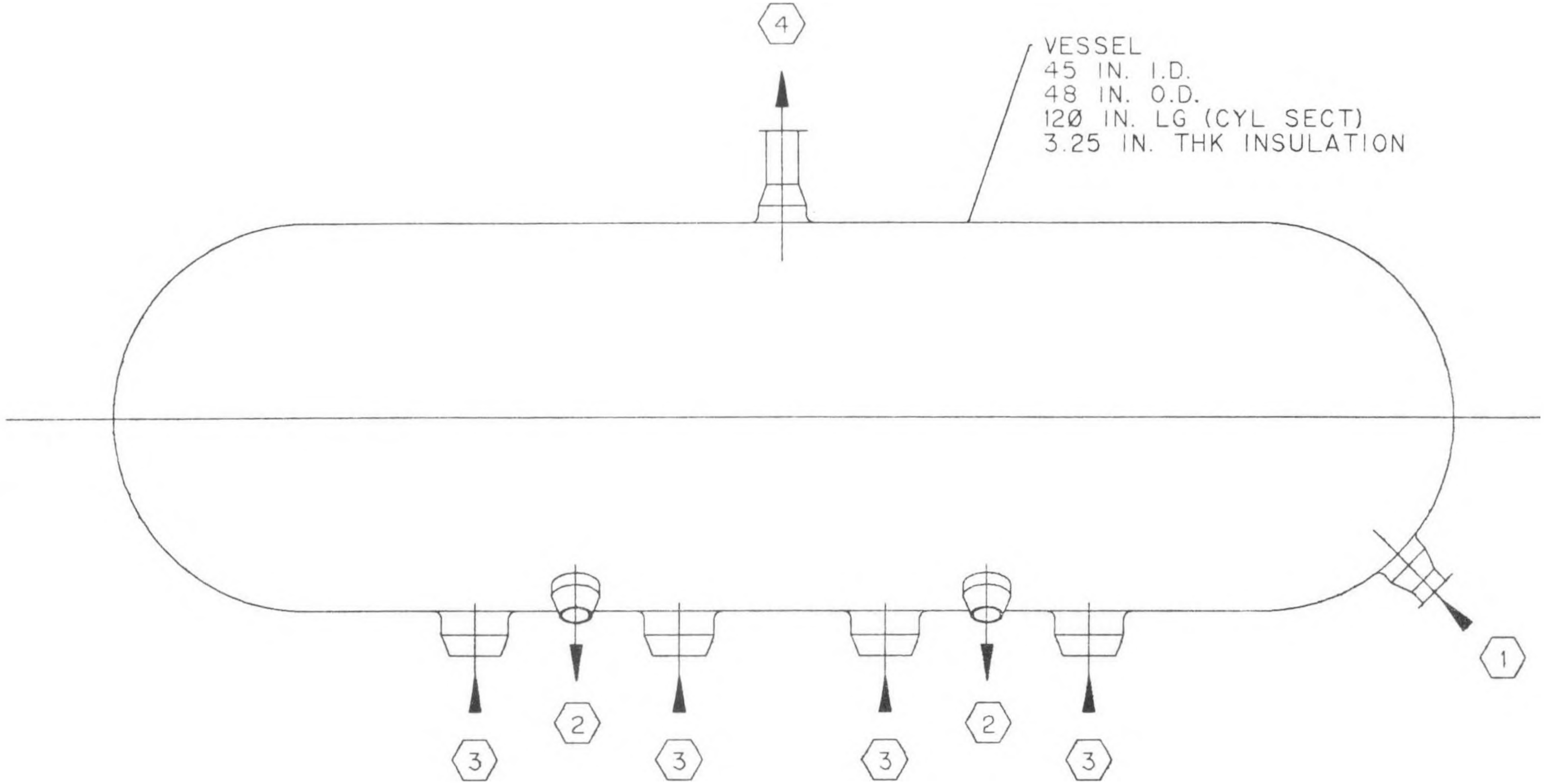
82-429-21

FIGURE 3-9



SYNGAS HEATER NO.3
15MWT TRAIN

FIGURE 3-10



1 FEEDWATER INLET (1)

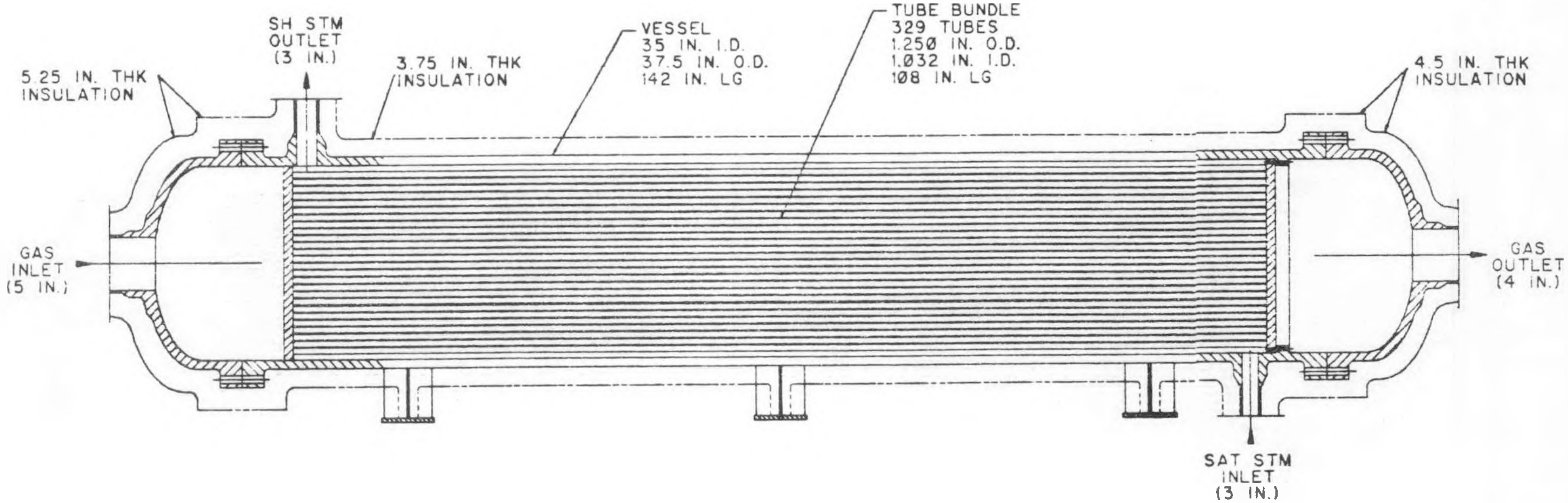
2 WATER FEED TO EVAPORATOR (4)

3 STEAM/WATER FROM EVAPORATOR (4)

4 SATURATED STEAM TO SUPERHEATER (1)

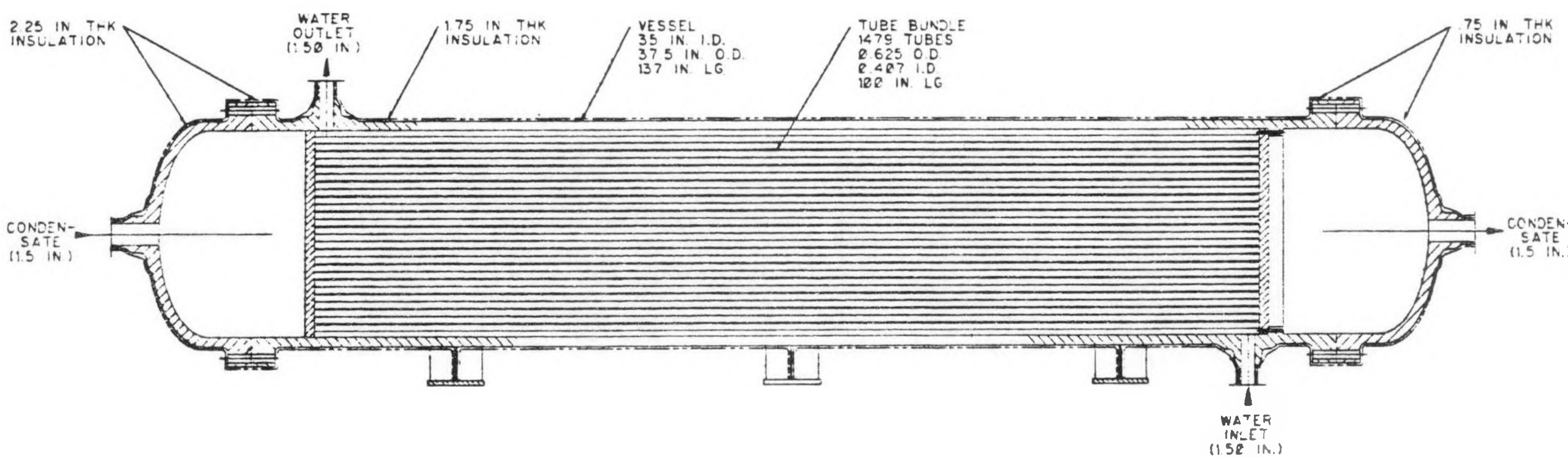
STEAM DRUM
15MWT TRAIN

FIGURE 3-11



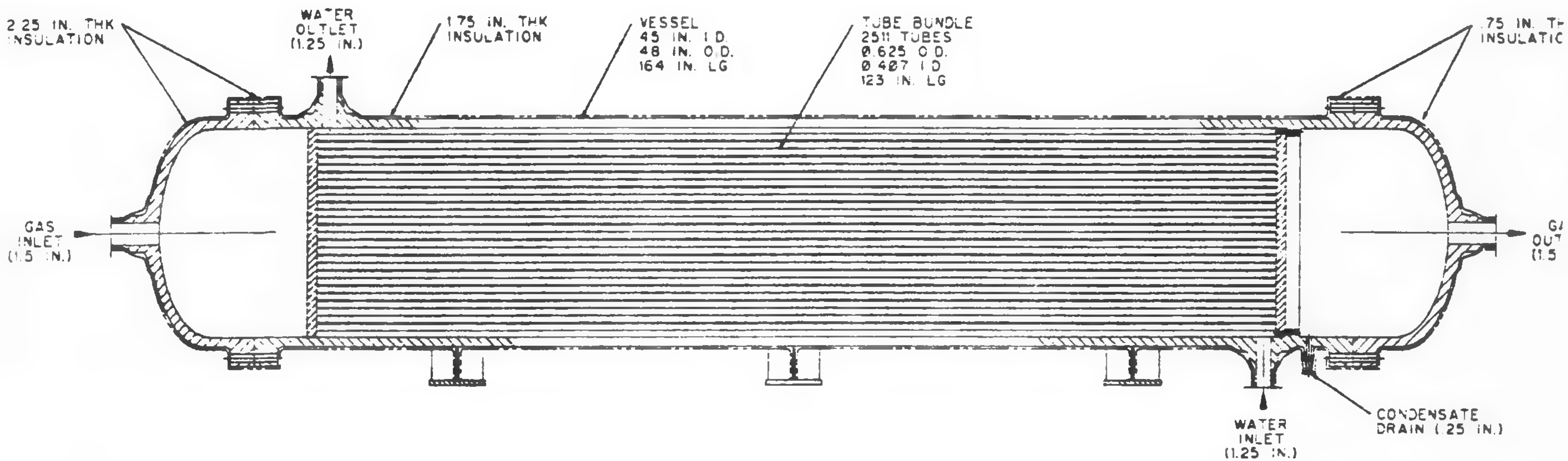
SUPERHEATER
15MWT TRAIN

FIGURE 3-12



FEEDWATER
HEATER NO.1C
15MWT TRAIN

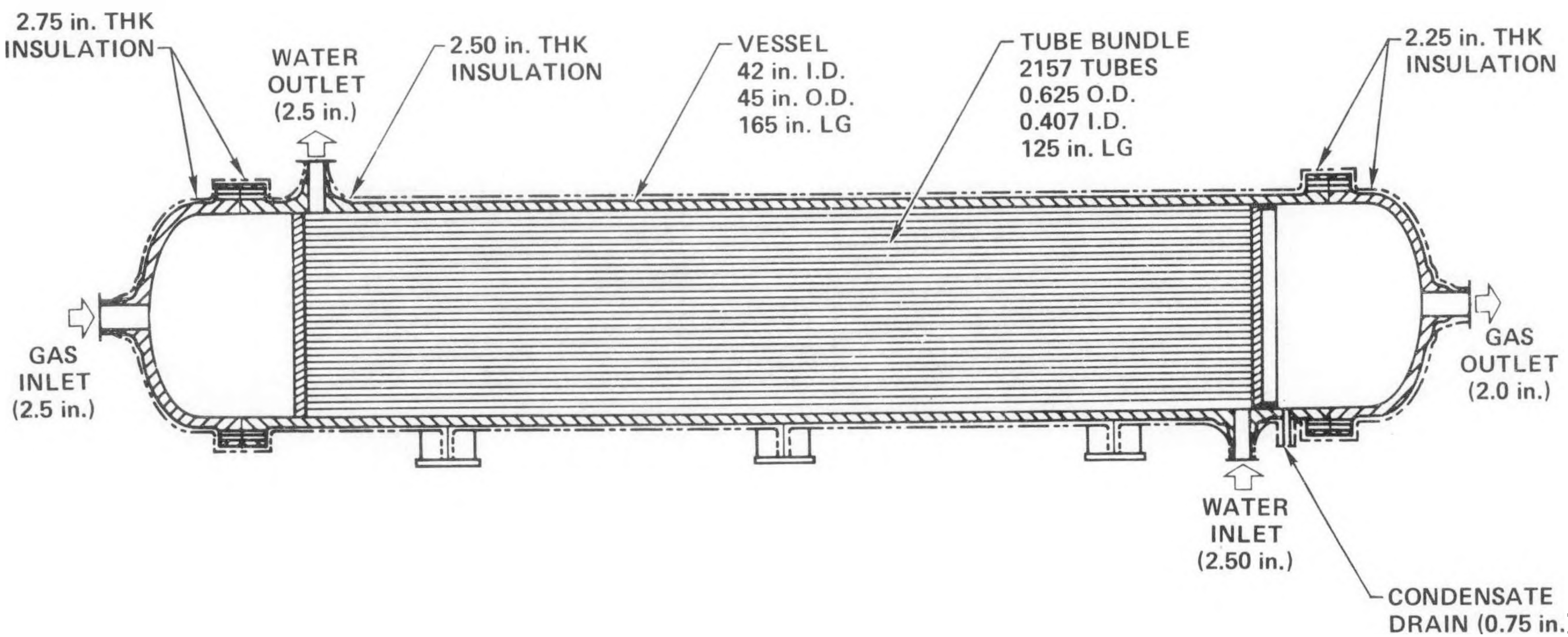
FIGURE 3-13



FEEDWATER
HEATER NO.1G
15MWT TRAIN

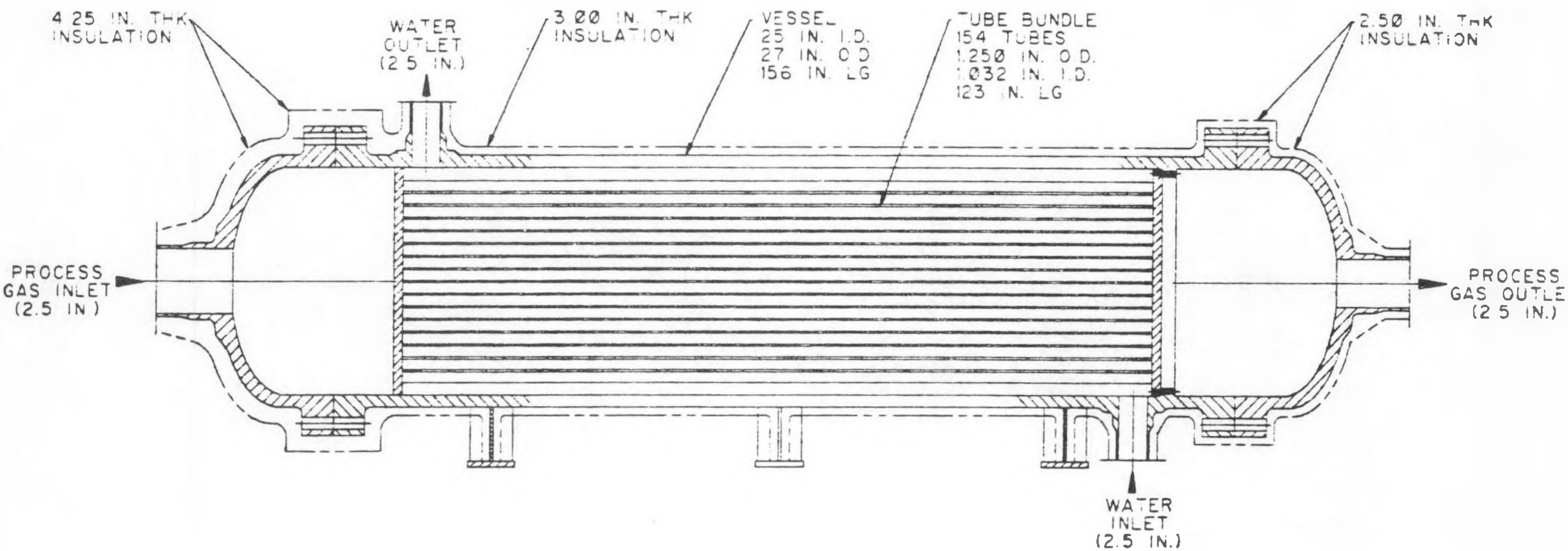
FIGURE 3-14

FEEDWATER HEATER NO. 2 15 MWt TRAIN



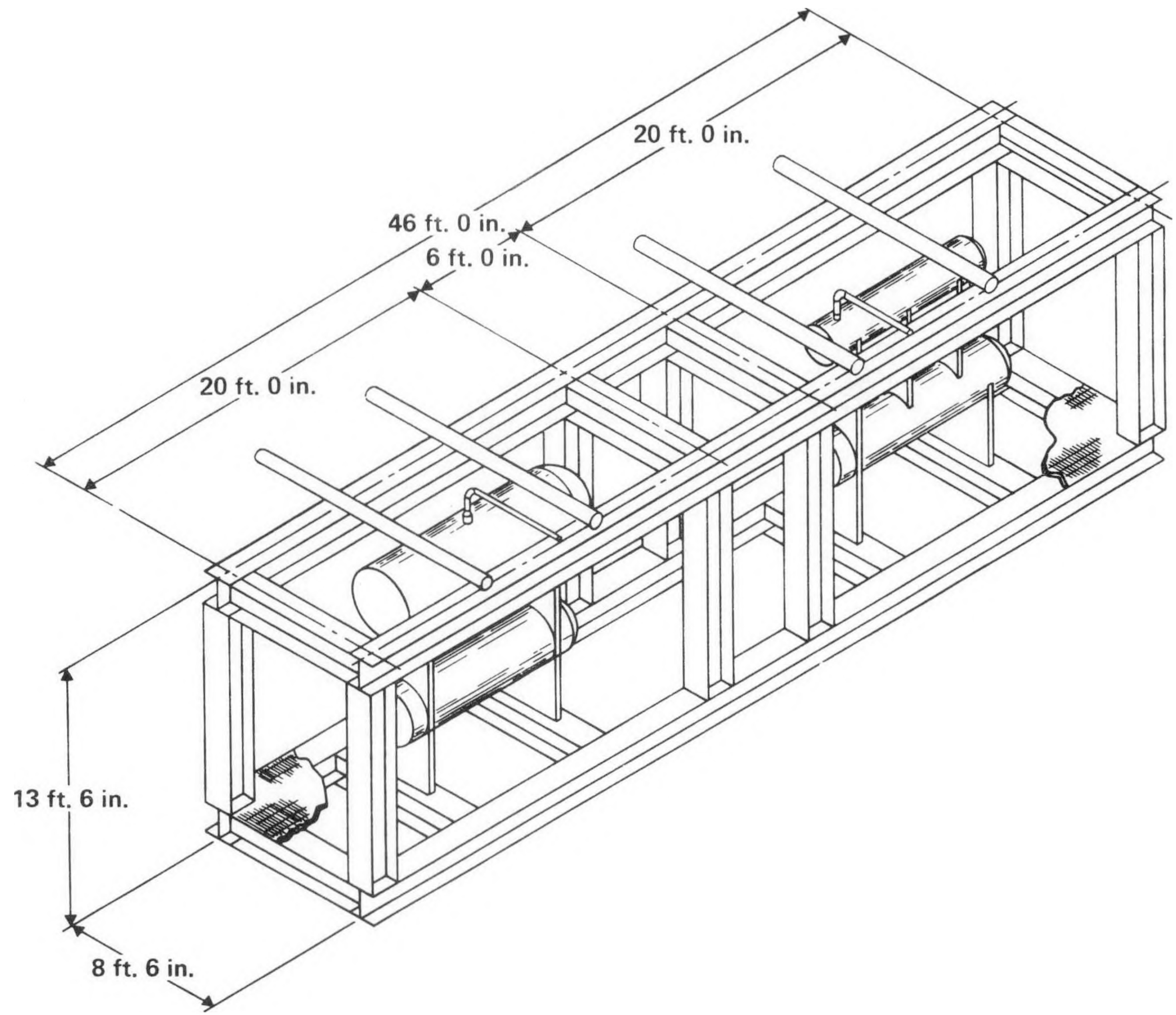
82-429-19

FIGURE 3-15



FEEDWATER
HEATER NO.3
15MWT TRAIN

TYPICAL EQUIPMENT MODULE



82-429-14

FIGURE 3-17

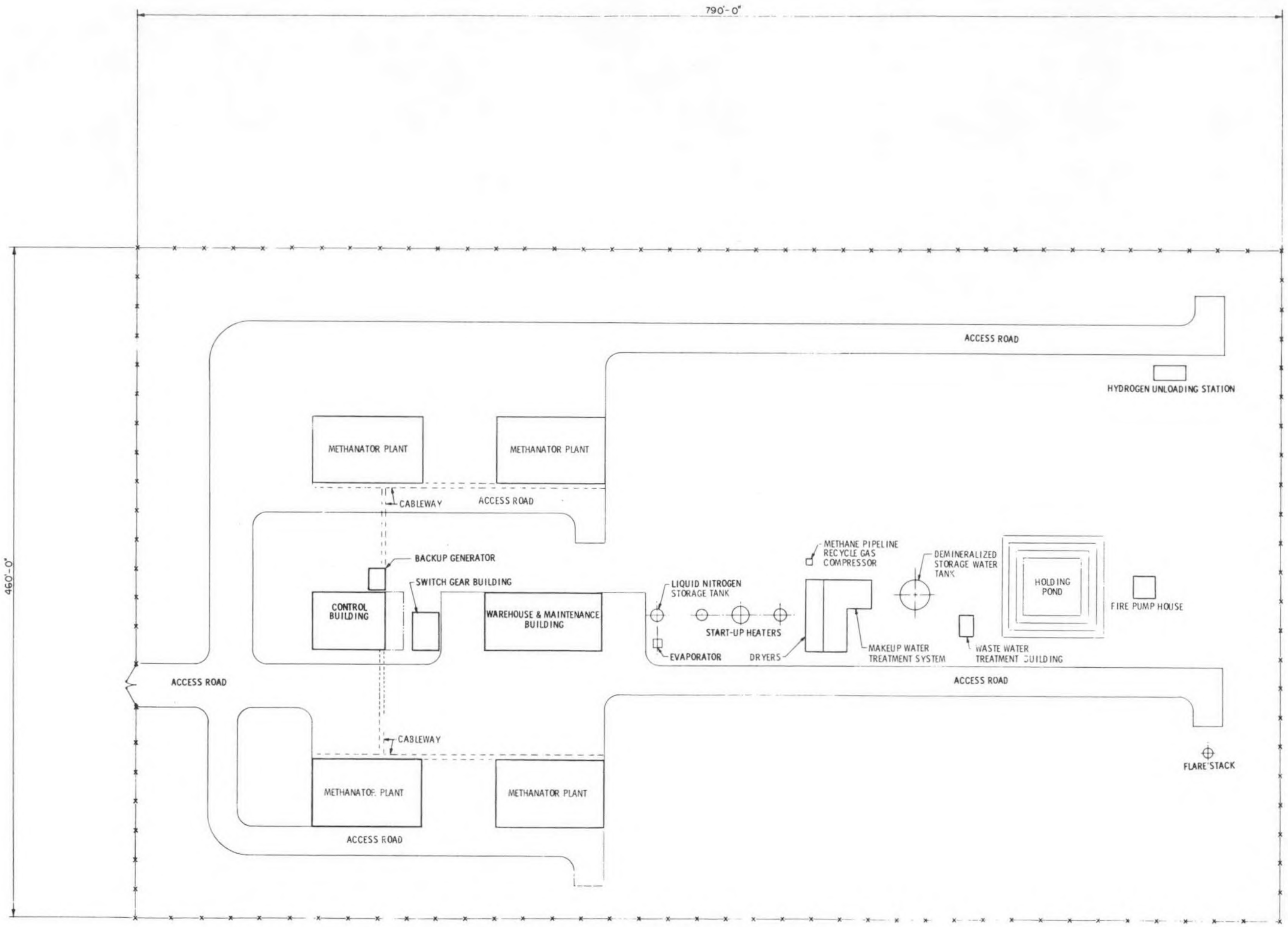


FIGURE 3-18

No.	DATE	REVISIONS	BY	CHKD	DESIGN SUPV	ENGR	PROJ ENGR	APPR
1	4/23/82	..D FOR FINAL REPORT						
A	4/8/82	PROGRESS PRINT						

SCALE 1" = 30'-0" | DESIGNED | DRAWN SMETHURST | CHIEF ENGR

BECHTEL GROUP, INC.
SAN FRANCISCO

G.E./HTGR R
METHANATOR STUDY

60 MW METHANATOR PLANT
PLOT PLAN

JOB No.	DRAWING No.	REV.
14508	C-4002	1

34 x 44 "E" SIZE

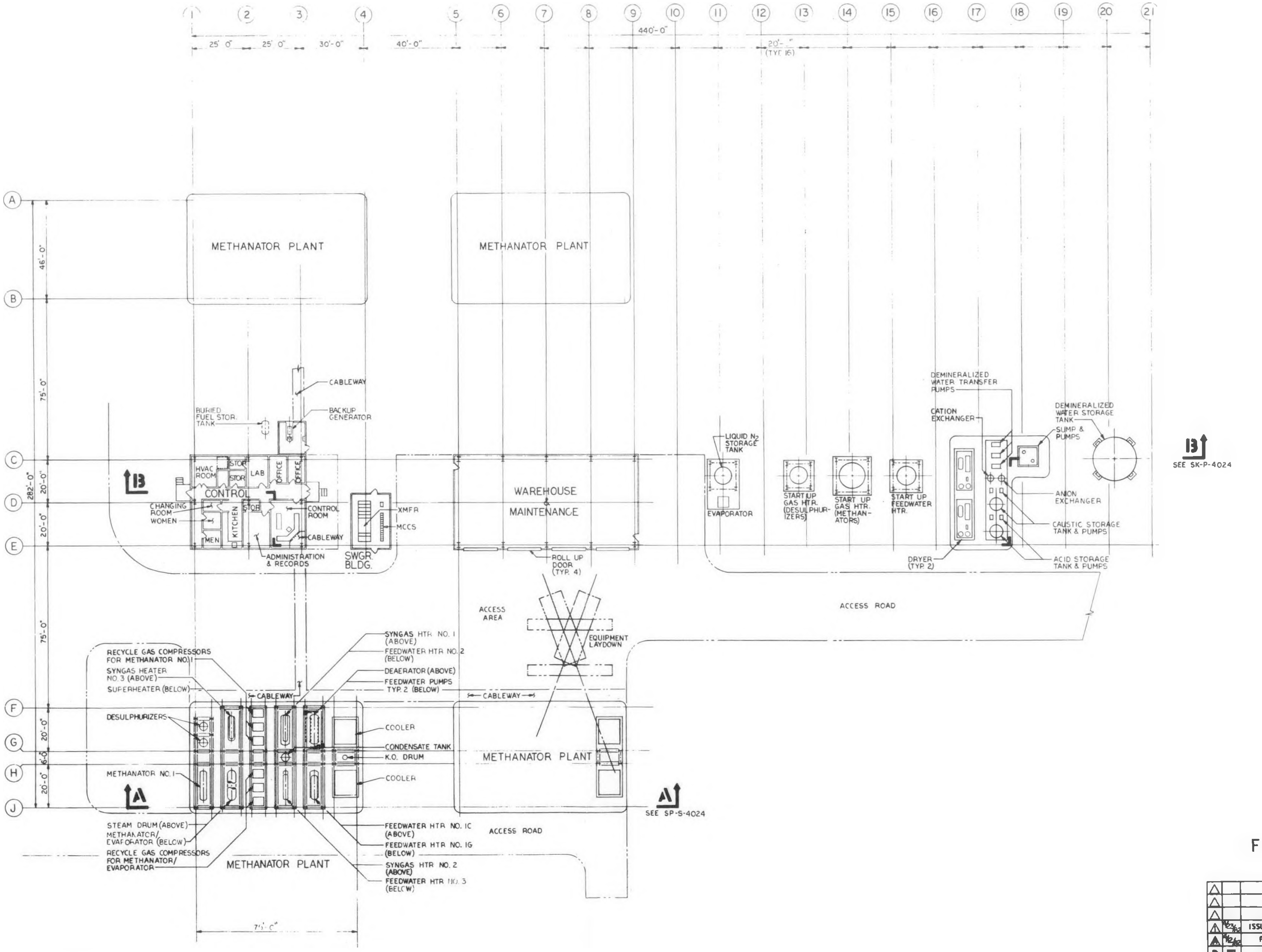
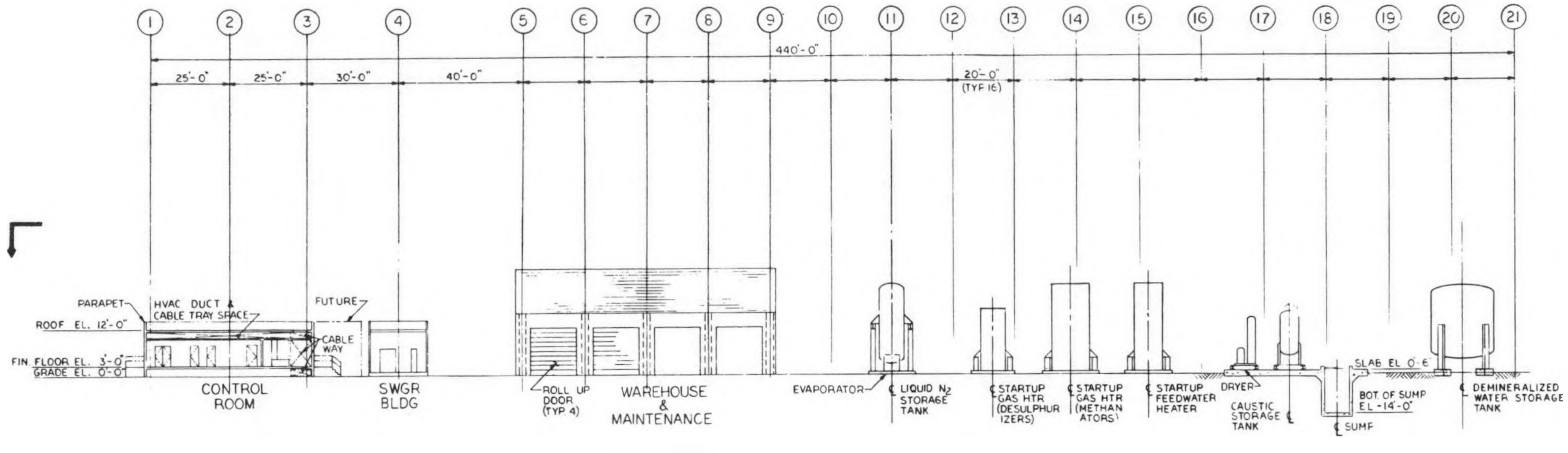


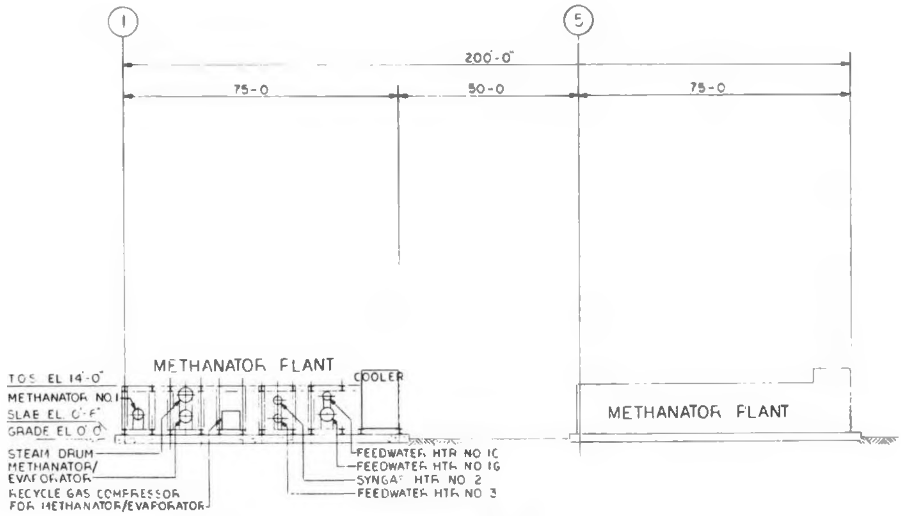
FIGURE 3-19

	ISSUED FOR FINAL REPORT	
	PROGRESS PRINT	

BECHTEL GROUP, INC.
SAN FRANCISCO
GE / HTGR R METHANATOR STUDY
60 MW METHANATOR PLANT PLAN BELOW EL. 50'-0"
14508 P-4021 1



SECTION B-B



SECTION A-A

FIGURE 3-20

1	4/23/82	ISSUED FOR FINAL REPORT	3/1/82
A		PROGRESS PRINT	3/1/82

SCALE 1/16" = 1'-0"

SMETHURST

BECHTEL GROUP, INC.
SAN FRANCISCO

GE / HTGR R
METHANATOR STUDY

60 MW METHANATOR PLANT
SECTIONS A-A & B-B

JOB No.	DRAWING No.	REV.
14508	P-4024	1

APPENDIX A

Application of Methanators to Industrial Steam

by B. A. Hutchins

There are a number of market reasons for continued emphasis on methanatior designs to serve the industrial process steam user.

1. Steam Requirements are Prodominantly Below 500°F

   Table 1 lists the fractions of industrial process steam demand (boiler output) in five temperature ranges as generated recently by GEA for four utility service areas and the nation. The four utilities are those chosen to give a cross-section of the U.S. for future detailed, site-specific studies. As seen in Table 1, only Northern Indiana P.S. has more than 10 percent of the demand above 500°F, and this is likely due to the large number of steel mills in that region.

   The table 1 demands are for process heat applications only. Steam mechanical drives or cogeneration normally demand steam at temperatures up to 800°F, as indicated for the Geismar region in Reference A-1. However, a comparison of these recent GEA results with those generated last year for all steam applications indicates that process heat accounts for at least 90% of the total demand at this time.

2. The Future Potential Steam Market is Much More Assured Than for Synfuels or Oil Shale

   Industrial process steam currently uses about 10% of all primary fuels consumed in the U.S. It also is projected to have a steady growth rate to 2020. Even without growth, the replacement of old industrial plants and conversions away from the high-priced fossil fuels represent a very large potential market.

A-1 A. T. McMain and J. D. Stanley, "1170 MWt HTGR-PS/C Plant Application Study Report: Geismar, Louisiana Refinery/Chemical Complex Application," General Atomic Co., GA-A16094, May 1981.

Shale oil and synfuels, on the other hand, have no presently-established energy demand. Recent energy projections by DRI and Pace show slower growth rates and stretchouts for synfuels production than in the more optimistic projections of a few years ago.

Table 2 gives projections of the energy markets available to the HTGR to 2020. The potential steam market is not only much larger than the sum of synfuels and shale oil, but also more definite.

3. <u>There Are Indisations that the Modular Reactor System Can Tolerate Higher Methanator Costs Without Significant Loss of Industrial Steam Market</u>

Application of the relative product steam costs to the industrial steam market as profiled by GEA indicates that the modular system could have a large market even if methanator and reformer costs each exceed 300 to 400 1980 $\$/10^6$ Btu. These results are certainly much too preliminary to ease up on efforts to lower methanator and reformer costs. However, they do provide hope for a very competitive position with the modular system in supplying industrial process steam.

Market assessments performed to date still indicate that the HTGR-R may have a large industrial steam market. It would seem to be important to evaluate in more detail any indications, such as those presented by C. Davis on November 2, that methanators designed to produce 500 to 700°F steam would be cheaper. Market analyses also continue to indicate no particular preference for methanator size (rating) up to about 50 MWt. The size below that should be primarily determined on the basis of minimum cost.

Finally, there is one other point that needs to be clarified. The condition of limiting minimum methanator output to at least 20 percent of peak is not dictated by market requirements. Actually a larger market could probably be captured for the higher steam costs without that condition. It was used in previous market analyses because this was felt to be the minimum practical power level for methanators. If hot standby using a fossil fuel is now deemed to be practical, that minimum condition should no longer be a constraint.

APPENDIX A

TABLE 1

INDUSTRIAL PROCESS STEAM DEMAND BY TEMPERATURE RANGE

(Results Of GEA Analyses)

Temperature (°F)	Fraction Of Steam Demand (Boiler Output)				
	PSE&G	Houston L&P	No. Indiana PS	Pacific G&E	U.S.
<350	0.50	0.51	0.29	0.50	0.49
350-500	0.42	0.40	0.20	0.45	0.41
500-750	0.07	0.08	0.27	0.04	0.08
750-1000	0.01	0.01	0.24	0.01	0.02
>1000	0	0	0	0	0
	1.00	1.00	1.00	1.00	1.00

APPENDIX A

TABLE 2

PROJECTIONS OF TOTAL ENERGY DEMANDS AVAILABLE TO THE HTGR

	Available Energy Demand (Quads/Yr)			
	1995	2000	2010	2020
1. Industrial Process Heat				
Steam (Boiler Output)	9.0	9.8	11.3	13.1
Direct (Process Requirement)	5.4	5.5	5.7	6.0
2. Electricity Produced (Peaking and Mid-Range)	7.8	8.6	11.1	14.0
3. Shale Oil Conversion Energy	0.3	0.5	1.4	2.6
4. Synfuels Conversion Energy				
Gas	0.1	0.2	0.5	1.3
Liquids	0.1	0.2	0.7	1.9
5. Energy Required To Reform Product				
Ammonia	0.5	0.6	0.8	1.1
Methanol	0.1	0.1	0.2	0.4

APPENDIX B

Advanced Methanation System Characteristics

by C. R. Davis

Planning for the design and development of an advanced methanation system, to be accomplished in FY-1982, has been completed. (See Table 1.) Consideration was given to the following data in determining the design requirements for the Methanator Plant:

1. Marketing - Latest marketing data indicates that about 1/3 of the industrial steam users operate 24 hours/day, another 1/3 operates 16 hours/day and the remaining 1/3 operates 8 hours/day. Over 90% of all industrial steam is used at temperatures $\leq$ 500°F.

2. Availability - The highest possible (approaching 100%) availability is desired, hence the use of multiple or redundant methanation trains are required.

3. Advanced Catalyst - Studies indicate that the use of monolithic catalyst can reduce the required catalyst volumes by as much as a factor of 5. This in turn reduces the size and cost of the methanator vessels.

4. Distribution Piping - A study was completed to optimize the pressure drop and heat loss in a distribution pipe, five miles long. Likewise consideration was given to returning condensate to the methanator plant from the steam users. Conditions to/from the distribution pipe are:

METHANATION PLANT	USER
Steam-40,000 lb/hr. 730°F, 900 psia	530°F, 670 psia
Condensate-32,000 lb/hr, 110°F, 40 psia	150°F, 200 psia

Feedwater is condensate plus 8,000 lb/hr makeup at 60°F and 40 psia.

Taking the information of the previous page into consideration, it was decided to size the methanation plant for 60 MWt, using four 15 MWt trains. The reduction in power from 40 MWt (FY-1981 design) to 15 MWt plus the use of monolithic catalyst reduces the size of all the methanators and heat exchangers to less than four foot diameter with lengths less than 14 feet. This will permit the use of modular construction with four to eight components each manufactured at the fabrication plant. All modules will be sized for railway transport. Field labor for piping, electrical and structures will be virtually eliminated.

Figure 1 depicts the operating power levels and time for each methanation train for the anticipated daily load demand. If the HTGR-R plant delivers a constant syngas supply at 40 MWt/hour level, storage for 20 MWt for eight hours must be provided. It is assumed that storage will be a part of the syngas pipeline system and not a part of the methanation plant. One possible alternative would be for the HTGR-R plant to be a "load follower" which would eliminate the need for storage.

APPENDIX B

TABLE B-1

60 MWt METHANATOR PLANT DESIGN REQUIREMENTS

ITEM	VALUE (per train)
No. Trains = 4	
Size (each)	15 MWt (Net)
Syngas In (Lbs/Hr, °F, psia)	29,800,100,900
CH_4 Out (Lbs/Hr, °F, psia)	15,279,TBD,870
Condensate Out (Lbs/Hr, °F, psia)	14,521,TBD,870
Feed Water In (Lbs/Hr, °F, psia)	40,000,100,40
Steam Out (Lbs/HR, °F, psia)	40,000,730,900
Temperatures:	
Maximum Process Gas	950°F
Maximum Vessel Shell	900°F
Pressure Drop:	
Thru Catalyst	0.2 To 1.0 psi/ft.
Gas thru Heat Exchangers	1.0 To 3.0 psi
Water/Steam thru Heat Exchangers	2.0 to 5.0 psi
Monolithic Catalyst:	
Active Surface	150 to 250 M^2/gm
Sulfur Tolerance	< 0.05 ppm
Life	5 yr/low temp 2.5 yr/high temp
Plant Life (Major Components)	40 year
Heat XGR Tubes & Tube Sheets	15 year
Pumps, Valves, Compressors	10 year Avg.
Train Assembly	To be pallatized in sub-assemblies for RR transport
ASME Code Design	Section VIII, Division 1.

60 MWt METHANATION PLANT
DAILY LOAD DEMAND

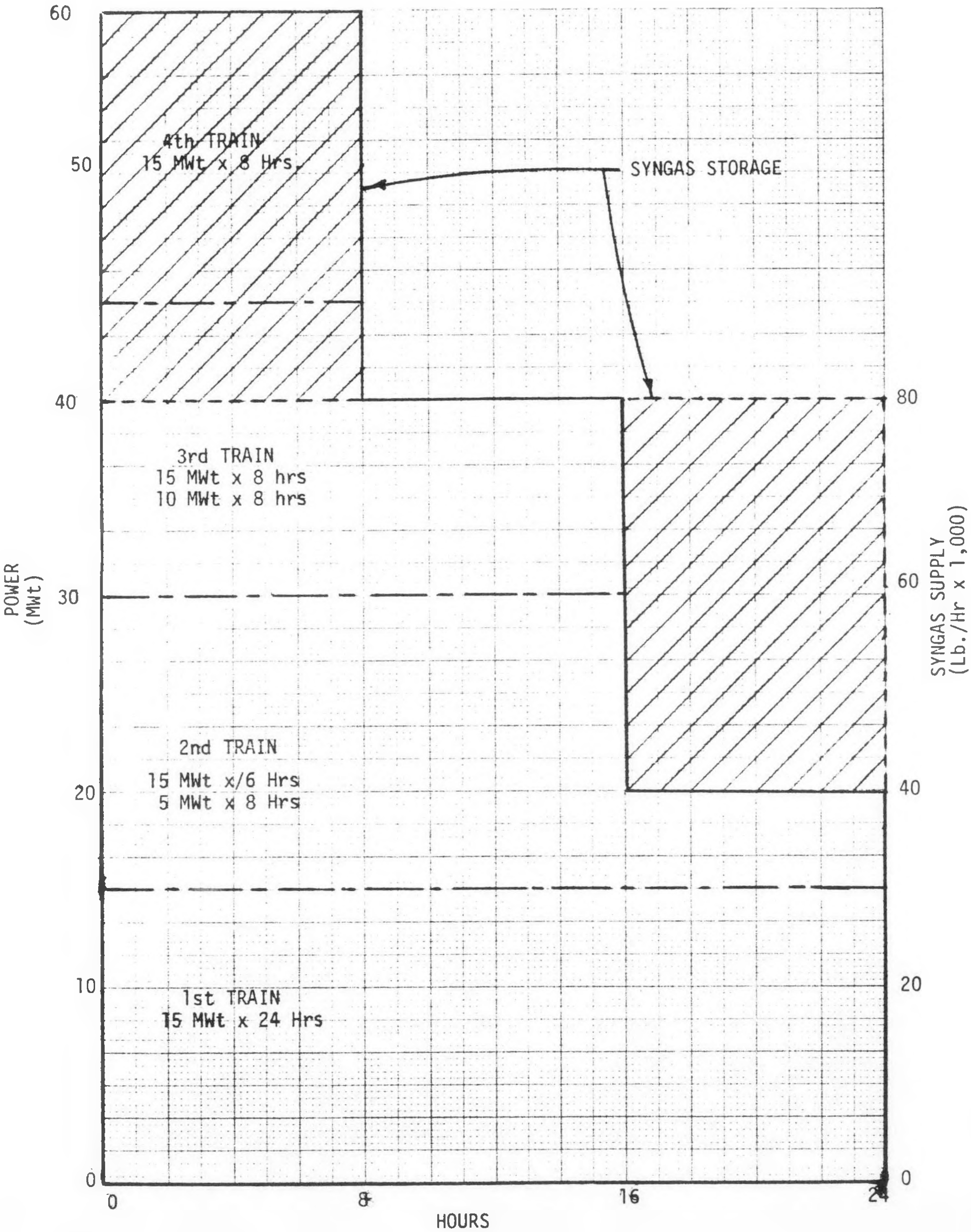


FIGURE B-1

APPENDIX C

Market Requirements for Advanced Methanation System

by B. A. Hutchins

This memo is to document the methanation system market requirements for the advanced design and development. The primary requirement is essentially for the methanation plant to provide a steam output ratio of average to peak of at least 0.67 to best serve a three-shift industrial process steam user market.

This results from several considerations. First, the duty factor of the methanators should be a high as practical because of their large contribution to the total system capital costs. Second, the national steam demand (in lb/hr) is nearly evenly divided between one-, two-, and three- shift users. And third, the two-shift users already have an average to peak demand of about 0.67.

The duty factor portrayed in Figure 1 of the reference is the situation if virtually all the local steam demand were supplied by the methanation plant. This may not be the case, since it may be difficult for the HTGR-TCP system to compete with coal for large users on three shifts. In order for the methanation plant to maintain a duty factor of 0.67 or greater, it is only necessary that it serve as much demand (in lb/hr) on three shifts as on one shift (since the two-shift plants are consistent with a duty factor of 0.67). Thus, the methanation plant could serve one- and three-shift users over a range from zero to that shown in Reference 1. If this range makes a difference in design and analysis efforts, perhaps it would be better to choose an intermediate case: say, where both the one- and three-shift demands are 10 MWt. Figure 1 would then show 60 MWt for the first 8 hours, 50 MWt for the second and 10 MWt for the third. The two-shift demand would be 40 MWt. It isn't known how this would impact reliability.

There are two other market areas that can impact the design. These are:

- The 1981 GEA results do indicate that 90% of the <u>process</u> steam entering industrial plants (boiler output) is at 500°F or less. This is a good target for considering methanators with lower output steam temperatures.

- The peak output of 60 MWt is a little large for the local market densities in the U.S. It would be better to have a peak of 45 MWt and an average of 30 MWt.

APPENDIX D

Methanation for a 60 MWt Steam Plant

by N. T. Arcilla

The design of a methanation process for a 60 MWt steam plant has been completed.

Four identical, semi-independent 15 MWt methanator modules will be run in parallel to give a total output of 60 MWt. The methanation process was designed to meet the following requirements:

1. Supply 9-10 MWt to make steam at 900 psia and 532°F

2. Supply about 2 MWt of superheat to bring the steam at 900 psia to 730°F

3. Attain a methane yield of over 90%

4. Minimize methanation steps and reaction volumes

Several process configurations were studied, using METH code to calculate the catalyst requirements and process conditions. The configuration selected uses a two-step methanation. Figures 1 and 2 show simplified schematics of the process. In this process, the design requirements are achieved by using:

1. A feed split of 31%-69% between the adiabatic and the isothermal steps. This balances the heat generation in the two methanators in the same ratio.

2. A 6.5 to 1.0 product recycle to feed mass ratio in the adiabatic step which sets the maximum temperature in the vessel to 940°F.

3. An internally cooled second methanator with a low 0.88 to 1.0 recycle ratio. In this vessel, 12 MWt is produced by the reaction

and 9.6 MWt goes to steam production. Methane yield is brought to 92% (dry basis).

4. A monolithic (or honeycomb) catalyst matrix which makes possible the greatly reduced reaction volume used in this process.

The methanators process at very high thruputs as a result of using the monolithic (or honeycomb) matrix to support the Ni on alumina catalyst. Figure 3 shows a typical geometry. The two major advantages to the honeycomb matrix are:

1. High catalyst utilization on the order of 5-20 times what is normally attained in particulate beds resulting in a much smaller quantity of catalyst required.

2. Negligible pressure drop through the unobstructed paths along the channels.

To illustrate thse advantages, Figure 4 shows a comparison of the relative sizes and catalyst requirements between a monolith and a particulate bed adiabatic methanator. Used adiabatically, a monolithic bed is significantly better.

In the internally cooled methanator, heat rejection to the coolant occurs at the same time as the reaction. A very serious limitation to the use of the mulitchannel monoliths in this design (i.e., heat transfer is perpendicular to gas flow and channel orientation) is the low average overall thermal conductivity of the honeycomb geometry. Absence of radial dispersion of gas eliminates the contribution of convective radial heat transfer owing to gas flow. Moreover, radial conduction of heat in a ceramic substrate is rather small.

To overcome these limitations, the tubes in the isothermal methanators for the 60 MWt Process Steam Plant are designed such that the monoliths are loaded in the tubes in sections spaced 3.25 inches apart by a blank centerbody. The monolith sections are 3.5-6.0 inches long. The sections are assumed to behave as quasi-adiabatic methanators in calculating the area required for radial heat transfer. The centerbody is sized to constrict the gas flow area to one-half the total flow area of the tube. Radial heat transfer occurs mostly in the bare tube sections. The centerbody is formed to effect good gas mixing. Figure 5 shows a sketch of a typical tube in the isothermal methanator.

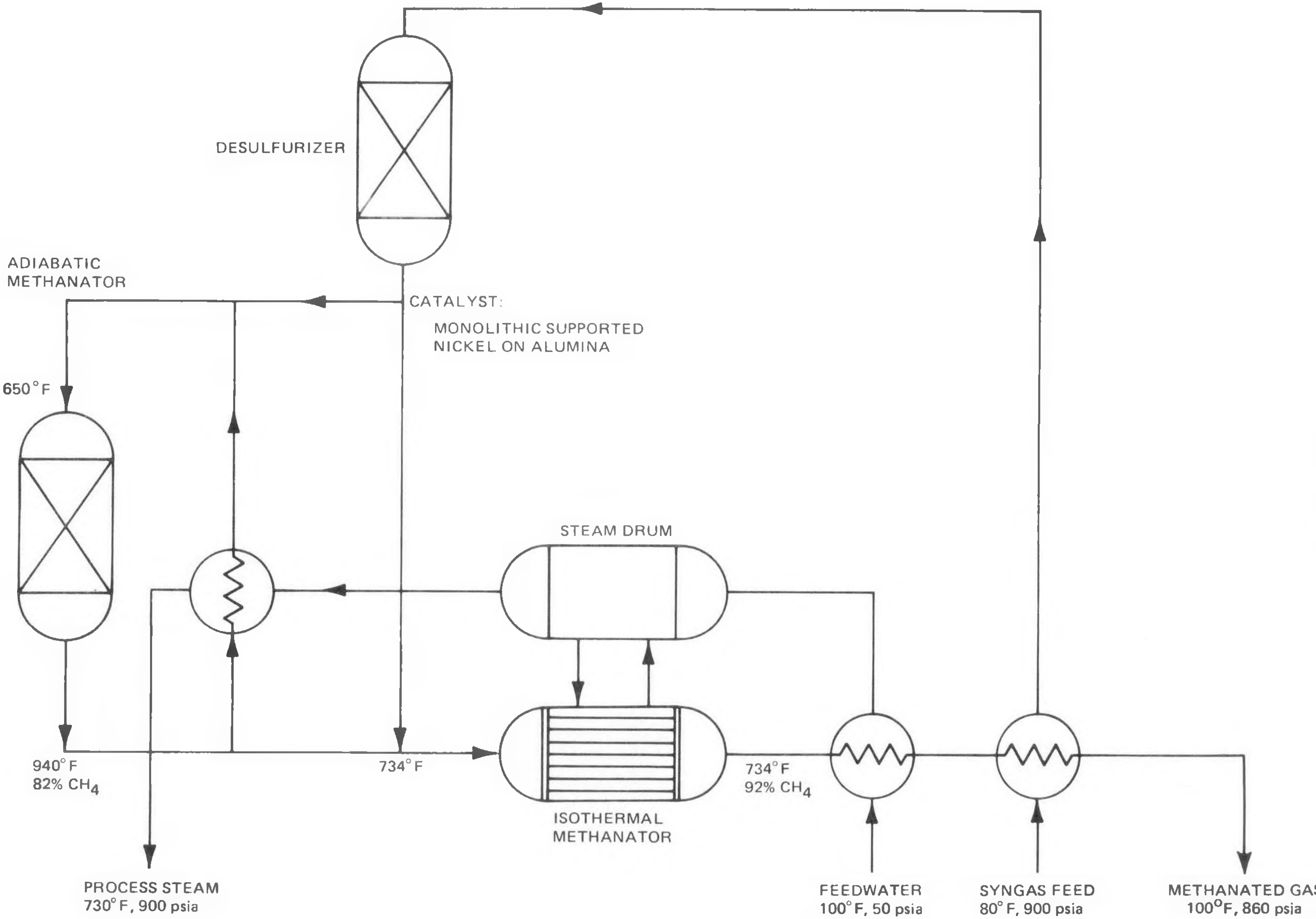
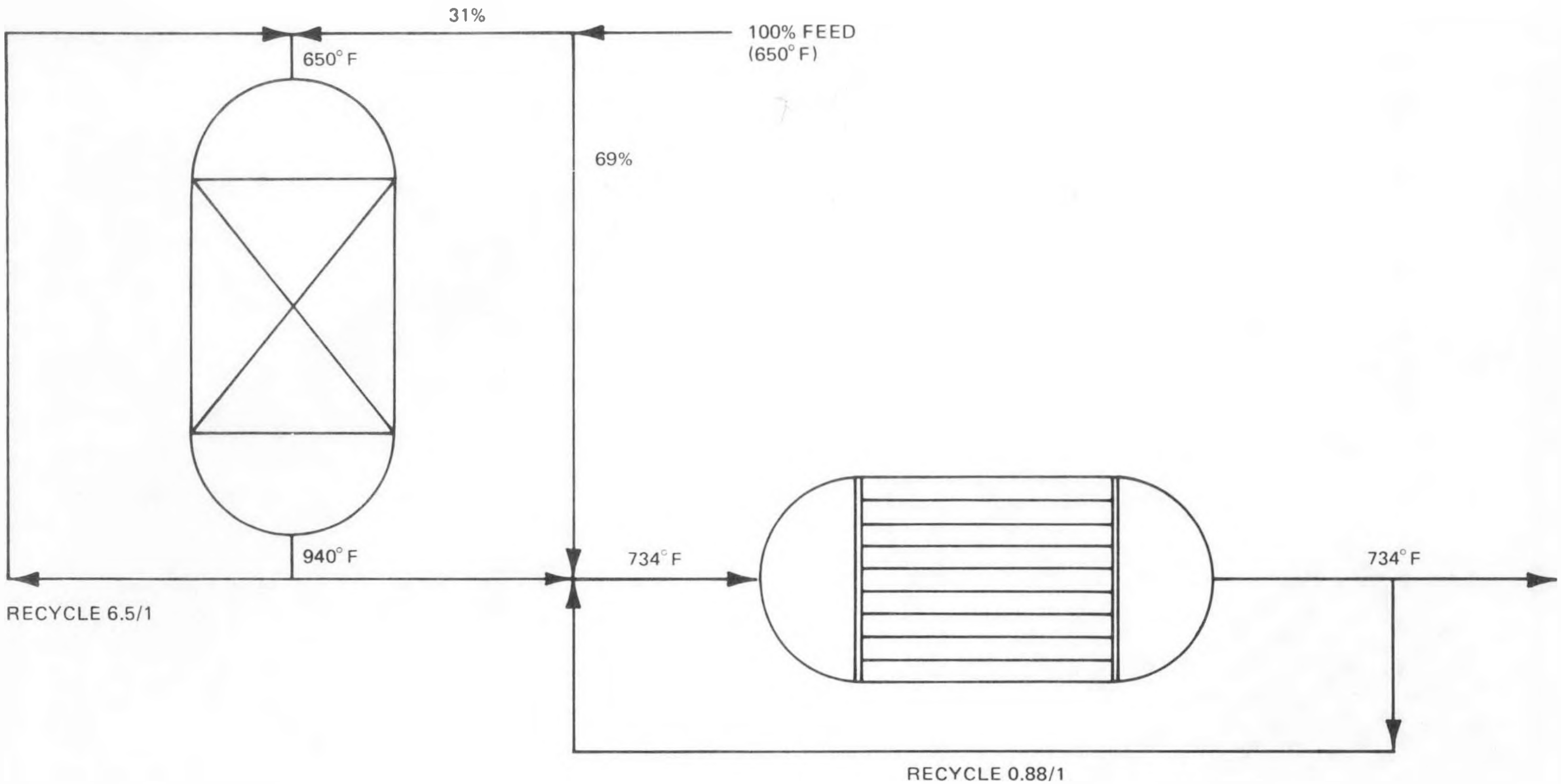


Figure 1 TWO-STEP METHANATION FOR PROCESS STEAM PRODUCTION

82-278-01



	ADIABATIC	ISOTHERMAL
CATALYST VOLUME	42.4 ft^3	20.7 ft^3
CH_4 (DRY BASIS)	81.6%	92.2%
HEAT PRODUCED	5 MW_T	12 MW_T

82-213-09

Figure 2 TWO-STEP METHANATOR 15 MW_T MODULE

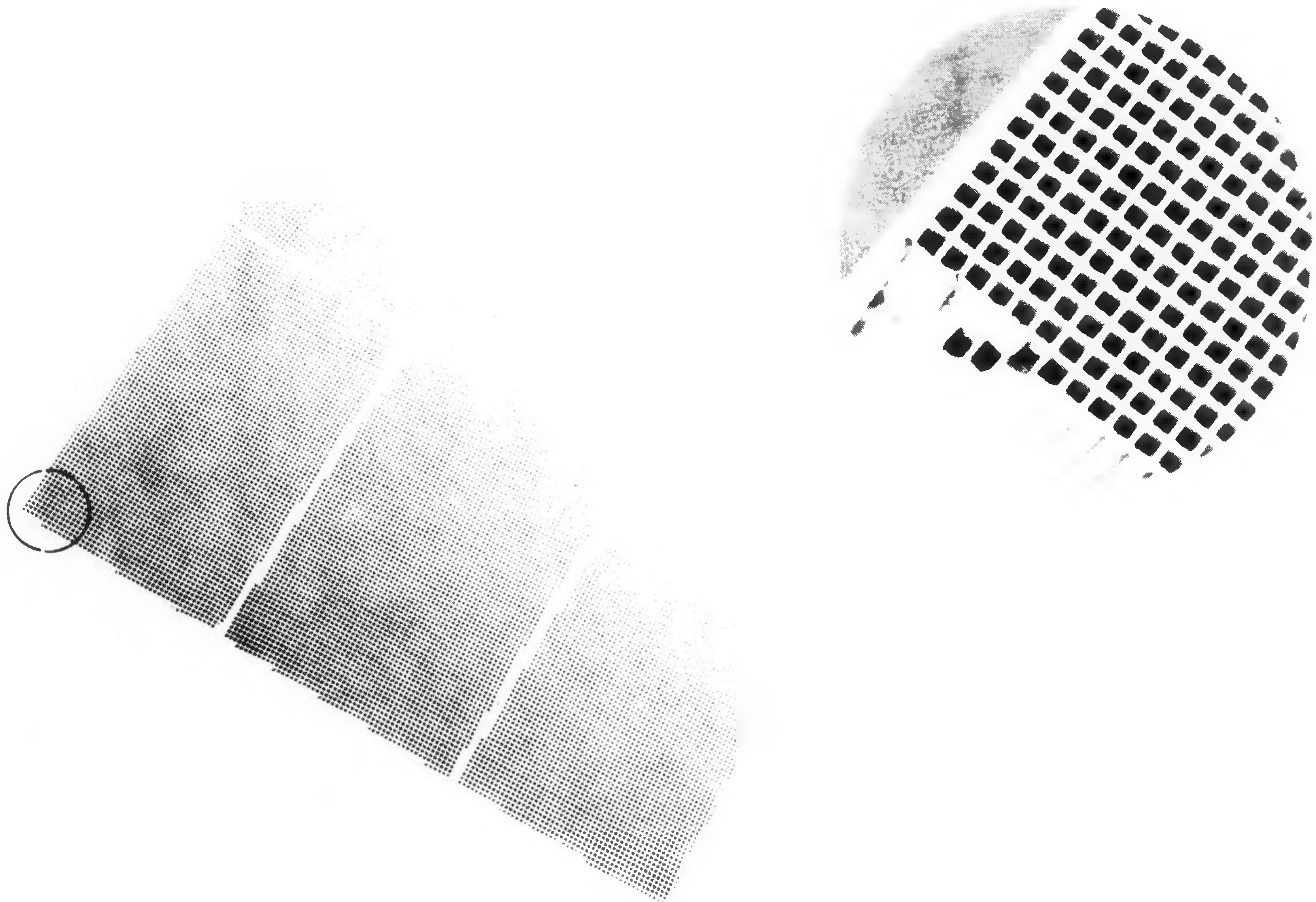


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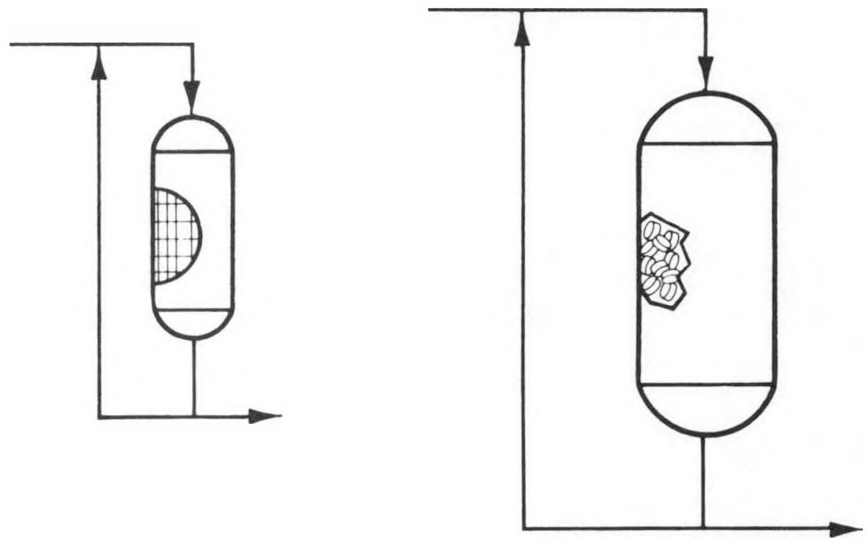
Figure 3 MONOLITHIC STRUCTURE FOR ADVANCED CATALYTIC SYSTEM

RELATIVE VESSEL SIZES FOR ADIABATIC METHANATORS



	MONOLITH BED	PARTICULATE BED
FEED, SCF/HOUR	1.66×10^6	1.66×10^6
GHSV, SCF GAS/CF CATALYST	60,000	15,000
BED VOLUME, CF	28	111
ΔP, psi	2	2
WEIGHT OF CATALYST + SUPPORT, lb	1060	6645
WEIGHT OF NICKEL, lb	33	1594

82-213-06

Figure 4 MONOLITH VS PARTICULATE

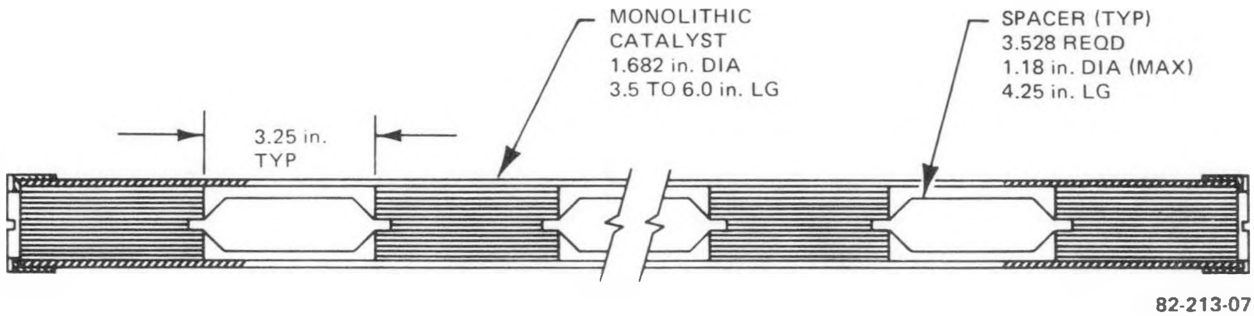


Figure 5 TYPICAL CATALYST TUBE IN THE ISOTHERMAL METHANATOR (60 MW_T PROCESS STEAM PLANT)

APPENDIX E

Preliminary Methanator Plant Design and Heat Balance
by M. M. Hui

The preliminary methanator plant system design for FY82 HTGR study has been performed and the heat balance for 15 MW methanator process train is shown in Figure E-1. The main characteristics of the methanator plant design are summarized as follows:

- o The size of each methanator train is 15 MWt net with four trains in the methanator plant.
- o Process steam is generated at 900 psia - 730°F, a reduction from 900°F of the FY81 design.
- o Maximum syngas/process gas temperature is 941°F.
- o The methanation process was derived from the two methanator system configuration, one adiabatic methanator and one methanator/evaporator unit.
- o Monolithic catalyst is used for the design of the methanation process.
- o The methanation process can be controlled by two process gas recirculation loops.
- o A steam drum is used as a steam/water separator and as a contact feedwater heater. The steam drum will locate on top of the evaporator/methanator unit to allow natural circulation of feedwater to the steam generator.
- o Direct steam feed from the drum to the deaerator to reduce feedwater flow and feedwater pumping.

- The operating conditions of the feedwater train was optimized to reduce heat transfer equipment and system cost.

- The efficiency of the methanator plant is about 98%.

Figure 2 shows the heat transfer process and the log mean temperature differences for the heat exchangers in the feedwater train. The plant heat balance and the process diagram will be used for the sizing and costing of the methanator plant equipment.

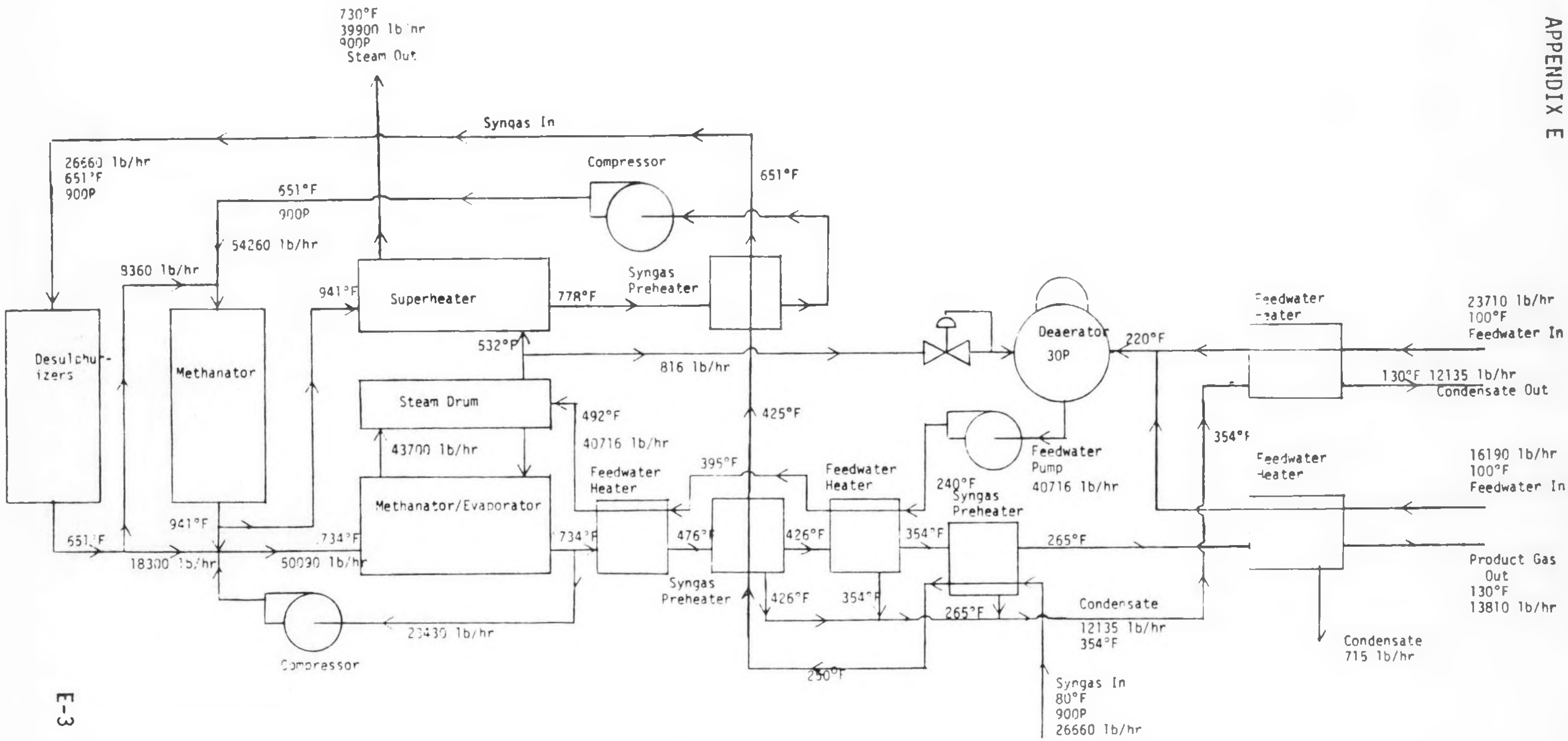


Figure 1 - PRELIMINARY METHANATOR PLANT HEAT BALANCE

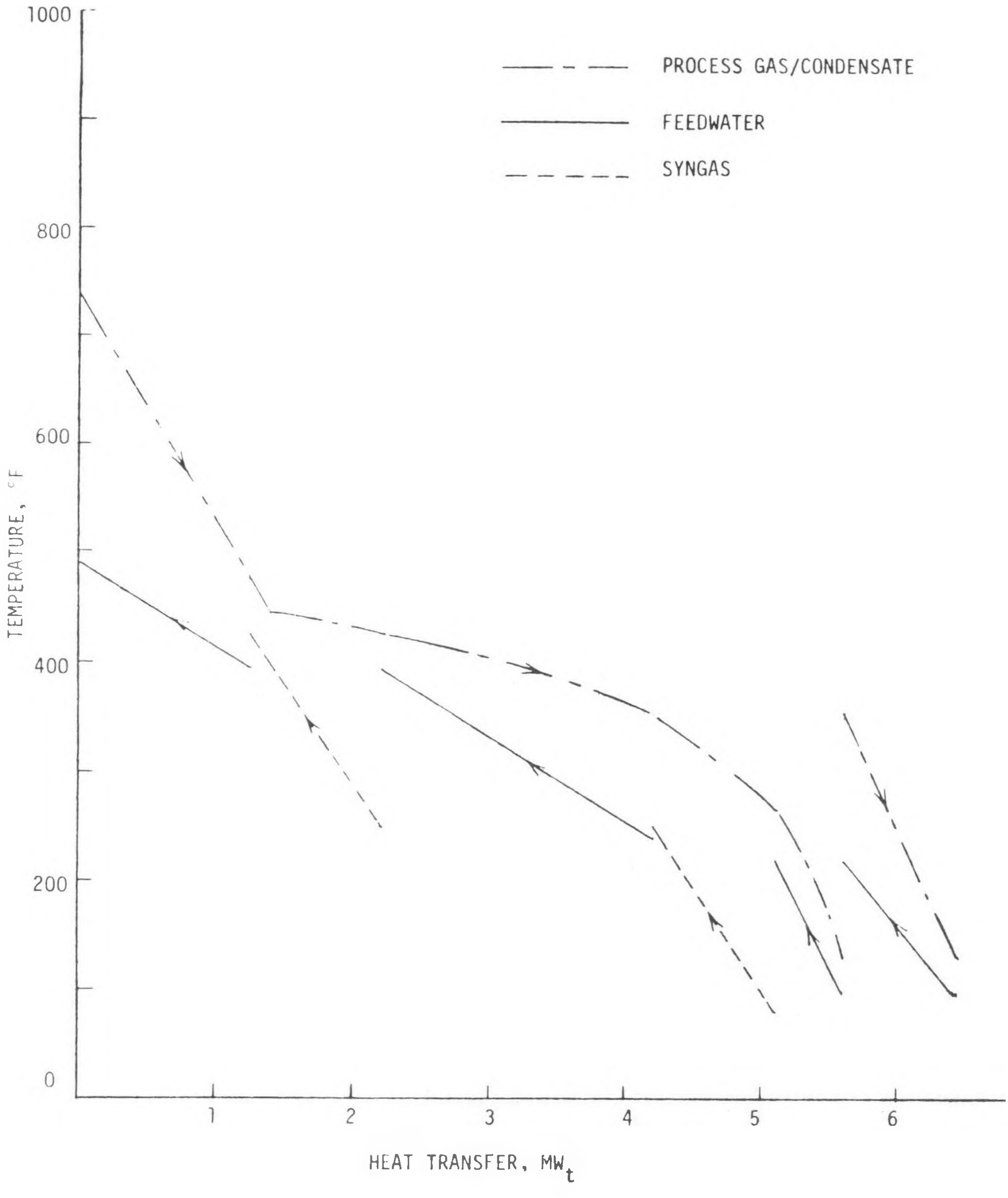


FIGURE 2 HEAT PROCESS FOR FEEDWATER TRAIN

APPENDIX F

Preliminary Design of Methanator Plant Heat Exchangers

by M. M. Hui

The preliminary design of the heat exchangers for the methanator plant has been performed and is presented in this memorandum. The information will permit design and costing of methanator plant.

The design of the methanator heat exchangers is based on the plant heat balance as shown in Figure 1. The heat exchangers and process gas composition are identified, respectively, in numerical number and alphabet. The methanator plant will be palletized so that there are four trains in each plant with 15 MWt nominal rating for each train. The process conditions for the feedwater trains are shown in Figure 2 and Table 1 summarizes the process gas composition in the trains.

The computer code HEATEX was used for sizing the methanator plant heat exchangers. HEATEX was developed by Core Materials as a quick and accurate means for calculating the size and characteristics of shell and tube-type heat exchangers. For feedwater heaters, a square pitch arrangement was adopted for the designs. Table 2 summarizes the major characteristics of the heat exchangers for the methanator plant.

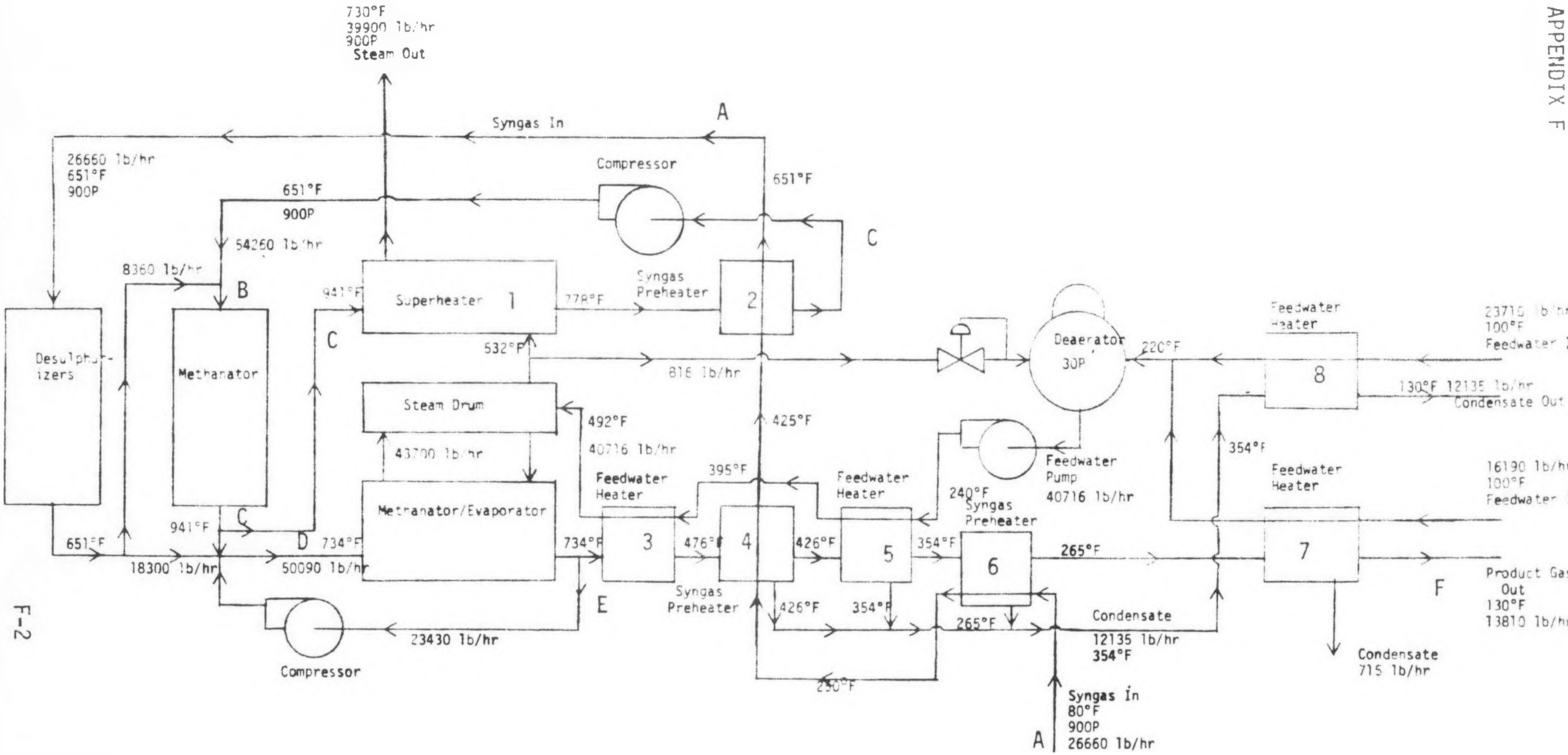


Figure 1 - PRELIMINARY METHANATOR PLANT HEAT BALANCE

APPENDIX F

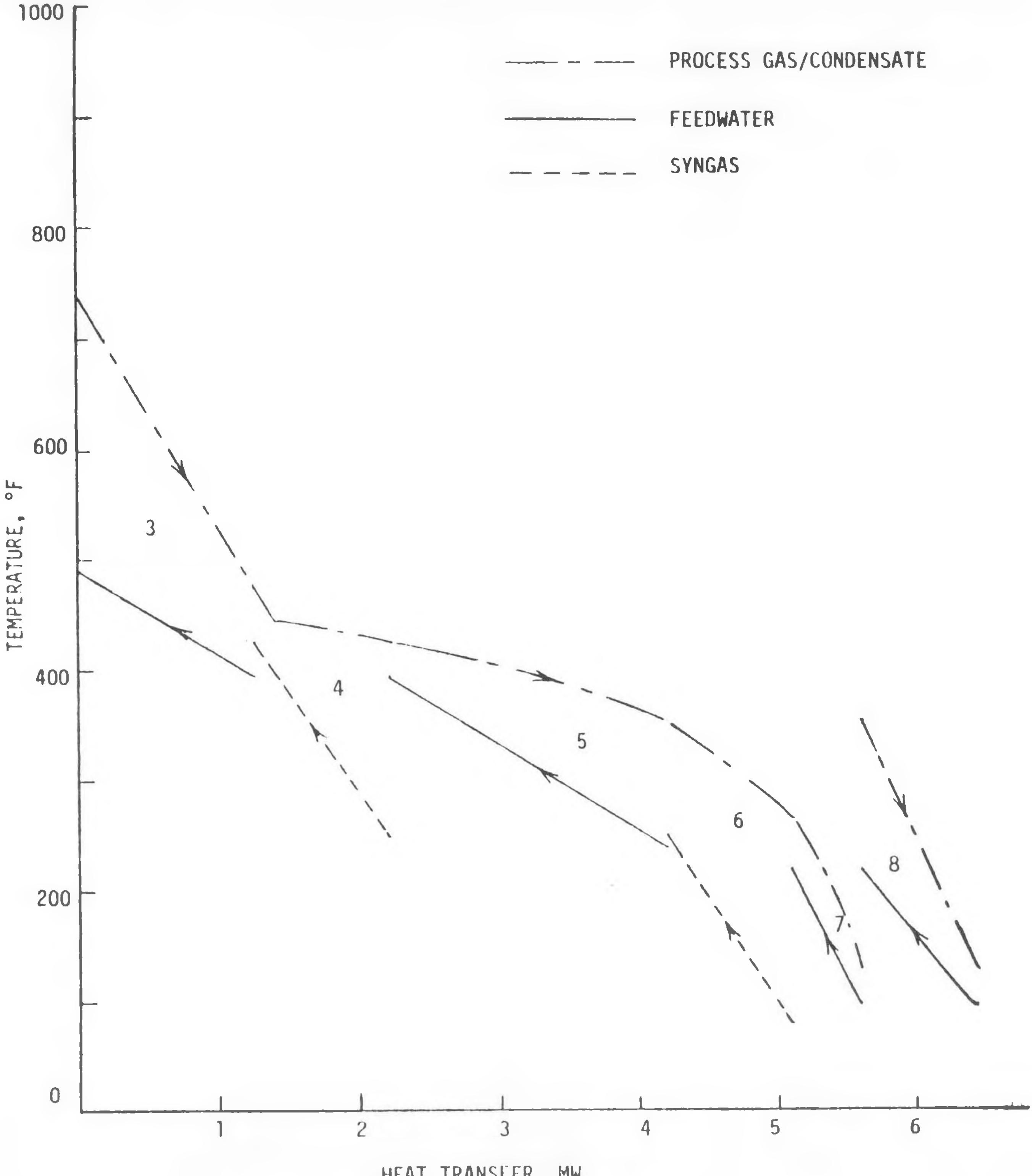


FIGURE 2 HEAT PROCESS FOR FEEDWATER TRAIN

APPENDIX F

TABLE 1 - PROCESS GAS COMPOSITION (MOLE %)

	A	B	C	D	E	F
CO	9.9	2.00	0.09	4.75	0	0
CO_2	9.5	3.53	2.10	5.17	0.86	1.6
H_2	67.7	20.12	8.68	34.91	3.43	6.2
CH_4	12.9	41.41	48.27	32.39	51.01	92.2
H_2O	0	32.94	40.80	22.78	44.70	0

TABLE 2 - METHANATOR PLANT HEAT EXCHANGER CHARACTERISTICS

No.	Type	Power MWt	Shell Side	Tube Side	Shell I.D. (In.)	No. of Tubes	Tube OD/ID (In.)	Pitch (In.)	Length (Ft.)	LMTD (°F)
1	Superheater	1.82	Steam	Gas	35	329	1.25/1.032	1.5625S*	9.6	231
2	Gas Preheater	1.41	Gas	Gas	25	198	1.25/1.032	1.5625T**	8.0	179
3	Feedwater Heater	1.25	Water	Gas	25	154	1.25/1.032	1.5625S	11.0	147
4	Gas Preheater	0.98	Gas	Gas***	25	316	1.00/0.782	1.25T	9.0	101
5	Feedwater Heater	1.91	Water	Gas***	42	2157	0.625/0.407	0.8125S	10.4	64
6	Gas Heater	0.91	Gas	Gas***	25	198	1.25/1.032	1.5625T	10	141
7	Feedwater Heater	0.55	Water	Gas***	45	2511	0.625/0.407	0.8125S	10.2	37
8	Feedwater Heater	0.83	Water	Gas	35	1479	0.625/0.407	0.8125S	9.3	62

* Square Pitch
** Triangle Pitch
*** With Condensate

APPENDIX G

BECHTEL JOB 14508

FOR

GENERAL ELECTRIC COMPANY

ADVANCED REACTOR SYSTEMS DEPARTMENT

METHANATOR PLANTS FOR THE HTGR - REFORMER THERMOCHEMICAL PIPELINE SYSTEM

Conceptual Cost Estimate

May 1982

RM1/5-1

APPENDIX G

CONTENTS

RM1/5-2

TABLES

APPENDIX G
PROJECT DATA

CLIENT:	General Electric Company
LOCATION:	Hypothetical site near Wilmington, Delaware
SCOPE:	Conceptual construction cost estimate for a methanator plant with 60 MWT capacity consisting of 4-15 MWT trains. As alternative case, the construction cost of the 60 MWT units built in multiples of 14, was also evaluated.
PURPOSE OF ESTIMATE:	To provide the client a conceptual construction cost estimate for a 60 MWT methanator plant. These results will be used to assess the economic viability of the project and to identify areas of potential cost savings for further investigation.
PRICING LEVEL:	First Quarter, 1980.
SCHEDULE:	Construction Schedule is not included in this study.

RM1/5-4

APPENDIX G

Introduction

1.1.2 General

In consideration of the economic feasibility of utilizing the heat generated from a high temperature gas reactor (HTGR), this study has been prepared to estimate the construction cost of one of 14 methanator plants of 60 megawatt thermal capacities. The basic concept utilizes a thermochemical pipeline (TCP) to transport energy from the HTGR to multiple process heat users.

A methane reformer facility, located nearby the HTGR nuclear power plant, reforms methane into syngas, mainly CO and H_2, by utilizing the rejected heat from the HTGR. The reformed gas is then transported, via the TCP, to a series of methanator plants where the gas is converted back to methane for return to the reformer. The heat generated from this exothermic methanation reaction is converted to superheated steam to meet the industrial steam demand.

1.2 Technical Scope

The scope of this study is limited to the methanator plants. Two construction costs of methanator plants have been estimated and included in this report. A brief description of these two plants are as follows:

CASE	THERMAL CAPACITY	DESCRIPTION
1	60 MWt	The plant consists of four trains, 15 MWt each train. Equipment, piping and electrical wiring required for each train are installed on modules at fabrication plant and the modules are delivered to construction site for the installation. The plant is designed to

		generate approximately 40,000 lb/hr of superheated steam at 730°F, 900 psia.
2	60 MWt	Plant design and capacity are identical to Case 1. However, the construction cost of one methanator plant has been reported when 14 such plants are constructed simultaneously at locations so that construciton facilities can be shared.

A brief process description of the base case is as follows:

The syngas is delivered to the adiabatic methanators equipped with a monolithic nickel catalyst at 651°F, 892 psia after it passes through the desulfurizer for residual sulfur cleanup. The product gas from the methanator is then sent to the combination methanator/evaporator for the final methanation. The synthesized methane is compressed and transported back to the reformer after the gas is cooled and dried.

A series of feedwater heaters are provided to recover the heat and preheat the water. A final heating is provided in the combination methanator/evaporator by the heat given off from the exothermic methanation reaction. A mixture of steam and water at 532°F and 905 psia generated from the evaporator is sent to the steam drum. The saturated steam from the steam drum is further heated to 730°F at superheater before it is supplied to the industrial process steam demand. The supporting equipment for the facility includes recycle gas compressor, startup heater, startup nitrogen storage and water treatment plant.

APPENDIX G

Construction Cost Estimate

The construction cost of the base case methanator plant (60 MWt) included in this report has been estimated and based on the following:

2.1 Estimate Bases

The estimates are based on the conceptual design and engineering information prepared by GE for the study in the form of engineering drawings, outline specifications and equipment lists. Estimating methods consistent with the conceptual nature of the design information were employed and rely on informal vendor contact, utilization of current Bechtel information as well as equipment cost provided by GE. The primary source of cost data is from the Bechtel Refinery & Chemical Division for process units similar to the methanator plant.

2.2 Pricing Levels

The estimate is at First Quarter, 1980, price and wage levels. No allowance has been made for future escalation.

2.3 Field Costs

The construction cost estimate is composed of field costs, engineering services and contingency. The largest category, field costs, comprises the direct installed cost of permanent plant equipment and the indirect cost of temporary construction facilities, services, equipment and non-manual supervision. The estimate anticipates an engineer/constructor direct-hire operation employing field construction labor forces, i.e., single responsiblity for engineering, procurement and construction (EPC).

2.3.1 Direct Field Cost

The direct field construction costs have been developed and included in the estimate based on the following:

Mechanical Equipment

Mechanical equipment required for the methanation plant is of the following two types:

- o Methanator modules including all the equipment for the methanation and its supporting equipment, interconnecting piping, electrical and instrumentation, cable trays and module steel structure are fabricated and equipment installed on module at off plot and then delivered to the construciton site for installation.

- o All the supporting equipment required for the operation of methanator plant are installed directly at the construction site and not on the modules.

Cost estimate basis for these are:

- o Methanator modules

  a) Methanator equipment

  The cost of major equipment including desulfurizer, methanator, steam drum, heat exchanger and others was provided by GE. This cost represents approximately 35% of all equipment cost.

  b) Supporting equipment

  The costs of this equipment were estimated based on a recent similar study.

c) Bulk materials for module

The costs of these items have been included in the estimate by splitting bulk material requirements of previous estimate based on Bechtel experience.

o Plant Supporting Equipment

The cost of this equipment was obtained through direct vendor contacts for the

- Product gas dryer
- Startup nitrogen storage
- Water treatment plant

The remainder of the equipment costs were estimated based on recent Bechtel pricing information. An allowance of undefined equipment has been included at 10% of the identified equipment cost.

Piping

Piping quantities required for the plant were not available for this estimate. Therefore, piping costs were estimated based on a previous study.

Other Bulk Materials

Other bulk materials include the following:

o Civil/Structural

o Electrical

- Instrumentation

The cost of these items have been included in the estimate as a percentage of major equipment costs based on experience with similar process units as defined in the Bechtel Refinery and Chemical Division Estimating Reference Manual.

Construction Labor

The direct hire construction manhours were estimated based on Refinery and Chemical Division standard unit manhours and a productivity factor of 1.5. A wage rate of $16.00 per hour has been estimated for this study and is based on a craft mix appropriate to the type of construction together with a 5% allowance for casual overtime. Sufficient manual labor to complete the project within the construction schedule is assumed to be available in the project vicinity.

The labor rate used for off-site fabrication plant is similar as above, however, due to more efficient labor practice at the off-site compared to construction site, 30 percent savings in fabrication labor has been allowed.

2.3.2 Indirect Field Cost

The indirect field costs are those items of construction cost that cannot be ascribed to direct portions of the facility and thus are accounted separately. They were estimated by modifying the experience on other similar facilities resulting in an assessment of 85% of direct labor costs to reflect the reduction of indirect field costs for the modular concept.

The items covered by indirect field costs are as follows:

- Temporary Construction Facilities: Temporary buildings, working areas, roads, parking areas, utility systems and general purpose scaffolding

- Miscellaneous Construction Services: General job clean-up, maintenance of construction equipment and tools, material handling and surveying

- Construction Equipment & Supplies: Construction equipment, small tools, consumable supplies and purchased utilities

- Field Office: Field labor of craft supervision, engineering, procurement, scheduling, personnel administration, warehousing, first aid, the costs of operating the field office

- Preliminary Check-out & Acceptance Testing: Testing of materials and equipment to insure that components and systems are operable

- Project insurance

2.4 Engineering Services

The engineering services include engineering costs, other home office costs and fee. Engineering includes preliminary engineering, optimization studies, detail engineering, vendor-drawing review, site investigation and support to vendors. Other home office costs comprise: procurement, estimating and scheduling services, quality assurance, acceptance testing and construction and project management. The sum of these three categories falls into historically consistent percentages in the range of 10%-20% of total field cost depending on the complexity and duration of the project. For this study a figure of minimum 12% of field construction costs has been used to reflect the reduction in engineering for the modular concept.

2.5 Contingency

Included in the estimate is an allowance for the uncertainty that exists within the conceptual design in quantity, pricing and productivity and that is under the control of the engineer/constructor and within the scope of the project as defined. Based on Bechtel historical experience for projects of a similar nature, and the degree of engineering definition for this study, a nominal figure of 25 percent of field construction cost and engineering services has been included in the base case estimate. However, the contingency is excluded on the equipment costs provided by GE based on GE's statement that these costs include appropriate contingency.

2.6 Qualifications

The following are the major qualifications in the estimate:

- The estimate was prepared assuming that the scope of services will be that of a prime contractor responsible to the owner for engineering, procurement and construction.
- Equipment and materials will be procured from U.S. sources, and lead times will be able to support the project schedule without cost penalties.
- Sufficient manual and non-manual personnel to complete the project within the construction schedule is assumed to be available in the project vicinity.
- Existing water sources and water will be adequate for the requirements of this project.
- Construction site is flat and normal effort is required for the site preparation and excavation.

2.7 Exclusions

The following items are excluded from the project scope and therefore, not included in the estimate:

- Methane compressor, drivers and auxiliaries (This compressor is considered as part of the TCP system.)

- Any special construction such as widening and strengthening existing roads

- Demolition and disposal of equipment and material at the end of project life

- Owner's costs such as financing, process royalties, licenses, permits and the like

- Site investigation, environmental reports and land acquisition

- State and local taxes

- Training of plant operators

- Plant startup and operation

- Future escalation

- All facilities beyond the hypothetical site boundary

2.8 Estimate Basis for Other Case

The estimate basis for Case 1 (60 MWt) has been described above. Case 2 for 1 of 14 60 MWt plants was estimated by adjusting Case 1 cost to reflect the cost savings due to the multiple construction concept and these adjustments are as follows:

- Equipment and bulk material cost. Twenty percent and ten percent discount for equipment and bulk material cost respectively have been included in the estimate to reflect larger quantity procurement.

- Direct Labor. A slight increase to cover the expenses for travel and subsistence to attract sufficient manual labor and to allow for drop in productivity resulting from utilization of less skilled labor force.

- Indirect Cost. Reduced from 85 to 70 percent of direct labor cost to reflect the savings resulting from the sharing of field office services, construction equipment, temporary construction facilities and related services.

- Engineering and fee. Duplication of design engineering and reduction of fee to allow for larger overall project. These costs are reduced from 12 to 7 percent of total field cost.

- Contingency. Decreased from 25 to 20 percent to allow for project design and construction developments in some plants that can be utilized in other plants.

2.9 Estimate Tables

2.9.1 Estimate Tables

The above discussion forms the basis of the estimates contained in the following tables:

Table G-1 Construction Cost Estimate for 1 60 MWt Plant (Case 1)

Table G-2 Construction Cost Estimate for 1 of 14 60 MW Plants (Case 2) (and comparison with Case 1)

APPENDIX G

TABLE G-1

CONSTRUCTION COST ESTIMATE FOR
1-60 MWt METHANATOR PLANT CASE 1

$1,000

	DELIVERED MODULE		CONSTR FIELD		TOTAL	
	MAT'L	LABOR	MAT'L	LABOR		
Methanation	$2,928		$		$2,928	
Supporting			1,050		1,050	
Piping	774		86		860	
Civil Structural	310		730		1,040	
Electrical	260		260		520	
Instrumentation	416		104		520	
SUBTOTAL	$4,688		$2,230		$6,918	
Methanation		132			132	
Supporting				50	50	
Piping		350		40	390	
Civil Structural		300		700	1,000	
Electrical		250		250	500	
Instrumentation		260		70	330	
SUBTOTAL		1,292		1,110	2,402	
TOTAL DIRECT COST						$ 9,320
ADD SUPPORTING MAT'L	180					180
LABOR SAVING ON OFF-SITE		< 400>				< 400>
Adjusted Total Direct Cost	4,868	892	2,230	1,110		9,100
Indirect Field Cost						940
Total Field Cost						10,040
Engineering & Fee						1,200
SUBTOTAL						11,240
Contingency						2,800
Additional Handling and Freight for Modular Construction						300
TOTAL CONSTRUCTION COST PRICE & WAGE LEVEL 1Q,80						$14,340

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TABLE G-2

CONSTRUCTION COST COMPARISON
FOR 1 OF 14-60 MWt METHANATOR PLANTS CASE 2

	STUDY CASE / CAPACITY, MWT	1 / 60	2 / *
Equipment		$ 4,160	$ 3,475
Piping		1,250	1,185
Civil Structural		2,040	1,850
Electric		1,020	925
Instrumentation		850	765
Total Direct Cost		9,320	8,200
Additional Supporting Mat'l		180	180
Labor Saving on Off Site		<400>	<400>
Adjusted Total Direct Cost		9,100	7,980
Indirect Field Cost		940	820
Total Field Cost		10,040	8,800
Engineering & Fee		1,200	600
Subtotal		11,240	9,400
Contingency		2,800	1,300
Additional Freight & Handling for Modular Construction		800	800
Total Construction Cost (Price & Wage Level, 1Q,80)		$ 14,340	$ 11,000

* The cost represents that for one 60 MWt plant when 14 such plants are to be built simultaneously at locations so that construction facilities can be shared.

APPENDIX H

Procedures and Operations

Reference 1, GEFR-00568

6.2 Methanator Process Support System

To startup the methanation system from ambient temperature, with new methanation catalyst, supplementary heat is required. A supply of nitrogen gas and hydrogen gas is also required. Supplementary heat and nitrogen are also required, in lesser quantities, for shutdown and hot standby operations. The components comprising the support system are:

a. Gas-fired nitrogen heater for the two desulfurizers.

b. Gas-fired nitrogen and hydrogen heater for the three methanators.

c. Gas-fired feedwater heater.

d. Liquified nitrogen supply tank.

e. Hydrogen supply tanker (only used for catalyst reduction).

In addition to the above equipment, the feedwater pump, the water recycle pump and the recycle gas compressor are used during startup, shutdown and hot standby operations.

Gas to fuel the three heaters is taken from either the syngas supply or methane return pipeline. If these pipelines are uncharged, then locally available natural gas or propane will be used.

6.2.1 Startup Procedure

The procedure for starting the methanation system from ambient temperature, with fresh catalyst, is as follows:

6.2.1.1 Charge the steam drum, the three feedwater heaters, the evaporator and the superheater with condensate or demineralized water using the feedwater and recycle water pumps.

6.2.1.2 Start the gas-fired feedwater heater and circulate water through this heater and all the methanation system components (6.2.1.1 above) using the feedwater and recycle pumps. Heat the water and all components at the rate of 150°F per hour until all are stabilized at 525°F at a pressure of 900 psia. Then reduce or turn-off the gas-fired feedwater heater.

6.2.1.3 Simultaneous with steps 6.2.1.1, above, charge the three methanators, the superheater and syngas heater with nitrogen at a pressure of 900 psia. Do the same for the two desulfurizers.

6.2.1.4 Circulate the nitrogen in the desulfurizers through their gas-fired heater and raise the temperature at a rate of 150°F per hour until the desulfurizers are stabilized at at temperature of 640-660°F. Reduce the heating to maintain this temperature.

6.2.1.5 Circulate the nitrogen in the methanators through their gas-fired heater and raise the temperature at a rate of 150°F per hour until the methanators are stabilized at a temperature of 750°F.

6.2.1.6 At this point supply a mixture of 20-40 v% hydrogen and 80-60 v% nitrogen to the gas-fired heater and heat it to 750°F at 900 psia. Slowly replace the nitrogen in the methanators with this mixture, exhausting the nitrogen to atmosphere via the "flare" stack.

6.2.1.7 Continue to flow the $H_2 + N_2$ mixture through the catalyst until reduction is completed per the following instructions:([1])

a. The catalyst should be reduced at a temperature of 700-750°F (370-400°C), attained by a reasonably uniform heating rate of no greater than 150°F/hr (83°C/hr). The maximum temperature difference between the catalyst temperature and gas temperature should be limited to 150°F (83°C). A means of measuring bed temperatures at three or more locations (inlet, outlet, and at least one in bed) should be provided in order to monitor the temperature during heatup and reduction. The reducing gas space velocity should be 1000-4000 volume/volume/hr. The superficial linear velocity should not be less than 0.2 ft./sec. (6.1 cm/sec.).

b. Most important during the reduction is that the water content of the reducing gas be kept as low as possible. Therefore, the reducing gas is to be exhausted to atmosphere via the "flare" stack until the water content is less than 0.4 mol %. If this value is exceeded, permanent catalyst deactivation will occur.

c. Reduction should be complete after 6-12 hours. This can be checked by measuring the concentration of water in the reduction gas into and exit from the catalyst bed. The catalyst has been reduced when it no longer consumes hydrogen and water is no longer evolved.

6.2.1.8 Replace the reduction gas with nitrogen and recycle it until the temperature of the catalyst in the three methanators is reduced and stable at the design inlet temperatures. (651°F for the C M/E and 536° F for methanators 2 and 3.)

6.1.2.9 Slowly start syngas flow and exhaust the nitrogen and syngas mixture to atmosphere via the "flare" stack.

6.2.1.10 As soon as a temperature rise is noted in the methanator outlets, increase the flow of syngas until the outlet temperatures of the methanators reach design value. At this point close off the "flare" stack and assume normal operations.

NOTE: For startup from ambient temperature without new catalyst, omit the reduction steps, 6.2.1.6, .7 and .8, above.

6.2.2 Normal Shutdown Procedure[1]

If the reactor is to be shutdown due to interrupted operation, it should always be purged and kept pressurized under an atmosphere of nitrogen or hydrogen. If the catalyst is to be removed from the vessel and discarded, it can be unloaded into used catalyst drums and wet down with water. If the catalyst is to be saved for possible reuse, it should first be subjected to controlled oxidation. However, the catalyst should not be oxidized unless absolutely necessary as it is probable that the activity will be decreased during the oxidation procedure. In the event the catalyst must be removed from the vessel, a recommended procedure is as follows:

After the flow of process gas is stopped, start a flow of nitrogen at a space velocity of 100-200 standard volumes per volume of catalyst per hour. The pressure and flow should be such as to not exceed the design linear gas velocity through the catalyst. When the catalyst bed temperature reaches about 350-400°F (117-204°C), add 0.5 mol percent air to the nitrogen. If, after about one hour the temperature rise has not exceeded 150°F (83°C), increase air percentage until a temperature rise of no greater than 150°F (83°C) is achieved. Do not exceed 3 percent air in the carrier gas. Continue the oxidation until an analysis of the gas from the reactor shows that no more than 10 to 20 percent of the oxygen added is being consumed. Increase the air to five percent and continue the oxidation for about four to six hours. The catalyst may then be cooled to near-ambient temperature and removed.

If at any time during the oxidation procedure any catalyst bed temperature should exceed 600°F (316°C), the quantity of air being added should be reduced, or removed as the particular circumstances dictates, until safe temperature levels can be maintained.

6.2.3 Hot Standby Operation

Hot standby operation may be used for many reasons. It is intended, primarily, to be used when the steam user's plants may be inoperative for extended periods. The procedure for hot standby is as follows:

6.2.3.1 Slowly reduce the flow of syngas into the methanation system.

6.2.3.2 Heat nitrogen with the gas-fired heaters to nominal syngas - methane temperatures and start nitrogen flow into the methanation system, replacing the decreasing syngas flow.

6.2.3.3 Exhaust the nitrogen-syngas-methane mixture to atmosphere via the "flare" stack.

6.2.3.4 After the syngas-methane train is purged with nitrogen shutoff the "flare" stack and recycle the nitrogen through the desulfurizers and the methanators and their respective gas-fired heaters. Reduce the temperatures of the nitrogen at a rate of 150°F per hour until the temperatures of the nitrogen stabilize at the following levels:

a. Desulfurizer Outlet - 600°F
b. Combination Methanator/Evaporator Outlet - 525°F
c. Methanator 2 & 3 Outlet - 525°F

6.2.3.5 Recycle water from the steam drum through the gas-fired boiler and the heat exchangers and the combination methanator/evaporator maintaining a water temperature of 525°F, using the feedwater and water recycle pumps.

6.2.3.6 To startup from hot standby, slowly introduce syngas into the methanation system and exhaust the nitrogen to atmosphere via the "flare" stack.

6.2.3.7 As soon as a temperature rise is noted in the methanator outlets, increase the flow of syngas until the outlet temperatures of the methanators reach design value. At this point shut off the "flare" stack and assume manual operation.

6.2.4 Emergency Shutdown

There appears to be only two conditions which would be classified as "emergency" requiring shutdown of the methanation system.

6.2.4.1 Power Failure

Pumps and the recycle compressor are provided with a time delay dropout to ride through power dips of short duration. For longer periods of power failure shut off the syngas feed and introduce hot nitrogen into the syngas - methane steam. Maintain the hot standby mode until power returns.

6.2.4.2 Pipe Break or Major Gas or Water Vessel Leak

If a pipe break or major leak in a vessel occurs, either gas or water, there will be a sharp reduction in pressure. The system is equiped with pressure sensors, which in the case of a sharp pressure drop would activate the shutoff valve of the syngas feed. With syngas off the system will grad-