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METHANATION PLANT DESIGN FOR HTGR PROCESS HEAT

By
C. R. Davis

September 1981

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METHANATION PLANT DESIGN
FOR
HTGR PROCESS HEAT APPLICATIONS

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ABSTRACT

This report describes a 40 MW_t Methanation Plant for generating process heat in the form of steam for industrial application. The plant receives syngas via pipeline from a HTGR-Reforming 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 100°F and a pressure of 900 psia. Methane and condensate are returned to the HTGR-R plant at a temperature of 100°F and a pressure of 870 psia. Steam is delivered to the steam distribution system at a temperature of 900°F and a pressure of 900 psia.

One isothermal catalytic reactor and two adiabatic catalytic reactors are used for the conversion of syngas to 94.1% (dry bases) methane. The isothermal reactor is also the evaporator used to convert saturated water to saturated steam. Three tube and shell type heat exchangers are used to raise the water temperature from 100°F to a saturation temperature of 532°F. A steam superheat is used to obtain the steam final temperature of 900°F. The methanation system includes also; desulfurizers, deaerator, steam drum, air cooler and dryer as well as appropriate pumps and compressors.

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1. INTRODUCTION

1.1 Background

A survey of existing methanation systems, suitable for adaptation to the supply of industrial process heat and steam was conducted by Flock & Vakil in 1978 (1). It seemed from this survey that the adiabatic fixed bed arrangement with interstage cooling of the methane/syngas steam was most suitable and that the RM Process, proprietary to the RM Parsons Co. of Pasadena, Cal., would allow higher pressures and temperatures to be generated than was possible with competing systems.

Under contract to GE-ARSD a 56 MW_t design for the production of process steam was prepared by the RM Parsons Co. in FY-80; a complementary but independent design was also prepared by KTI of Pasadena, Cal. in the same year. The result of these studies was disappointing and confusing; the capital costs of the Parson's design were approximately eleven times greater than Flock & Vakil had estimated while those of KTI were some six times greater than those estimated by Flock & Vakil. This was the nature of the disappointment. The nature of the confusion was that since little independent work had been conducted within ARSD and because a proprietary data rights agreement existed between ARSD & RM Parsons, no satisfactory understanding or conciliation of the two to one cost difference between similar designs by reputable corporations could be achieved.

It was resolved that in FY-81, work would continue within GE-ARSD to develop an understanding of methanation systems and practices such that an accurate and independant methanator cost estimate could be made. This work should also be an endeavor to reduce the methanator costs to a point that they did not cripple the economics of the HTGR-R/TCP system and that

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fundamental assumptions with respect to methanator deployment and employment would be re-examined in the light of continuing market studies.

1.2 Objectives

The objectives of the work reported here were as follows:

(a) To develop a process steam methanation plant design using existing methanator technology or that thought to be obtainable in the near term. This design must be of sufficient detail to permit cost data of greater accuracy to be determined which would resolve previous uncertainties.

(b) To redesign the system to emphasize the energy extraction qualities and characteristics. This recognizes the fact that the paramount function of existing methanation systems is to convert syngas to acceptable quality synthetic natural gas rather than to efficiently extract energy from the process.

(c) To attempt to reduce costs by combination of system functions within equipment, thereby reducing the number of equipment items where possible.

(d) To adapt the industrial process steam methanation plant design to the requirements for process steam and heat established for oil shale reduction.

(e) To commence investigation of the possibilities for cost reduction, performance enhancement, reliability, ease of maintenance and other desirable qualities which might be possible with advanced methanation systems; specifically monolithic catalyst and fluidized bed systems.

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1.3 Assumptions

Substantial uncertainty exists at the present time as to the nature and extent of the United States industrial process steam and process heat requirments. This uncertainty is further added to, when it is considered that methanation systems would be deployed at some future time when the U.S. economy will be expanded compared to the present and a change in the relative importance of industries may exist.

It is of crucial importance that fundamental assumptions for the design of the methanation system arrangements meeting the objectives stated in 1.2 should be such that the utility of the work reported herein maintains stability as superior perceptions of methanator applications emerge. Thus the assumptions listed in the following, leading to a Basis of Design (3.0) are followed, each in turn, by a short discussion on the preservation of the utility of the work.

(a) The industrial process steam methanator design should be for a complete and self contained power plant; that is, capable of operation and maintenance without reliance on services from the load or facility to which it is coupled.

This assumption permits a design which, via steam distributing piping, can serve an industrial area with many small users and which can also function as as adjunct to a specific large scale user.

(b) The thermal rating of the process steam methanator plant should be some 40-50 MW_t.

The distribution of live steam over a five mile radius is a practical maximum from the standpoint of steam quality degradation and distribution piping capital cost. The aggregate process steam requirement of industrial users within a ten mile diameter in the PSE&G utility service area is 40-50 MW_t and this aggregation is thought to be typical of other major industrial service areas.

Also at 40-50 MW_t two or more such plants would serve the needs of single large users such as refineries or synfuel plants with no particular cost penalty. Conversely a plant designed for a 40-50 MW_t rating may be scaled, both technically and from a cost standpoint to 20-25 MW_t should such a requirement arise.

(c) The steam produced by the process steam methanation plant will be distributed at commencing pressure and temperature of 900° psia and 900°F.

Surveys of the small user industrial process steam market suggest that the preponderance of user requirments are for steam at a temperature less than 650°F. Deterioration of steam quality in a distribution pipeline can be limited over five miles such that the most distant user still receives steam suitable for the bulk of his needs. A temperature of 900°F and 900 psia also permits the steam to be used for electricity generation in some applications and avoids any necessity to use expensive high temperature materials in the plant.

(d) The process steam methanation plant will be designed to operate at base load between outages for maintenance or catalyst replacement and be designed for a service life of 40 years. Provisions for part power operation will be made providing such provisions do not compromise the basic design.

Continued assessment of the industrial process steam market has indicated that the process steam requirements of second and third shift users is considerably greater than had been thought so that the plant should be capa-

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ble of operating at about 25 - 35% of full power for 8 hours; at about 60-75% for 8 hours and at 100% of full power for 8 hours.

(e) The process heat methanation system for various synfuel plants will be designed as a stand-alone adjunct to the oil shale or coal reduction plants and would supply hot methane to a series of heat exchangers provided by the synfuel plant designs. It is apparent that economics in plant design would be realized should the process heat methanators be intergrated within the synfuel plant, but at this stage of design evaluation is not apparent which of the numerous synfuel processes is most adaptable to the supply of nuclear heat.

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SECTION 1 - REFERENCES

(1) Vakil, H. B. and Flock, J. W., "Closed Loop Chemical Systems for Energy Storage and Transmission (Chemical Heat Pipe)", Appendix A1 Final Report No. COO-2676-1, Prepared for V. I. Department of Energy dated February, 1978.

2.0 SUMMARY

The design of a reference methanation system to supply process steam to industrial users has been completed. The design effort required GE-ARSD to become knowledgeable of catalytic reactor kineties and to intergrate this technology with the technology of mass and energy balance, heat transfer and the mechanical design of heat exchangers and pressure vessels.

2.1 Conclusions

The completed 40 MW_t Methanation Plant design satisfies all the design requirements and most all of the methanator objectives for FY 1981. Accomplishments which can be demonstrated and evaluated include:

(a) Completion and validation of the METH computer code. The code was written by GE, reviewed and upgraded by Professor Calvin Batholomew of Brigham Young University and validated by comparison to experimental data obtained from United Catalyst, Inc. manufacturer of methanation catalyst. The METH code is a valuable tool for determining the exothermic heat generated, the packed-bed size and pressure drop through the catalyst bed for adiabatic methanators. The code can also be used to size the catalyst volume required for an isothermal reactor.

(b) Completion of the HEATEX computer code which sizes tube and shell type heat exchangers. The code determines the tube diameters, the square footage of heat transfer area required and pressure drops.

(c) Completion of the physical design of all major components in the Methanation System.

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(d) Completion of the identification and sizing of all components in the Methanator Support System.

(e) *Completion of the Methanator Plant Arrangement Design.

(f) *Completion of the cost estimate for the fabrication of all equipment and the construction of the methanator plant.

2.1.1 Scaling the Reference Methanation System

One-half of the objective (reference system must be scalable up to 200 MW_t and down to 10 MW_t) was accomplished. The system was scaled up to 180 MW_t using three isothermal reactor and four adiabatic reactors. The 180 MW_t designed included the design of all major components in the methanation system and Bechtel was able to compute a cost estimate by scaling. A scaled down version was not accomplished, but there are no apparent reasons why a smaller methanation system would be infeasible.

2.1.2 Cost Estimates of the Reference Methanation System

Cost estimates indicate the 40 MW_t Methanator Plant would be $16 x 10^6 (FY 1980 $) for 14 plants to process the 560 MW_t of syngas generated by one 850°C IDC HTGR-Reforming plant. Although this cost is higher than anticipated or desired it is the most realistic cost estimate to date. Cost of one system designed in FY1980 was estimated to be over $25 x 10^6, but a detailed break-down of costs are not available. The cost of the other system designed in FY1980 was $12 x 10^6, but this cost estimate was incomplete. It did not include the Support System required for startup, shutdown and hot standby operation.

*These activities were accomplished by Bechtel National, Inc., by subcontract to GE-ARSD.

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2.2 Recommendations

There are several ways of reducing the costs of the proposed methanation system and at the same time increase reliability and availability of the methanator plant. All the following recommendations are proposed to accomplish these objectives.

2.2.1 Advanced Methanation Catalyst

Sufficient research and development needed to produce an advanced catalyst such as a monolithic or fluidized-bed should be accomplished. The most promising appears to be a nickel alloy catalyst on a cermic or metal monolith (monolith is a honeycomb type structure). The use of monoliths greatly increases the surface area per unit of volume which in turn reduces the volume required and hence the size of the vessels. The volume of monolithic catalyst required may be only 1/2 or 1/3 the volume required for tablets.

2.2.2 Re-evaluation of L/D Ratio of Catalyst Beds

The reference methanator design uses a L/D ratio of 1.1 to 1.3. If this ratio can be increased to 2.0 or 3.0 for the same volume of catalyst the diameters of the methanator vessels can be reduced and likewise the wall thickness of the vessels. If all vessel outer diameter can be reduced to a maximum of 60 inches they can be centrifugal castings, which eliminates longitudinal welding and improves reliability.

2.2.3 Scale Down the Reference Methanator Size

Availability of 100% can be achieved by having a spare methanator train. A 40 MW_t methanator plant should have three 20 MW_t methanator trains or five 10 MW_t trains and one support system. This would halve the capacity of all methanators, desulfurizers and heat exchangers. The reduced size components could be assembled at the factory, including the piping, mounted and shipped on pallets as 1 to 3 subassemblies. The labor required for installation at the site would be reduced by a factor of 4 to 5. Smaller methanator trains would also permit greater flexability in sizing methanator plants i.e., they could be any size desired at intervals of 10 or 20 MW_t.

3. BASIS OF DESIGN

3.1 Application

The referenced design 40 MW_t Methanation Plant will provide steam to industrial and commercial users at pressures and temperatures up to 700 psia and 700°F. This allows for a 200 psia and 200°F loss in pressure and temperature in the distribution lines which may be as long as 5 miles. Application of steam at lesser values may be obtained by use of metering values and expansion tanks at the users facilities. It is estimated that 25-35% of the users will operate three shifts (24 hours per day), another 25 to 35% of the users will operate one shift (8 hours per day) and the remaining 50 to 70% of the users will operate two shifts (16 hours per day). The ability to adjust steam generation to meet load demands is included in the Methanation Plant design. If, for any reason the load demand goes to zero, the plant is operated in a hot stand-by mode by use of gas-fired gas heaters and boiler.

3.2 Plant Inputs and Outputs

The 40 MW_t Methanation Plant receives syngas from a HTGR-Reformer Plant and make-up water from a local (on-site) supply.

3.2.1 Syngas to Methane Conversion

The Methanation Plant receives 1,202 lbm/min of dry syngas via pipeline from the HTGR-R Plant. The syngas is received at 900 psia and 100°F. Its composition is 67% w CO and CO_2, 13% w H_2, and 20% w CH_4. Passage of the syngas through the first methanator converts 88% w of the CO, CO_2 and H_2 to CH_4 and H_2O. Continued passage through the second and third methanators results in conversion of 96-97% w of the CO, CO_2 and H_2 to CH_4 and H_2O. On

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a dry bases the gas is now 95% w CH_4, 4% w CO_2 and 1% w H_2 which is returned to the HTGR-R Plant via pipeline at 870 psia and 100°F. The H_2O obtained by conversion is also returned to the HTGR-R Plant via a separate parallel pipeline.

3.2.2 Steam Generation

Water for steam generation is mostly condensate returned by the steam users. Make-up water may be from any local source. Appropriate water treatment will be used prior to pumping the water into the methanation plant system. It is assumed that water enters the deaerator at 100°F from which it is pumped through feed water heaters and then to a steam drum. The steam drum holds a mixture of water and steam at 905 psia and 532°F. Saturated steam flows from the steam drum through the superheater and then to the steam distribution pipeline(s) at 900 psia and 900°F.

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4. DESIGN CRITERIA

In order to accomplish the objectives set forth in 1.2 above, a methanator design task force was established in the first quarter of FY-81. The objective of the task force was to consider and to integrate the attributes of three technologies:

(a) Kinetics of catalytic reactors
(b) Mass and energy balance and heat transfer
(c) Structural and mechanical design of pressure vessels and heat exchangers

The final results of the task force's work was the development of the Methanation Plant Design Requirements and the selection of methanator types.

4.1 Design Requirements

The design requirements for the reference $40MW_t$ Methanation Plant are presented in Table 4.1. In addition to the methanation system design requirements it was decided that the first methanator would be combined with the water evaporator in order to eliminate one heat exchanger and to limit the temperature rise in the first methanator. Thus the combination methanator/evaporator is an isothermal process which offers the following advantages over an adiabatic methanator ([1]):

(a) Product recycle may be reduced and one-pass operation at high CO concentrations is practical.

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TABLE 4.1

Reference Methanation Plant Design Requirements

ITEM	VALUE
PLANT SIZE:	40 MW_t GROSS
SYNGAS IN	100°F; 900 psia
FEEDWATER IN	100°F; 920 psia
STEAM OUT	900°F; 900 psia
CH_4 OUT	100°F; 870 psia
1st METHANATOR:	
MINIMUM RXN ZONE TEMP	650°F
MAXIMUM RXN ZONE TEMP	1,400°F (1)
2nd & 3rd METHANATORS:	
MINIMUM RXN ZONE TEMP	500°F
MAXIMUM RXN ZONE TEMP	N/A
MAXIMUM VESSEL SHELL TEMP	900°F
MAXIMUM PROCESS GAS PRESSURE	900 psia
MAXIMUM PROCESS GAS TEMP	1100 °F (2)
PRESSURE DROP THRU CATALYST	0.5 TO 1.5 psi/ft
CATALYST:	
ACTIVE SURFACE	$50 m^2/gm$
SULFUR TOLERANCE	<0.1 ppm
LIFE	3 YEAR/HIGH TEMP 5 YEAR/LOW TEMP
PLANT LIFE (MAJOR COMPONENTS)	40 YEAR
LIFE XGR TUBES & TUBESHEETS	15 YEAR
INTERNAL INSULATION THICKNESS	"0" DESIRED
COMPONENT SIZE	RR TRANSPORTABLE
ASME CODE DESIGN	SECTION VIII, DIVISION 1

(1) Approximate equilibrium for 650°F Syngas Inlet Temperature.
(2) Obtained by Recycle in 1st Methanator.

(b) Heat exchange duties and cooling costs are lowered.

(c) Avoiding low inlet temperatures and chemical-reaction-limited kinetic regimes reduces catalyst requirements.

(d) Avoiding extremely high reactor exit temperatures and hot spots prevents sintering of the catalyst, thereby preserving catalyst life.

4.1.1 180 MW_t Process Steam Methanation Plant.

The design requirements and the combination of the 40 MW_t referenced plant methanator/evaporator design was used successfully to size a 180 MW_t. The Process Steam Methanation Plant for coal gasification is described in Section 6.7, below.

4.1.2 80 MW_t High Temperature Process Heat Plant.

The design requirements and the combination methanator/evaporator design of the reference 40 MW_t reference plant could not be used in the design of the 80 MW_t Process Heat Methanation Plants for oil shale reduction because the temperatures of the syngas to methane conversion must be greater than 1500°F. Hence, the design of the 80 MW_t High Temperature Process Heat Plants for both PARAHO and TOSCO II oil shale reduction processes employ only adiabatic methanators as described in Section 6.8, below.

4.2 Plant Availability

An individual methanator train as depicted in the 40 MW_t design is expected to be on-stream continuously, except for scheduled shutdown and minor unscheduled shutdown occurrences for an availability factor greater than 95%.

A scheduled shutdown of 14 days/year to coincide with refueling the HTGR is anticipated. Unscheduled shutdowns are not expected to be more than 2 days/year.

4.3 Service Life

It is anticipated that pumps, valves, and compressors will require repair and/or replacement on time intervals of 3 to 10 years (the same as for other industrial applications).

Heat exchangers will probably require maintenance (tube-plugging) at a rate equivalent to a commercial power plant. All heat exchangers are designed for tube bundle removal and replacement, which is estimated at 15 to 20 year intervals. This includes the tubes and tube sheets of the combination methanator and evaporator.

The desulfurizers, adiabatic methanators, steam drum and knock-out pots have a design life of forty years for the vessels. The catalyst in the desulfurizers needs to be replaced about every 4 years while the catalyst life in the Combination Methanator/Evaporator is estimated at 2.5 years and the catalyst life in the adiabatic methanators is estimated to be 5 years.

Overall the design life of the 40 MW_t Methanation System is 40 years with tube bundles (tubes and tube sheets) being replaced 1 or 2 times during the life of the plant.

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4.4 Design Safety Requirements (1)

The 40 MW_t Methanation Plant is designed in accordance with modern practice for refinery, chemical plant and steam generating plants. Since there is no radiation, neither shielding nor radiation access control is required. Applicable standards and codes used to design the 40 MW_t Methanation Plant are:

Component	Standard	Code
Foundations and Other Concrete Work	ASTM	
Structural Steel	ASTM	
Vessels	ASTM	ASME
Heaters and Furnaces	ASTM	ASME
Shell and Tube Exchangers	ASTM	ASME, TEMA-R
Pipe, Valves and Fittings	ASTM	ASA, ASME
Pumps	API-610	
Gas Compressors	API-617, 618	
Electric Power Distribution and Lighting		NEC
Steam Generating Equipment	ASTM	ASME
Plant Arrangement	OISG*	

*OISG - Oil Insurance Safety Guidelines.

4.5 Inspection (1)

The annual inspection consists of many items of which some major items are given below:

(a) Non-destructive inspection of methanator vessels, heat exchangers, piping, valves, etc. for metallurgical stability - cracks, loss in wall thickness, corrosion, erosion, etc. Techniques include X-ray, ultrasonic, visual.
(b) Boiler inspection per code.
(c) Inspection of fired heaters.
(d) Overhaul of pumps.
(e) Overhaul of compressors.

4.6 Quality Control

The laboratory equipment essential to successful operation of the methanation system are process gas chromatograph and sulfur analyzer. Analyses are needed to monitor and record the feed gas composition, desulfurized effluent and all the methanator effluent compositions.

Adequate sample connections are provided for obtaining appropriate samples from the process for analysis as desired.

Most of the samples at the take-off points from the process lines are hot, high pressure and contain water vapor which would condense upon cooling at process pressure. To facilitate material balance considerations, complete analysis is desired of all stream components including water vapor. The major design considerations are two fold - reduce the pressure to analytical levels and maintain the temperature above the dew point.

A pressure reducing control valve is used to reduce and maintain the downstream pressure at 15-20 psig. The high pressure piping, preferable 1/2 inch size, is as short as possible to minimize sample-line lag time. A filter is placed next to the sample connection block valve to prevent particulate matter from collecting in the control valve and downstream equipment. The pressure, in turn, sets the minimum temperature to be maintained on each side of the regulating valve, to insure that no condensation will occur.

A relief valve downstream of each pressure reducing control valve will be set at 45 psig, or as required to protect downstream equipment. The sample lines are converged into a solenoid network which permits switching the desired stream to the appropriate analyzer.

Two of the samples are selected for continuous analysis by gas chromatograph on a regular cycle. These are the streams into and out of the methanation unit, namely; feed gas to the unit and methanated product gas returning to the pipeline. Analysis of the other methanation samples are performed at appropriate intervals to identify the plant material balance.

The two chromatograph analyzers would be identical in analytical capability and may be operated simultaneously and continuously by one master data processor. Each analyzer in turn contains a microprocessor which has full control over the instrument in which it resides. The analyzer microprocessor performs all analysis function and also handles all alarm and safety functions of the analyzer system. Should a failure of any safety feature occur, the microprocessor will shutdown power to the analyzer and transmit an alarm to the master processor.

Use of two chromatograph analyzers permits continuity for analytically monitoring the process. In the event of malfunction of either analyzer, the remaining instrument can handle analysis of the plant stream in a cycle order most appropriate of that time.

When all components are measured as presently contemplated, the preferred calculation method is to use relative response factors with internal normalization. No calibration standard is required. Instead, relative response factors (RRFs) are input for all components into analyzer microprocess memory. Area of each component is corrected by the RRF to provide a corrected area. The corrected areas are then totalized, and each corrected area is divided by the total to provide the individual component concentration. Standard gas is needed only for trouble-shooting.

In contemporary analyzers, a number of special calculation software packages are available. Among these are: element balance, average molecular weight, specific gravity, Btu, and ratio of components, such as steam/dry gas.

The master data processor operates both chromatograph analyzer and can transmits analyzer data output to bargraph recorders, trend recorders line printer, cassette deck or a writer as preferred. The program can be entered or changed as needed.

Monitoring of hydrogen sulfide and organic sulfur in the feed gas streams is important to protect the methanation catalyst against sulfur poisoning. Samples to be analyzed are the following: feed gas to the unit; effluent from 1st Desulfurizer; effluent from 2nd Desulfurizer prior to methanation.

Several types of analyzers are available. The Houston-Atlas analyzer uses a photo cell to read the darkening of lead acetate-treated paper by reaction with H_2S. For total sulfur analysis, the sample is passed through a heated catalytic cell in the presence of hydrogen. The sulfur compounds present are converted to hydrogen sulfide which is detectable on the lead acetate paper. The analyzer may be checked by comparator samples. An optional gas sample preparation kit is available for calibration purposes.

Houston-Atlas recommends a 10 to 15 minute cycle on each mode of analysis. Only one range of detection is used for each analyzer although a 10 fold increase is possible. One recorder is typically used with each analyzer.

A second type of analyzer is the chemiluminescence analyzer (Beckman). H_2S reacts directly with ozone to give measurable radiation in the ultraviolet region. Carbonyl sulfide and other organic sulfur compounds luminesce in the same region, and with greater intensity than H_2S emits. The H_2S $-O_3$ reaction is slow relative to others. Advantage is taken of this difference in reaction rate. In this manner, the organic sulfur is determined first followed by measurement of the H_2S emission.

4.7 Codes and Standards (1)

Standard refinery practices and codes are to be adhered to in the design of the methanation system. The following table gives the codes and standards for the major equipment.

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TABLE 4.2

Codes and Standards for Major Components

Major Equipment	Code	Std.
Methanators & Desulfurizer	ASME Boiler & Pressure Vessel Code - Section VIII, Div. 1.	
K.O., Flash & Steam Drums	ASME Boiler & Pressure Vessel Code - Section VIII, Div. 1.	
Shell & Tube Heat Exchanger	ASME Boiler & Pressure Vessel Code - Section VIII, Div. 1.	TEMA-R
Air Cooler	ASME Boiler & Pressure Vessel Code - Section VIII, Div. 1.	API-660
Fire Heater		API-530
Compressors:		
Centrifugal		API-617
Reciprocating		API-618
Pumps	Hydraulic Institute	API-610
Piping		ANSI B31.1

SECTION 4 - REFERENCES

(1) "Design of Methanation Plants for Peaking Powr Generation and Process Plant" Prepared by: The Ralph M. Parsons Company, For: G.E.-ARSD Project No. P2773, dated: June, 1980.

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5. GENERAL INFORMATION

The 40 MW_t methanation plant receives feed gas from a pipeline transporting reformed gas from the High Temperature Gas-Cooled Reactor-Reforming (HTGR-R) system. The feed gas contains 5 ppmv sulfur, 50% tert-butyl mercaptan, and 50% dimethylsulfide. Prior to methanation the total sulfur content of the reformed gas must be reduced to less than 0.1 ppmv to prevent sulfur poisoning of the methanation catalyst.

The desulfurized feed gas then flows through a fixed-bed combination methanator/evaporator (isothermal process) followed by two fixed-bed adiabatic catalytic reactors. Between reactors, exothermic heat of reaction is removed from the system by the generation of high pressure steam and by feed-effluent exchanged in conventional heat exchangers. As the flow progresses through the reactors and heat exchangers the bulk of the syngas is methanated. The temperature of the process gas is progressively lowered, finally resulting in a reduced temperature favorable for achieving a high conversion of hydrogen and carbon oxides to methane.

The methane gas from this plant is returned to the HTGR-R plant via a pipeline for the steam-reforming operation and thus to close the long distance energy transport system. The condensate generated by converting syngas to methane may be returned to the HTGR-R plant via a pipeline or it may be used as feedwater feed for the methanation plant.

5.1 Desulfurization

To remove sulfur from the reformed-gas (syngas) feed requires that the syngas be heated to between 600-700°F and then passed through a sulfur removing catalyst.

5.1.1 Desulfurization Catalyst

Desulfurization catalyst used in the 40 MW_t methanation plant is manufactured in the form of 1/8" - 3/16" spheres or tablets. It is packed in beds in two pressure vessels. Each vessel contains 39,480 lbs. of catalyst, of which 1/6 (6,580 lbs) is Cobalt-Molybdenum (Co-Mo) and the remainder (32,900 lbs) is Zinc-Oxide (ZnO). The Co-Mo catalyst is loaded on top of the ZnO bed.

There is normally little to no temperature change across the desulfurization catalyst bed during operation and no special treatment is required to activate the reagent. The capacity of a ZnO bed, depending upon the temperature, ranges from 7 to 20 percent of its weight as sulfur. During use, the ZnO is converted to zinc sulfide and therefore its absorption function is progressively lost.

Spent zinc oxide is pyrophoric and may be hydrolyzed to produce hydrogen sulfide. Since it is pyrophoric, it must be cooled to ambient temperature before exposure to the atmosphere. The spent catalyst may then be removed and put into containers for sale through metal recovery channels.

Feed gas from the pipeline is available at 100°F and 900 psia. The gas is preheated to 651°F by methanator effluent gas and then flows to the desulfurizers where sulfur reacts with the zinc oxide reagent. Normal flow is through two vessels in series. When the reagent in the first vessel is spent, it is replaced with a fresh charge and this vessel is brought back on stream as the second in series.

5.1.2 Desulfurization Vessels

Two desulfurizer vessels are designed for operating temperatures of 651°F and a pressure of 900 psia. Each vessel will be provided with inlet

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baffle plates and a slotted-and-screened cap on the gas outlet nozzle. The vessels are 5'-0" ID x 13'-0" high with 3" thick walls. The catalyst is supported by a perforated steel grate. A dump manway is provided at the bottom level of the catalyst bed. A differential pressure instrument is provided to indicate the pressure drop through the bed.

5.2 Methanation

In the methanation of synthesis gas mixtures, these chemical equations represent the net reactions taking place:

a) Water-Gas Shift

$$CO + H_2O = CO_2 + H_2;\ -\Delta H_{25°C} = 9.85\ \text{kcal/mole} \qquad (1)$$

b) Carbon Monoxide Methanation

$$CO + 3H_2 = CH_4 + H_2O;\ -\Delta H_{25°C} = 49.3\ \text{kcal/mole} \qquad (2)$$

c) Carbon Dioxide Methanation

$$CO_2 + 4H_2 = CH_4 + 2H_2O;\ -\Delta H_{25°C} = 39.4\ \text{kcal/mole} \qquad (3)$$

Water-gas shift (Eq. 1) takes place rapidly and is essentially in thermodynamic equilibrium. In synthesis gas mixtures, the carbon monoxide reaction (Eq. 2) dominates. Only when carbon monoxide depletes to ppm levels does carbon dioxide methanation (Eq. 3) takes place. The reactions are catalyzed by various metals. Nickel on alumina supports is the most commonly used, commercially.

The rates of conversion and heat generation in a catalytic methanation process are determined by one) the rate of reaction at the catalyst sites

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and two) the rates of mass and heat transport in the gas and in the catalyst pores. To accurately predict the process temperature changes, bulk conversions and catalyst volume required, the methanation model used in designing the GE-ARSD process takes into account the reaction paths and factors in the contributions of the following reaction parameters:

a) catalyst activity in terms of active metal sites per gram of catalyst,

b) mass transfer rates of reacting gases through the gas and catalyst pores,

c) temperature differences between gas and catalyst particles,

d) reaction rates at the metal sites based on kinetic expressions determined experimentally and summarized in Table 5.1.

5.2.1 Methanation Catalysts

At the core of a methanation process is the methanation catalyst that promotes the exothermic methanation and water-gas shift reactions. The reactions are catalyzed by a number of transition metals, the more active ones being Ni, Ru, Pd, Co, Fe, Mo, Pt and Rh. The metal is dispersed on a porous support such as alumina, silica, or magnesia to increase the amount of metal sites available for catalysis and to reduce metal agglomeration during reaction. At present, essentially all methanation catalysts employ Ni as the active metal, with alumina being the preferred support. Catalyst for the methanation in the GE-ARSD process are listed in Table 5.2. The catalyst is manufactured as nickel oxide on alumina. It is activated prior to use by reduction of the nickel oxide to nickel in a reducing gas mixture of from 20 to 100 percent hydrogen.

TABLE 5.1

Kinetic Rate Expressions for Methanation Reactions on Nickel

I. Carbon monoxide methanation as determined by Sughure and Bartholomew (1)

$$-R_{CO} = \frac{k_1 k_2 P_{H_2}}{k_1 (1 + K_{H_2}^{1/2} P_{H_2}^{1/2} + K_{CO} P_{CO})^2 + k_2 (1 + K_{CO} P_{CO})^2}$$

II. Carbon dioxide methanation as determined by Weatherbee and Bartholomew (2)

$$-R_{CO_2} = \frac{k_1 P_{H_2}^{1/2} P_{CO}^{1/2}}{(1 + k_2 P_{H_2}^{1/2} + k_3 P_{CO_2}^{1/2} + k_4 P_{CO})^{5/2}}$$

Constants = f(T)

III. Water-gas shift reaction as determined by Grenoble (3)

$$- R_{CO} = A \exp(-E/RT) P_{CO}^{-0.14} P_{H_2O}^{0.62}$$

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TABLE 5.2

Catalyst Specifications for the GE-ARSD Process

	Methanator 1	Methanator 2 and 3
Form	Tablet	Tablet
Size, Inch	0.25 x 0.25	0.25 x 0.25
Temperature Range, °F	600 - 1200	450 - 850
Pressure Range, psig	up to 1500	up to 1500
Composition (Dry), Wt %		
NiO	32	75
Al_2O_3	62	22
Others	6	3
Bulk Density, Pf^3	60	58
Surface Area (BET), m^2/g	110	225
Pore Volume, cc/g	0.25 - 0.45	0.45 - 0.55
Crush Strength, psi	> 700	> 500

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5.2.2 Combination Methanator/Evaporator

The combination methanator/evaporator (C M/E) consists of catalyst filled tubes with water flow between the tubes and the pressure vessel wall. This isothermal process was chosen for the following reasons:

a. Combining the first methanator with the evaporator eliminates the evaporator as a separate heat exchanger and hence reduces cost.

b. The use of water as a coolant helps control the temperature of the process gas within the desired limits.

c. By limiting the temperature of the C M/E to 1015°F maximum, the tubes, tubesheets, vessel and vessel ends can be fabricated of ASME approved materials.

d. The need for internal refractory insulation as normally used for high temperatures adiabatic methanators is eliminated. This is also a cost savings.

The C M/E consists of 336 stainless steel tubes 4.00 in. O.D. - 3.62 in. I.D. x 270 in. long. The botton tube sheet is secured to the pressure vessel while the top tube sheet is free to float with temperature changes. The catalyst in the tubes are supported by a perforated metal cone. It is estimated that the life of the catalyst will be 2-1/2 - 3 years. The vessel is 8'-0" ID x 34'-8" high with a wall thickness of 4 in. The body of the vessel is designed for 550°F, the bottom end for 650°F, the top end for 1015°F, and all parts for a pressure of 900 psia.

5.2.3 Adiabatic Methanator

Two adiabatic methanators, located in series, are used for final conversion of syngas to methane and for polishing the methane prior to cooling it for return to the HTGR-Reforming Plant. The temperatures of these methanators are from 550°F to 650°F so that internal refractory insulation is not required. The adiabatic methanator vessels are 8'-0" I.D. x 18'-1" high with a wall thickness of 4 in. The vessels are designed for a pressure of 900 psia.

Both adiabatic methanators have packed-beds of catalyst supported on a monolithic ceramic grate. Manways are located in the top and in the side, near the bottom of the catalyst bed, for loading and unloading the catalyst. It is anticipated that the life of the catalyst in these two methanators will be 5 years.

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SECTION 5 - REFERENCES

1. Sughrue, E. L. and Bartholomew, C. H., "Kinetics of CO Methanation on Nickel Monolithic Catalysts," I&EC Fundamentals, June 1981.

2. Weatherbee, G. E. and Bartholomew, C. H., "Hydrogenation of CO_2 on Group VIII Metals, I. Specific Activity of $Ni/Si/O_2$," Journal of Catalysis, 68, 67 (1981).

3. Grenoble, D. C., Estadt, M. M. and Ollis, D. F., "The Chemistry and Catalysis of the Water Gas Shift Reaction, 1. The Kinetics Over Supported Metal Catalysts," Journal of Catalysis, 67, 90 (1981).

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6. REFERENCE PLANT DESIGN DESCRIPTION

The reference methanation plant is designed as a single methanation system with a capacity of 40 MW_t. Syngas is received from the HTGR-R plant via pipeline at the rate of 72,144 lbm/hr at a temperature of 100°F and pressure of 900 psia. The syngas is converted to 50 w% methane and 48 w% water condensate, which is returned to the HTGR-R plant by separate pipelines at a temperature of 100°F and pressure of 870 psia.

Condensate is received at the methanation plant from local steam users and mixed with make-up water at the rate of 100,000 lbm/hr at a temperature of 100°F and pressure of about 20 psia. Superheated steam is delivered to the users at the same flow rate at a temperature of 900°F and pressure of 900 psia. (This is exclusive of temperature and pressure losses in the steam distribution piping.)

6.1 Methanation Process System

The methanation process system converts syngas (H_2 + CO_2 + CO) to methane and water (CH_4 + H_2O) by passing the gas through a nickel catalyst which also produces exothermic heat. Part of the exothermic heat is used to preheat the syngas for desulfurization and to start the catalytic reaction. The remaining heat is used to convert 100°F water to 900°F superheated steam.

6.1.1 Normal Operation

During normal operation the methanation system is self-sustaining except for electrical power used to drive pumps and a recycle gas compressor. (Even these components could be driven by steam turbines fed by steam

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generated by the methanation process, but electrical drives would still be needed for startup operations.)

6.1.1.1 Syngas to Methane Conversion

Syngas at 100°F and 900 psia flows from the pipeline to the syngas preheated where its temperature is raised to 651°F. From the syngas preheater it flows through the desulfurizers where the sulfur content is reduced to less than 0.1 ppmv. From the desulfurizers the syngas flows to the entrance of the first methanator at which point it is mixed with recycled gas which is 47 w% methane. The mixture flows through the first methanator which is a combination methanator/evaporator (C M/E). The C M/E is a tube and shell heat exchanger with the catalyst inside the tubes. As exothermic heat is generated inside the tubes a portion of the heat is transferred to water which flows between the tubes and pressure vessel shell. The syngas and methane mixture leaves the C M/E at 1015°F and about 880 psia. The process gas (syngas and methane) then flows through the superheater where heat is transferred to saturated steam to raise its temperature to 900°F. Next the process gas flows through the syngas preheated and then through the third of three water heaters. Between the syngas preheater and the third water heater a portion of the process gas is recycled to the entrance of the C M/E. Upon leaving the third water heater the process gas flows through the second methanator, which is an adiabatic packed-bed in a pressure vessel. Upon leaving the second methanator the process gas temperature is 637°F and the methane content is 48 w%. The process gas then flows through the second waterheater; next through the third methanator (another adiabatic packed-bed in a pressure vessel) and then through the first water heater. At this point the process gas is 50 w% CH_4 and 48 w% H_2O. The gas is then routed through a cooler, knock-out drum and dryer to separate the water from the methane. The dry methane (95 w% or 37,448 lbm/hr) is then piped back to the HTGR-R plant. Likewise, the condensate (34, 696 lbm/hr) is returned to the reforming plant.

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6.1.1.2 Water to Steam Conversion

Feedwater for the methanation plant is primarily condensate from the local steam users, mixed with make-up water. It is assumed that the feedwater flows into the methanation plant deaerator where it is mixed with saturated steam at 228°F and 20 psia and that entrained gases are vented to atmosphere. The flow of feedwater is 100,000 lbm/hr and the flow of steam is 50,000 to 70,000 lbm/hr. The feedwater pump pumps the water from the deaerator through feedwater heaters 1, 2 and 3 at temperatures of 228°F increasing to 424°F and pressures of 920 psia decreasing to 911 psia. After leaving the third water heater, 100,000 lbm/hr is routed to the steam drum and the remainder is returned to the deaerator. The feedwater is mixed with saturated water and steam in the steam drum to obtain a stable temperature of 532°F at 905 psia. A recycle pump pumps saturated water at the rate of 150,000 to 200,000 lbm/hr through the C M/E and back to the steam drum. Saturated steam flows from the steam drum through the superheater and then to the steam distribution system. Upon leaving the superheater the steam is 900°F at 900 psia.

6.1.1.3 Tables and Diagrams

The above normal operation of the methanation process system is detailed in the following:

a. Table 6.1, "Mass and Volumes - Syngas to Methane."
b. Table 6.2, "Mass and Volume - Water and Steam".
c. Figure 6.1, "Flow Sheet, 40 MW Plant, 3 Methanators".
d. Figure 6.2, "Piping Diagram, 40 MW Plant, 3 Methanators".

TABLE 6.1

Mass and Volumes
—Syngas to Methane

REFERENCE METHANATION SYSTEM
40 MW_t FLOW SHEET & PIPING DIAGRAM

Element \ Line No.	1	2	3	4	5	6	7	R_1	R_2	8	9	10	11	12	13	14	15
Total Flow: (Lbm/Min)	1,202.4	1,202.4	3.282.5	3,282.5	3.282.5	3,282.5	1,202.4	2,080.1	2,080.1	1,202.4	1,202.4	1,202.4	1,202.4	624.14	624.14	578.25	578.25
CO (Lbm/Min)	320.87	320.87	330.29	11.50	11.50	11.50	4.21	7.28	7.28	4.21	2.05	2.85	-	-	-	-	-
CO_2 (Lbm/Min)	483.86	483.86	637.97	243.89	243.89	243.89	89.33	154.54	154.54	89.33	54.45	54.45	22.40	23.23	23.23	-	-
H_2 (Lbm/Min)	158.30	158.30	188.15	47.28	47.28	47.28	17.32	29.96	29.96	17.32	10.06	10.06	6.02	5.99	5.99	-	-
CH_4 (Lbm/Min)	239.37	239.37	1,213.32	1,540.04	1,540.04	1,540.04	564.13	975.92	975.92	564.13	578.43	578.43	595.72	594.92	594.92	-	-
H_2O (Lbm/Min)	-	-	912.77	1,439.80	1,439.80	1,439.80	527.41	912.40	912.40	527.41	557.41	557.41	578.26	-	-	578.26	578.26
Flow (Ft^3/Min-STP)	41,556	41,555	88,225	73,697	73,697	73,697	26,996	46,701	46,701	26,996	26,287	26,287	26,109	14,977	14,977	9.26	9.26
Pressure (psia)	900	899	900	880	879	878	878	878	900	877	876	875	874	873	870	873	870
Temperature (^oF)	100	651	651	1015	852	651	651	651	656	536	637	536	615	280	100	280	100
Flow (Ft^3/Min-P.^oF)	827.4	1,642	3,485	3,947	3,515	2,980	1,092	1,888	1,842	980	1,052	956	1,026	394	308	10	9.33
Pipe Size (Dia.-Sch.)	4-40	5-40	8-40	10-80	8-80	8-80	540	6-40	6-40	4-40	4-40	4-40	4-40	3-40	2.5-40	2-40	2-40
Velocity (FPM)	9,360	11,810	10,040	7,610	11,080	9,390	7.860	9,410	9,180	11,090	11,900	10,810	11,610	7,680	9,260	430	400

TABLE 6.2

Mass and Volumes
— Water and Steam

REFERENCE METHANATION SYSTEM
40 MW_t FLOW SHEET & PIPING DIAGRAM

Element \ Line ◇	A	B	C	D	E	F	G	H	I	J	K	L	M
Flow (Lbm/Hr)	178,000	178,000	178,000	178,000	106,800	178,000	100,000	78,000	78,000	200,000	200,000	100,000	100,000
Condition*	f	f	f	f	f	f	f	f	fg	f	fg	fg	g
Density (Lbm/Ft^3)	61.99	59.32	55.19	54.11	50.13	52.54	52.54	52.54	44.59	47.17	49.03	1.989	1.176
Flow (Ft^3/Min)	47.86	50.01	53.75	54.83	35.51	56.46	31.72	24.74	29.15	70.67	67.99	837.94	1,417.2
Pressure (psia)	40	920	915	913	911	911	911	911	20	910	905	905	900
Temperature (^{0}F)	100	228	360	389	475	424	424	424	228	532	532	532	900
Pipe Size (Dia-Sch)	5-40	4-40	4-40	4-40	3.5-40	5-40	3.5-40	3-40	3/4"-40	5-80	1-1/4-80	5-80	5-80
Velocity (fpm)	344	566	608	620	517	406	462	482	7,875	560	7,630	6,630	11,220

*f = fluid
fg = saturated steam/water
g = superheated steam

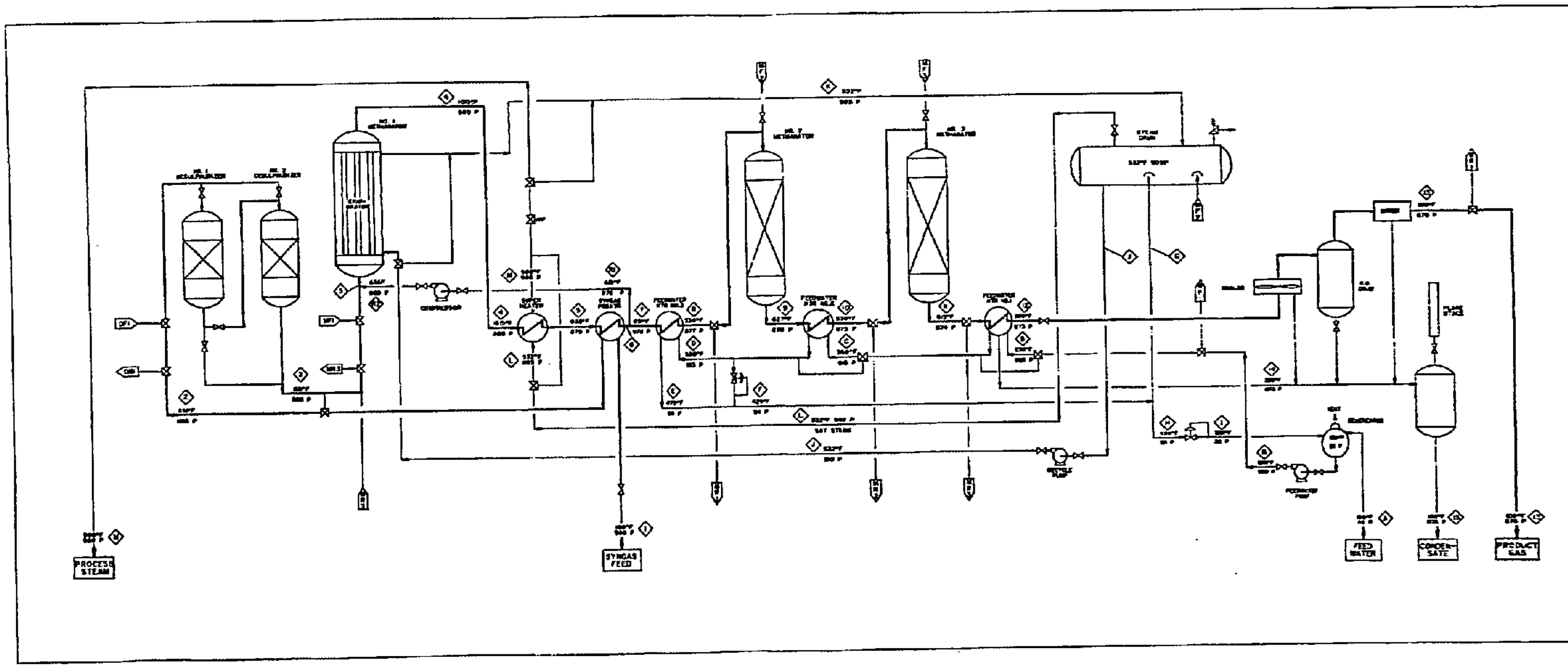

FIGURE 6.1. Flow Sheet, 40 MW Plant, 3 Methanators

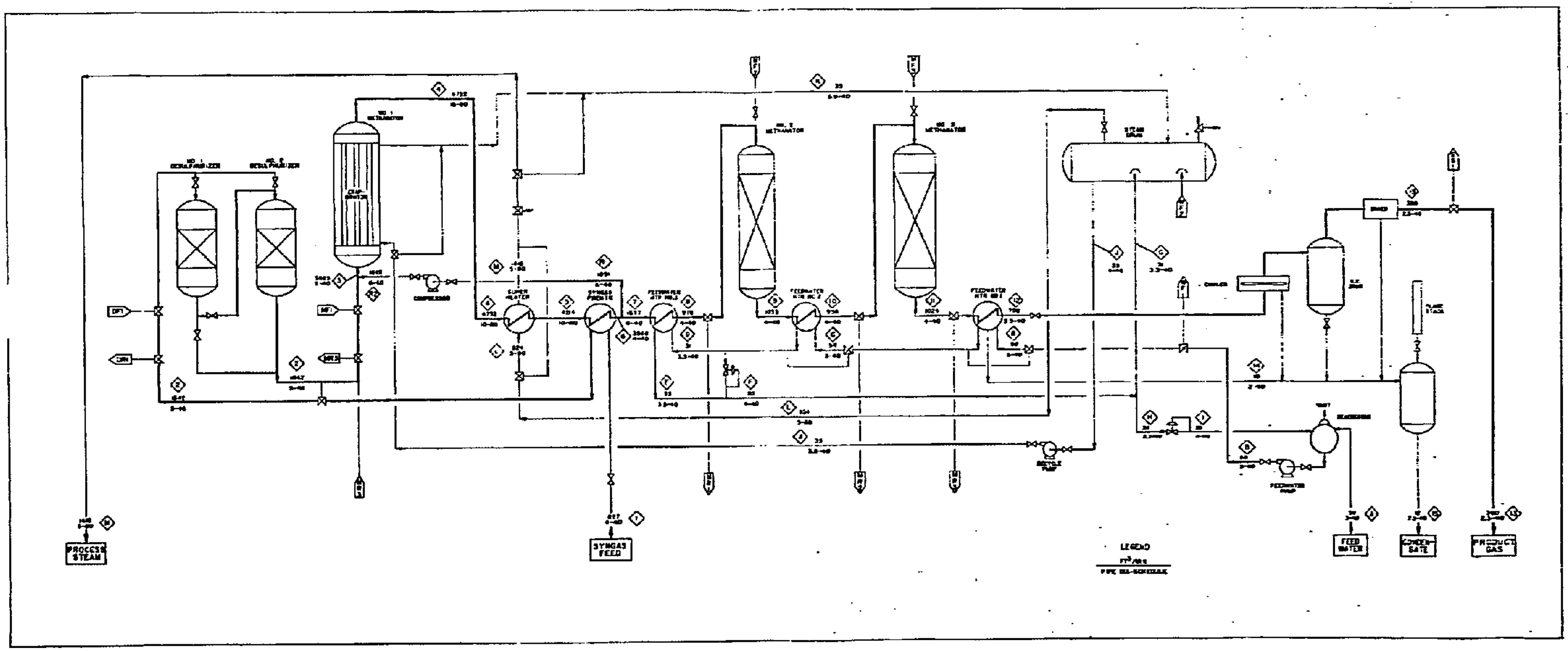

FIGURE 6.2. Piping Diagram, 40 MW Plant, 3 Methanators

6.1.2 Partial Capacity Operation

The reference 40 MW_t Methanation System can be operated at any desired capacity; from as low as 20% (8 MW_t) up to 110% (44 MW_t) of its nominal rating.

The 110% operation requires a 10% increase in syngas feed flow and 12% increase in feedwater flow. This would require an increase in pressures of about 6 psia for both. The end results would be no changes in the temperature of the syngas to methane stream but the superheated steam would be delivered to the distribution system at about 890°F and 906 psia.

For decreased capacity operation the syngas feed and feedwater flow would be reduced proportionally, with the pressures held at nominal 900 psia. The ratio of recycle gas to feed syngas to the combination methanator/evaporator must remain the same as for the rated capacity operation. At reduced capacity operation all heat exchangers will be oversized. To maintain desired mass and energy transfer in the heat exchangers, all heat exchangers will have a "by-pass" with flow control valves in the water lines. This procedure will maintain all temperatures and pressures of both syngas to methane and water to steam the same as for rated capacity operation and only the quantity of methane and steam produced per unit of time will be reduced.

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:

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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 stabil-

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ized 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

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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

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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.

6.2.4.1 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-

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ually give up its heat, at a rate much lower than 150°F per hour, so no damage to the equipment would result.

6.3 Methanation System Components

Brief descriptions of the major components of the Methanation System are as follows: (Presented in the sequence of syngas-methane flow).

6.3.1 Syngas Heater

The syngas heater is a tube and shell heat exchanger. Syngas from the pipeline at 100°F and 900 psia flows through the shell side where it is heated by process gas (tube side) to 651°F and then flows to the desulfurizers. See Figure 6.3.

6.3.2 Desulfurizers

Syngas flows from the syngas heater through the desulfurizers which uses a packed bed of Cobalt-Molybdenum and Zinc Oxide catalyst to absorb sulfur. There are two desulfurizers in series, either of which is able to absorb most all the sulfur in the syngas. The syngas leaves the desulfurizers with less than 0.1% ppmv sulfur content. The desulfurizers have very little, if any effect on the syngas temperature and pressure, see Figure 6.4.

6.3.3 Combination Methanator/Evaporator (C M/E)

The syngas flows from the desulfurizers to the inlet of the C M/E where it mixes with recycle process gas. The mixture then flows through catalyst in tubes which is the first methanator. The mixture enters the C M/E at 651°F and 898 psia and exits at 1015°F and 880 psia. The C M/E is an iso-

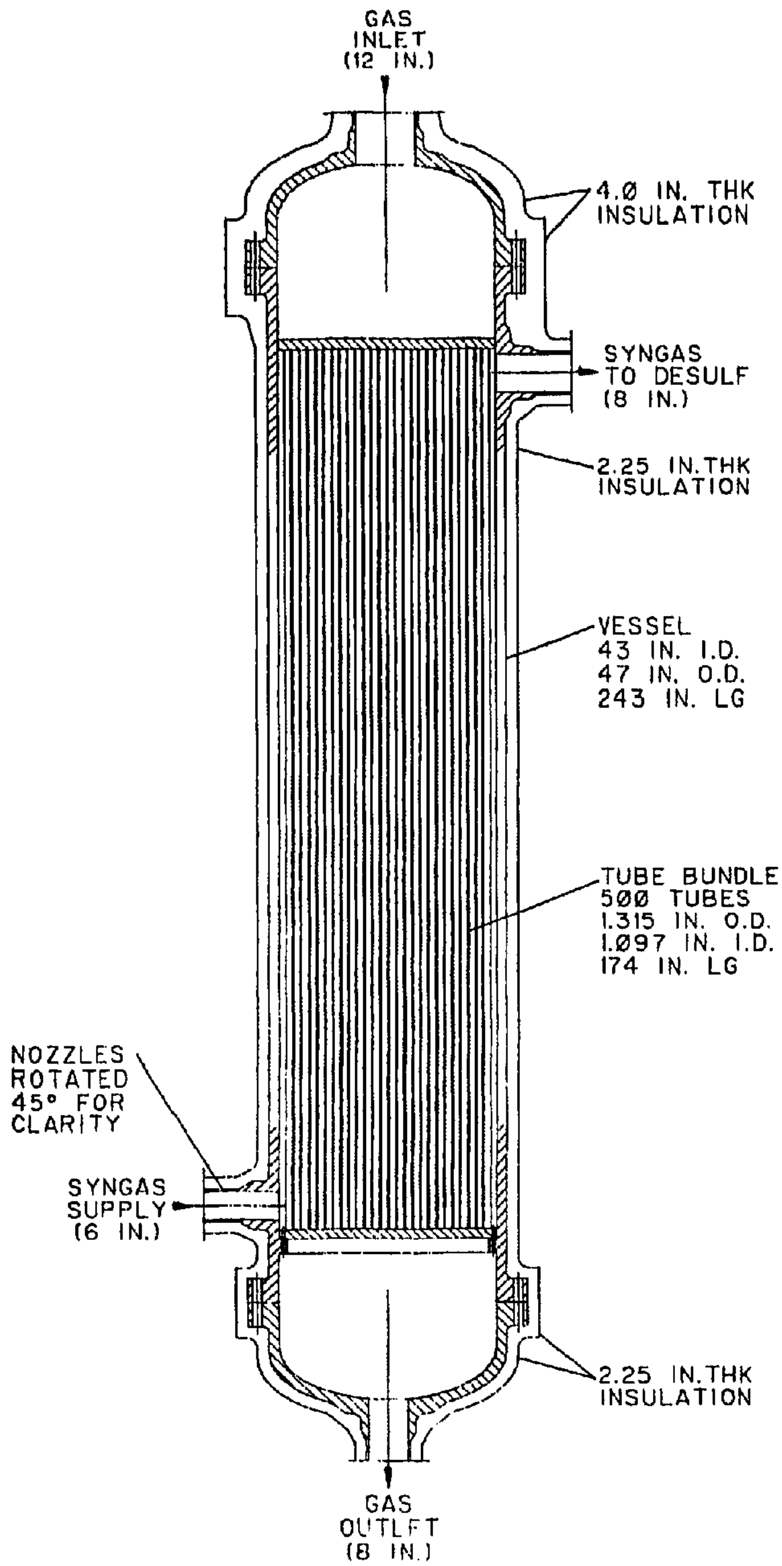

FIGURE 6.3. Syngas Heater 40 MW Plant

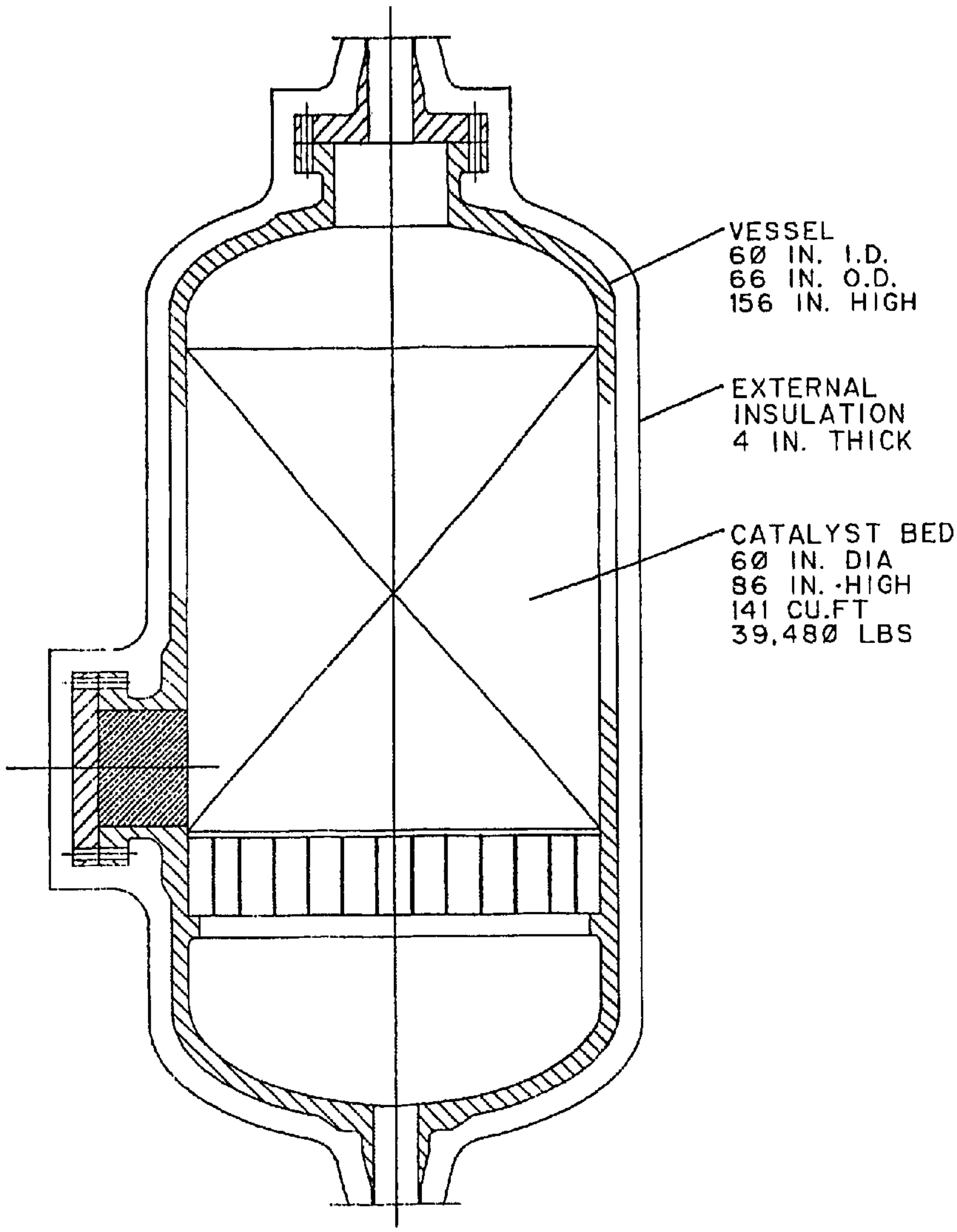

FIGURE 6.4. De-Sulfurizer 1 & 2 40 MW Plant

thermal reactor in that 38.6 MW_t of heat is produced by conversion of syngas to methane and 21.6 MW_t is transferred to the water between the tubes outer diameters and the vessel wall. Water enters the C M/E at 532°F and 910 psia and exits at 532°F and 905 psia. It is desired that about 50% of the water leaving the C M/E be saturated steam. See Figure 6.5.

6.3.4 Superheater

Process gas (the syngas is now converted to mostly methane and steam) flows from the C M/E through the tubes of the superheater, where heat is transferred to the steam on the shell side of the heat exchanger. Process gas enters the superheater at 1015°F and 880 psia and leaves at 852°F and 879 psia. Saturated steam flows from the steam drum to the superheater at 532°F and 905 psia and leaves the superheater at 900°F and 900 psia. See Figure 6.6.

6.3.5 Syngas Heater

The process gas flows through the tubes of the syngas heater described in 6.3.1, above. Upon leaving the syngas heater the process gas flow is divided with one portion being recycled to the entrance of the C M/E and the remainder flowing to the Feedwater Heater No. 3.

6.3.6 Feedwater Heater No. 3

The process gas flows through the tubes of this water heater and water flows through the shell side. The process gas enters at 607°F and 878 psia and leaves at 536°F and 877 psia. The water flowing from Feedwater Heater No. 2 enters at 389°F and 913 psia and leaves at 475°F and 911 psia. See Figure 6.7.

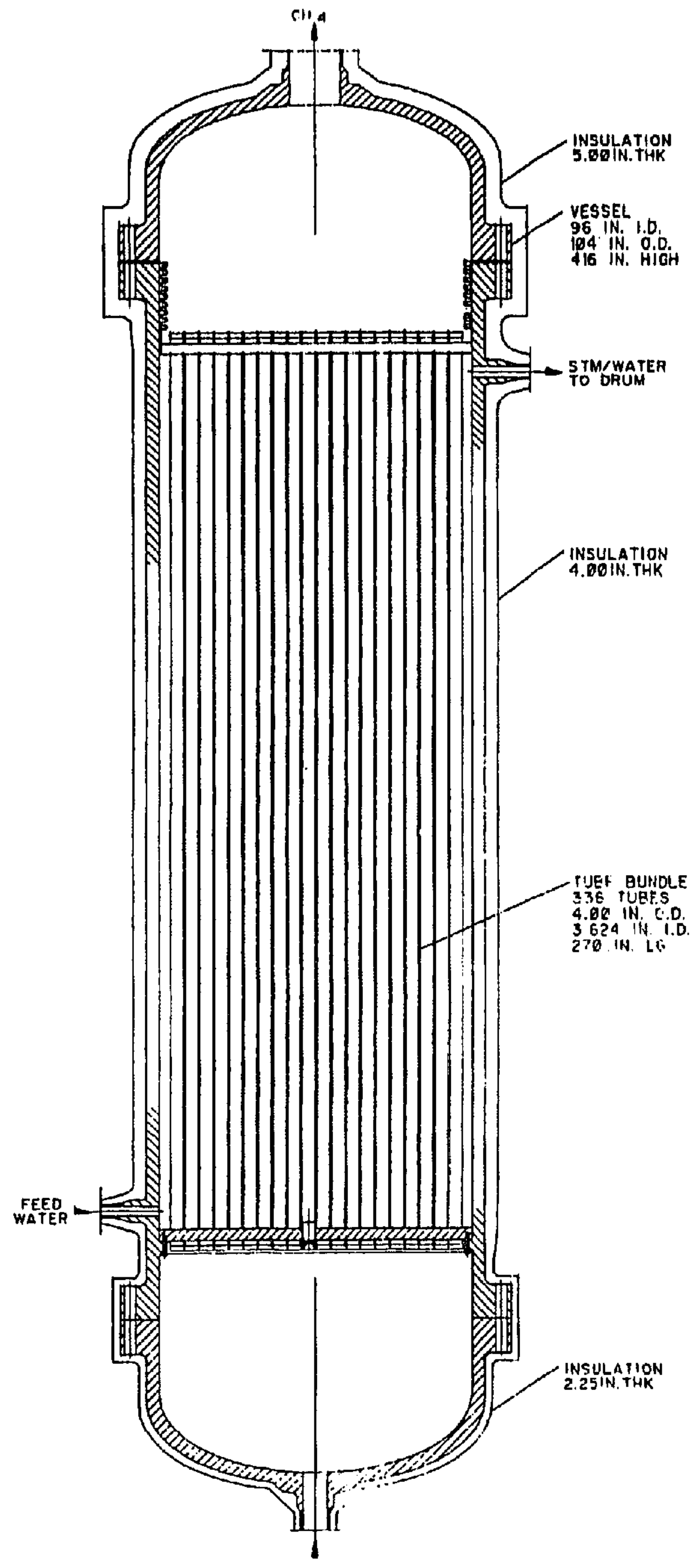

FIGURE 6.5. Comb Meth/Evap 40 MW Plant

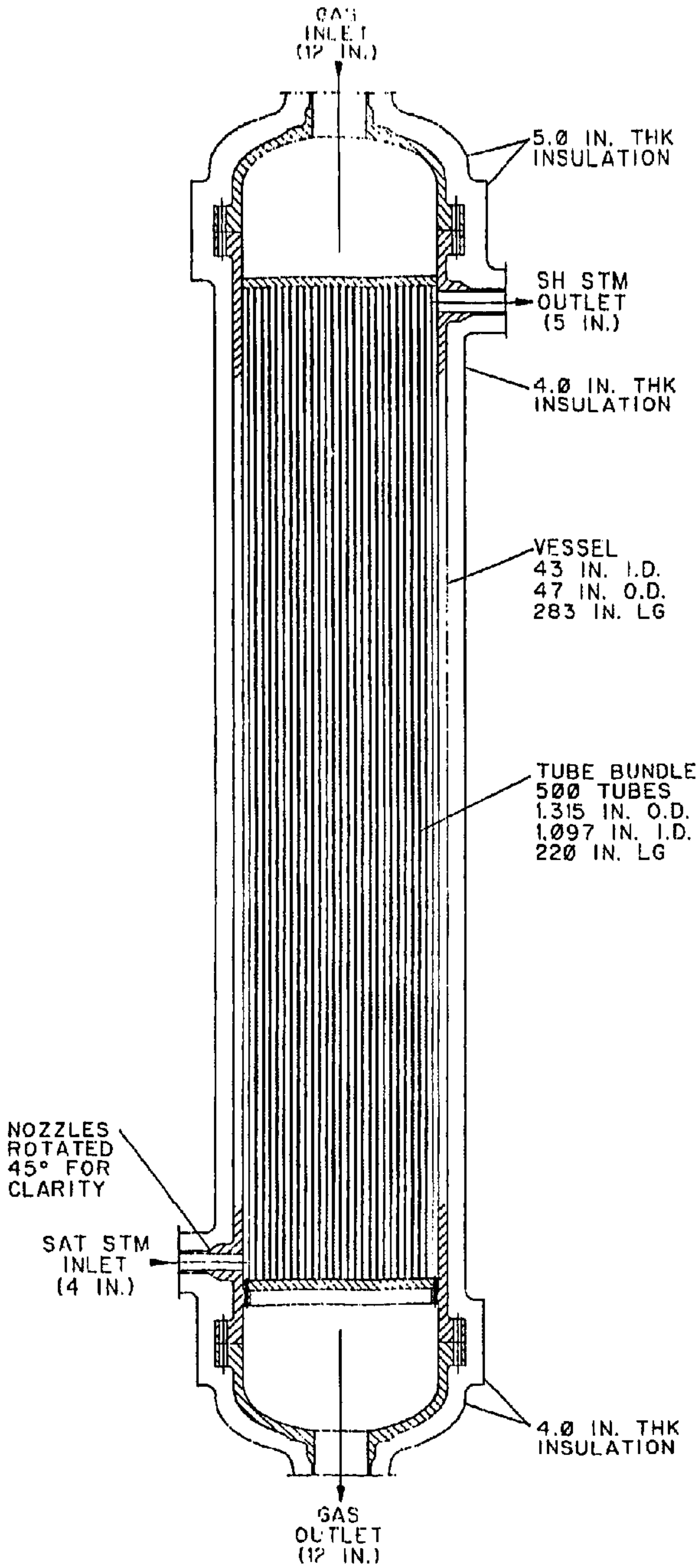

FIGURE 6.6. Superheater 40 MW Plant

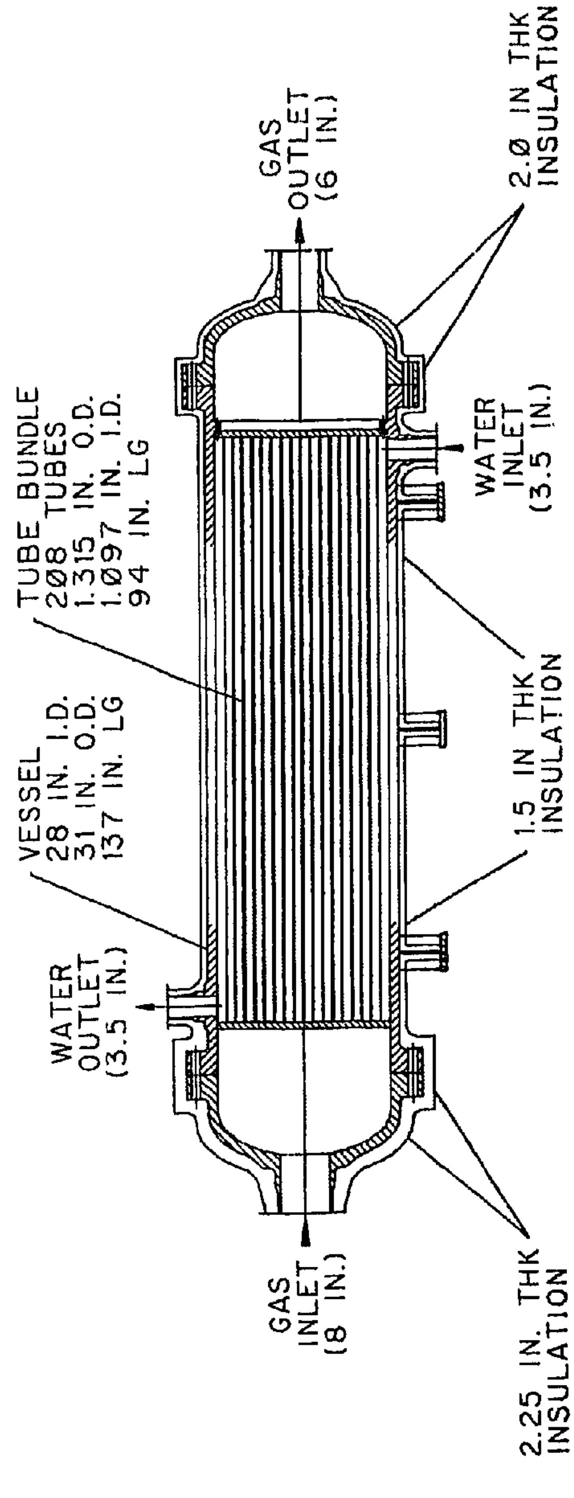

FIGURE 6.7. Feedwater Heater No. 3 40 MW Plant

6.3.7 Methanator No. 2

The process gas flows from Feedwater Heater No. 3 to Methanator No. 2. This methanator is an adiabatic packed bed catalytic reactor. As process gas flows through the bed, additional conversion of syngas to methane occurs and 2 MW_t of heat is added to the process gas. The process gas enters Methanator No. 2 at 536°F and 877 psia and leaves at 637°F and 876 psia. See Figure 6.8.

6.3.8 Feedwater Heater No. 2

The process gas flows through the tubes of this heat exchanger and transfers heat to water flowing in the shell side. The process gas enters Feedwater Heater No. 2 at 637°F and 876 psia and leaves at 536°F and 875 psia. Water from Feedwater Heater No. 1 enters at 360°F and 915 psia and leaves at 389°F and 913 psia. See Figure 6.9.

6.3.9 Methanator No. 3

This methanator is an adiabatic packed-bed catalytic reactor, identical to Methanator No. 2. As process gas flows through the bed the final conversion of syngas to methane occurs and 1.5 MW_t of heat is added to the process gas. Process gas enters this methanator at 536°F and 875 psia and leaves at 615°F and 874 psia.

6.3.10 Feedwater Heater No. 1

The process gas flows from Methanator No. 3 thorugh the tubes of this heat exchanger and transfers heat to water flowing on the shell side. Process gas enters Feedwater Heater No. 1 at 615°F and 874 psia and leaves at 280°F and 873 psia. Since the process gas temperature passes through the saturation temperature of 528°F most of the steam in the process gas conden-

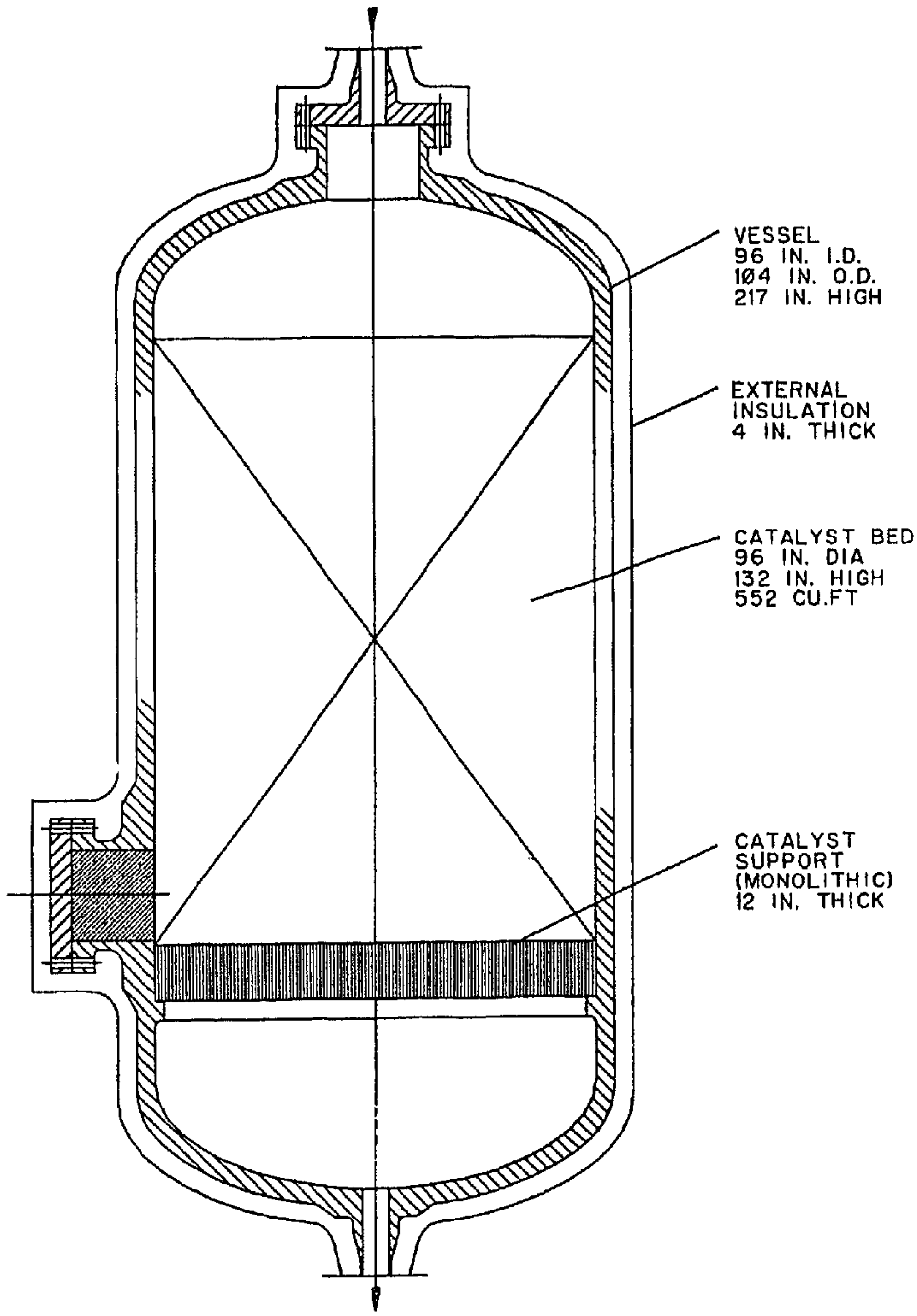

FIGURE 6.8. Methanator 2 & 3 40 MW Plant

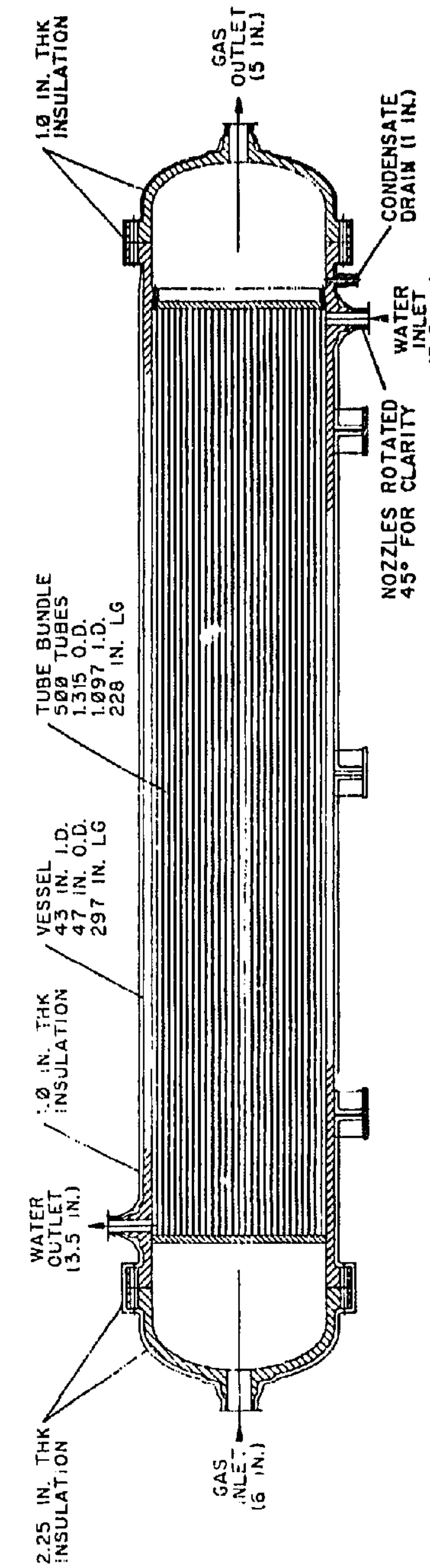

FIGURE 6.9. Feedwater Heater No. 2 40 MW Plant

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ses and the condensate is drained to a condensate holding tank. Water from the Feedwater Pumps enters the Feedwater Heater No. 1 at 228°F and 920 psia and leaves at 360°F and 915 psia. See Figure 6.10.

6.3.11 Remaining Syngas to Methane Components

From Feedwater No. 1 the process gas; now 95% methane on a dry basis, flows through an air-cooler, knock-out drum and a dryer. Condensate from these components flows to the condensate drum and then to the HTGR-R plant via pipeline. The dry methane at 100°F and 870 psia is also returned to the HTGR-R plant via a pipeline.

6.3.12 Deaerator

Condensate returned from the steam users, mixed with demineralized make-up water enters the deaerator at about 100°F and 40 psia. In the deaerator it mixes with 228°F at 20 psia steam and entrained gases are vented to atmosphere. The design capacity of the deaerator is 200,000 lbm per hour.

6.3.13 Feedwater Pump

Two feedwater pumps are provided to pump water from the deaerator through the feedwater heaters and then to the steam drum. The capacity of each pump is 210 gpm and each is driven by a 160 HP electric motor. The inlet/outlet pressure is 20/925 psia.

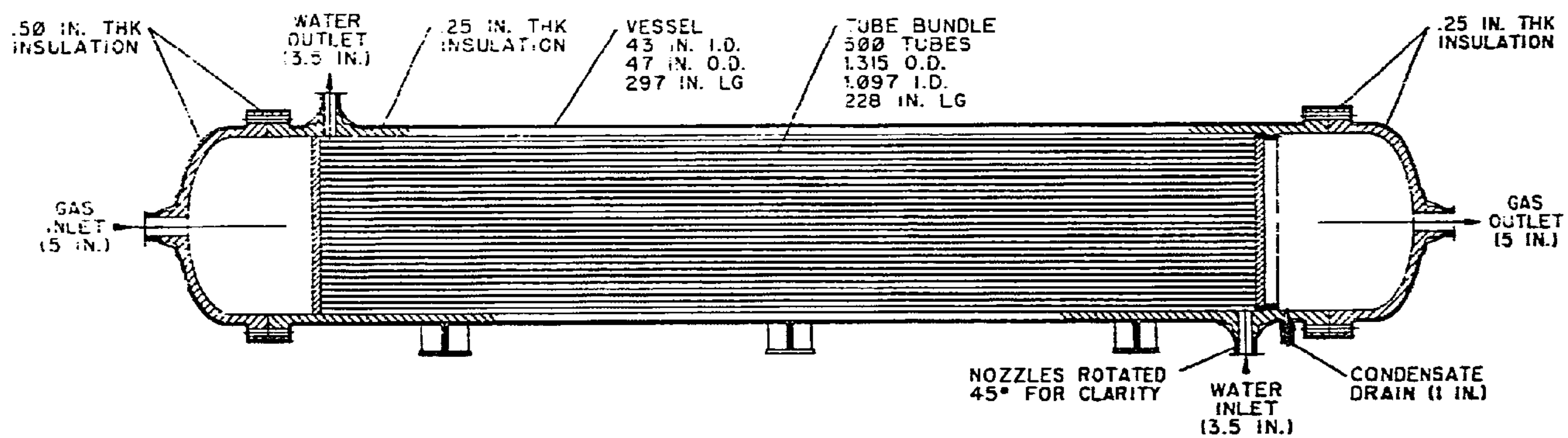

FIGURE 6.10. Feedwater Heater No. 1 40 MW Plant

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6.3.14 Steam Drum

The steam drum receives water from Feedwater Heater No. 3 at 424°F and 905 psia at the rate of 100,000 lbm/hour. The recycle pumps pump up to 200,000 lbm/hour water from the steam drum through the C M/E and back to the steam drum. The water leaves the steam drum at 532°F and 905 psia and returns half-water and half-steam at 532°F. Steam flows at the rate of 100,000 lbm/hour from the steam drum to the superheater where the temperature is increased to 900°F before being distributed to the steam users. See Figure 6.11.

6.3.15 Recycle Pump

A recycle pump is used to pump condensate from the steam drum, through the combination methanator/evaporator and then back to the steam drum. The recycle pump capacity is 80 gpm and it is driven by a 5 HP electric motor. The inlet/outlet pressure is 905/910 psia.

6.3.16 Air Cooler

An air cooler is used to reduce the temperature of the methane and steam, flowing from Feedwater Heater No. 1, from 280°F to 100°F. The heat removal is 6×10^6 Btu/hr. The air cooler uses two 4.5 ft. diameter fans, each driven by a 7.5 HP electric motor.

6.3.17 K. O. Drum

One K. O. Drum is used to separate moisture and particulates from the methane after is has been cooled by the air cooler. The K.O. Drum is 27 in. diameter by 132 in. in vertical height. It is designed to operate at a temperature of 100°F and a pressure of 870 psia.

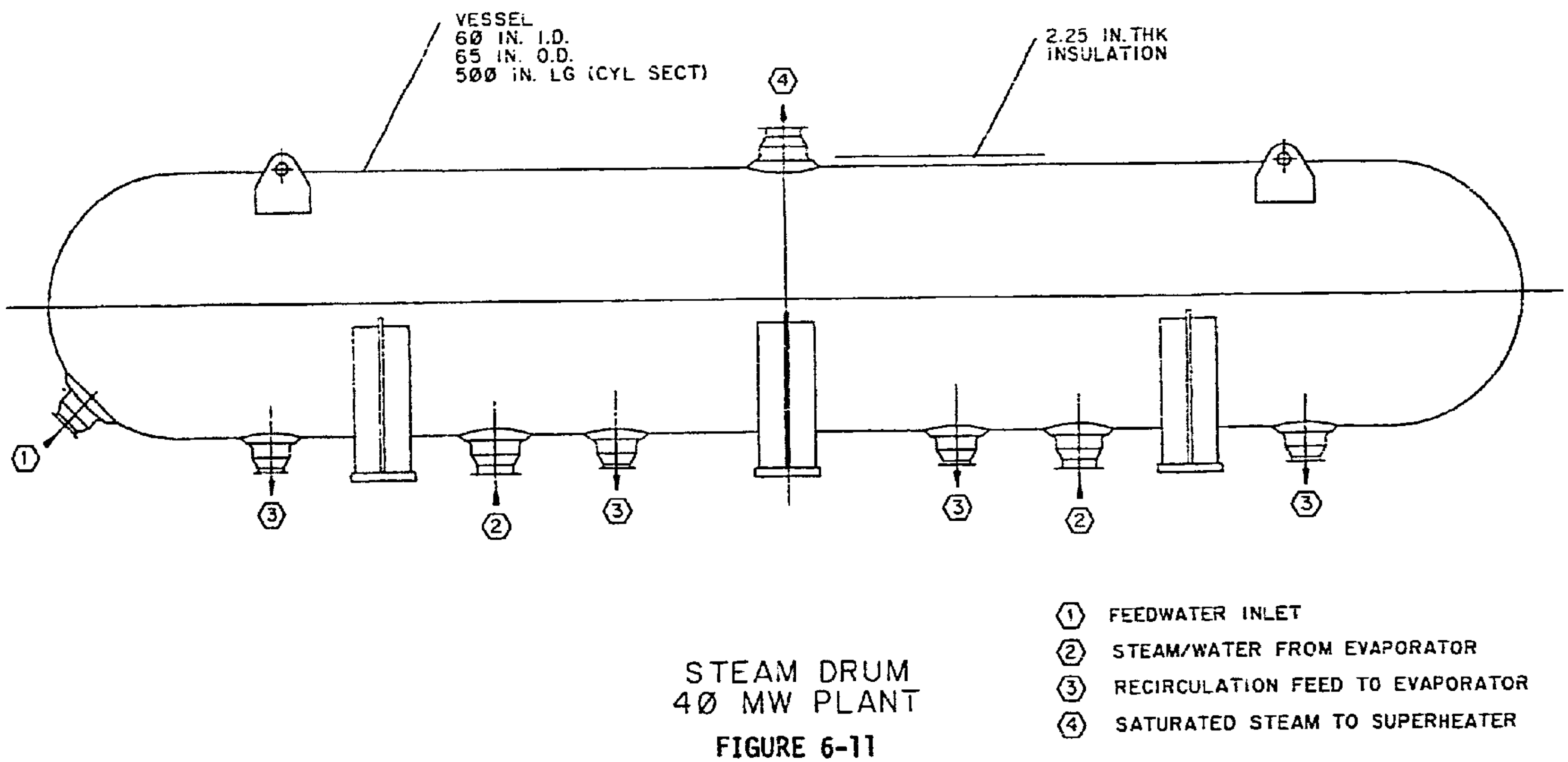

FIGURE 6.11. Steam Drum 40 MW Plant

6.3.18 Product Gas Dryer

The methane dryer is a triethylene glycol (TEG) scrubbing package unit with absorber and regenerator. It is designed to operate at 100°F and at a pressure of 870 psia. The moisture removed is 7 lbs per 1 x 10^6 scf of gas. The reboiler duty is 100,000 Btu/hr. The absorber is 2.5 ft. in diameter and 15 ft. in vertical height and the regenerator is 2 ft. in diameter and 7 ft. in vertical height.

6.3.19 Recycle Gas Compressor

Two recycle gas compressors are used to recycle process gas from the syngas heater to the combination methanator/evaporator. They are single stage centrifugal compressors, each driven by 210 HP electric motors. The inlet temperature/pressure is 651°F/878 psia and the outlet temperature/ pressure is 668°F/918 psia.

6.4 Support System Components

The components comprising the methanation system support system are used for startup, shutdown and hot-standby operation of the methanator plant.

6.4.1 Startup Gas Heater-Desulfurizer

A gas-fired heater is required to raise the temperature of the desulfurizers from ambient to 650°F at a maximum rate of 150°F per hour for startup and to maintain that temperature during hot-standby operations. The medium used for heating is nitrogen. The capacity of this gas heater is 5 x 10^6 Btu/hr and it is 8 feet in diameter and 20 feet in vertical height. The fuel may be syngas, methane or locally supplied natural gas.

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6.4.2 Startup Gas Heater-Methanators

A gas-fired heater is required to raise the temperature of the combination methanator/evaporator from 525°F to 651°F at a maximum rate of 150°F per hour during start-up. This heater is also used to raise the temperature of the two adiabatic methanators from ambient to 536°F during startup and to maintain this temperature during hot-standby operations. The medium used for heating is nitrogen or a mixture of nitrogen and hydrogen if catalyst reduction is required. The capacity of the methanators startup heater is 15 x 10^6 Btu/hr and it is 12 feet in diameter and 30 feet in vertical height. The fuel may be syngas, methane or locally supplied natural gas.

6.4.3 Startup Heater-Feedwater

A gas-fired heater is required to raise the temperature of the steam drum, three feedwater heaters, steam superheater and the combination methanator/evaporator to 525-530°F at a maximum rate of 150°F per hour, and to maintain this temperature during hot standby operation. The heating medium is water circulated by the feedwater and condensate recycle pumps. The capacity of the feedwater heater is 10 x 10^6 Btu/hr and it is 9 feet in diameter and 28 feet in vertical height. The fuel may be syngas, methane or locally supplied natural gas.

6.4.4 Startup Nitrogen Supply

Startup nitrogen is supplied from a permanently installed liquid nitrogen storage tank. Necessary recycle pumps, values and controls are also included. The storage tank capacity is 1 x 10^6 scf and the tank is 9 feet in diameter and 40 feet in vertical height.

6.4.5 Condensate Storage Tank

The condensate storage tank receives condensate from the methane/steam gas from Feedwater Heater No. 1, the Air Cooler, the K.O. Drum and the Product Gas Dryer. From this storage tank, condensate flows to a pipeline for return to the HTGR-R plant. The capacity of the condensate storage tank is 2,000 gallons. It is designed for operation at a temperature of 300°F and at a pressure of 900 psia. Size of the tank is 50 inches in diameter and 24 feet in vertical height.

6.4.6 K.O. Drum With Flare Stack

The K.O. Drum with Flare Stack is used to release nitrogen to atmosphere and to burn hydrogen, syngas, methane, and mixtures of these gases. This is a self-supporting packaged unit with 2 pilot ignitors which require gas at the rate of 160 scfh. The capacity is 350 cfm. The unit is designed to operate at a temperature of 750°F and at a pressure of 1,000 psia. The K.O. Drum base is 4 feet diameter and 9 feet vertical height and the Flare Stack is 14 inches in diameter and 32 feet in vertical height.

6.4.7 Emergency Power Supply

Emergency power is supplied by a 230 KW diesel generator. Fuel required is 150 gallons/day.

6.5 Materials and Structure Analysis

Since most of the components in the Reference Plant Design are subject to pressures of 920 psia maximum and temperatures of 651°F maximum, conventional materials approved by ASME for fabrication of Section VIII, Division 1 components are readily available. Three of the components were given special consideration for material selection.

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6.5.1 Combination Methanator/Evaporator (C M/E)

It is recommended that the vessel of the C M/E be fabricated of 2-1/4 Cr + 1 Mo with 3/16" thick internal cladding of 316 stainless steel. The same material and cladding is recommended for the tube sheets. The tubes, which contain the catalyst, should be standard pipe made of 304 stainless steel.

Structurally, the recommended materials are quite adequate for the design temperatures and pressures except for the upper tubesheet. The upper tubesheet has a stabilized temperature of 1015°F with 532°F at its bottom surface (saturated steam temperature for 905 psia). The ΔT of 483°F requires a tubesheet thickness of 24 inches with minimal allowance for fatigue failure due to thermal cycling. It is recommended that the detail design of the tubesheet includes thermal insulation to reduce the ΔT.

6.5.2 Superheater

It is planned that the same materials used for the C M/E be used in the fabrication of the superheater. The tubesheet ΔTs for the superheater are 115°F and 320°F so thermal stresses and fatigue due to cycling are much less severe. However, thermal insulation for the tubesheet with ΔT = 320°F should be considered.

6.5.3 Syngas Heater

Again it is planned to fabricate the syngas heater of the same materials as the C M/E and the superheater except that stainless steel cladding of the internal vessel walls, in contact with the syngas, is not required. The tubesheet ΔTs for the syngas heater are 201°F and 551°F. Thermal insulation of the tubesheet with 551°F ΔT will certainly be required.

6.6 Methanator Plant Arrangement

The methanator plant arrangement design was selected by GE-ARSD from three alternative arrangements developed by Bechtel. The selected arrangement provides the following features:

(a) Minimum lengths of piping and electrical and instrument conduit.

(b) A second methanation system can be added with good access to shared facilities: control building, warehouse and supporting system.

(c) Good access for mobile cranes for infrequent major maintenance operations.

The methanator plant arrangement is shown in the following figures:

Figure 6.12 - Plot Plan
Figure 6.13 - Plan Below El, 50'-0"
Figure 6.14 - Sections A-A and B-B
Figure 6.15 - Equipment Removal and Laydown Study

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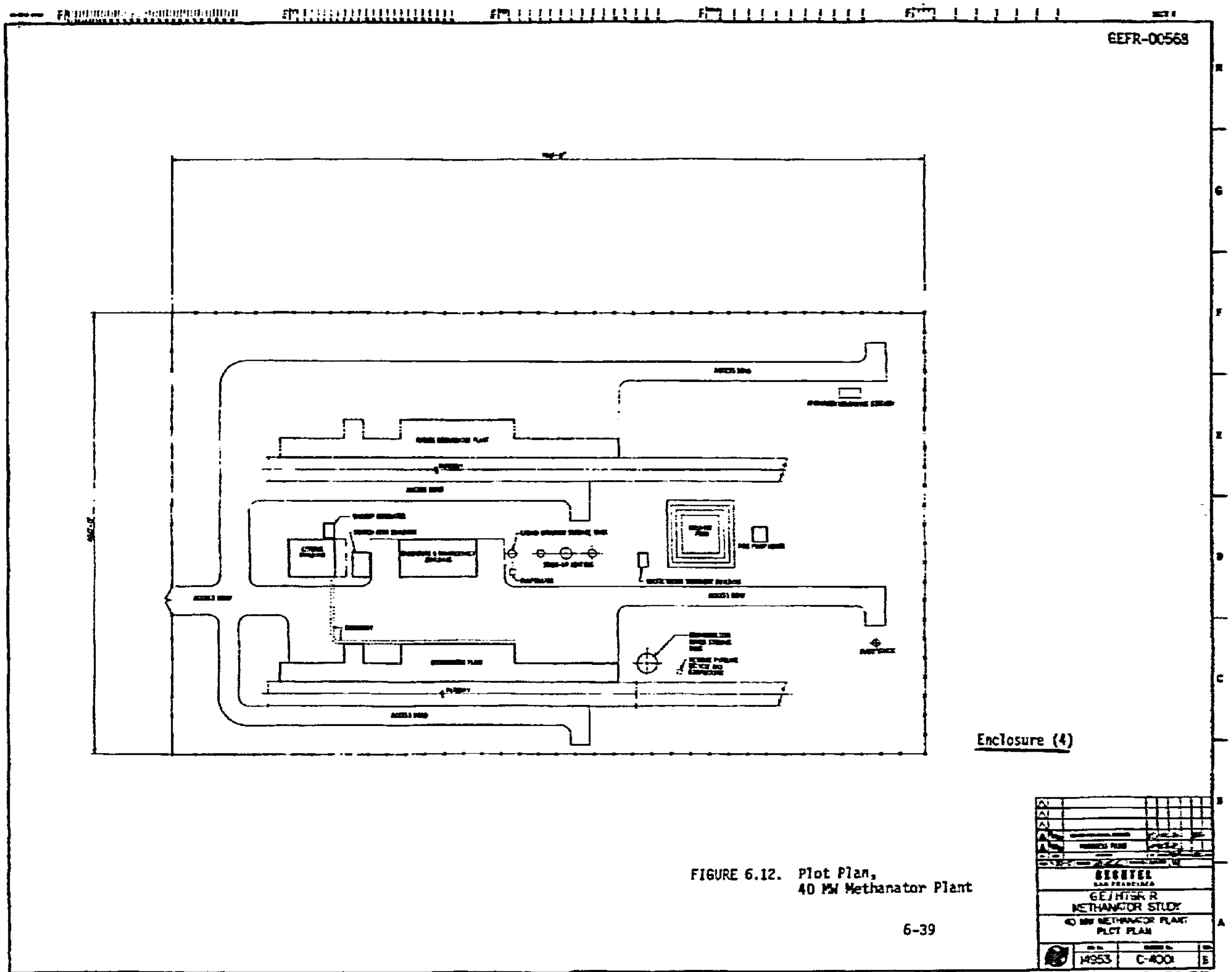

FIGURE 6.12. Plot Plan,
40 MW Methanator Plant

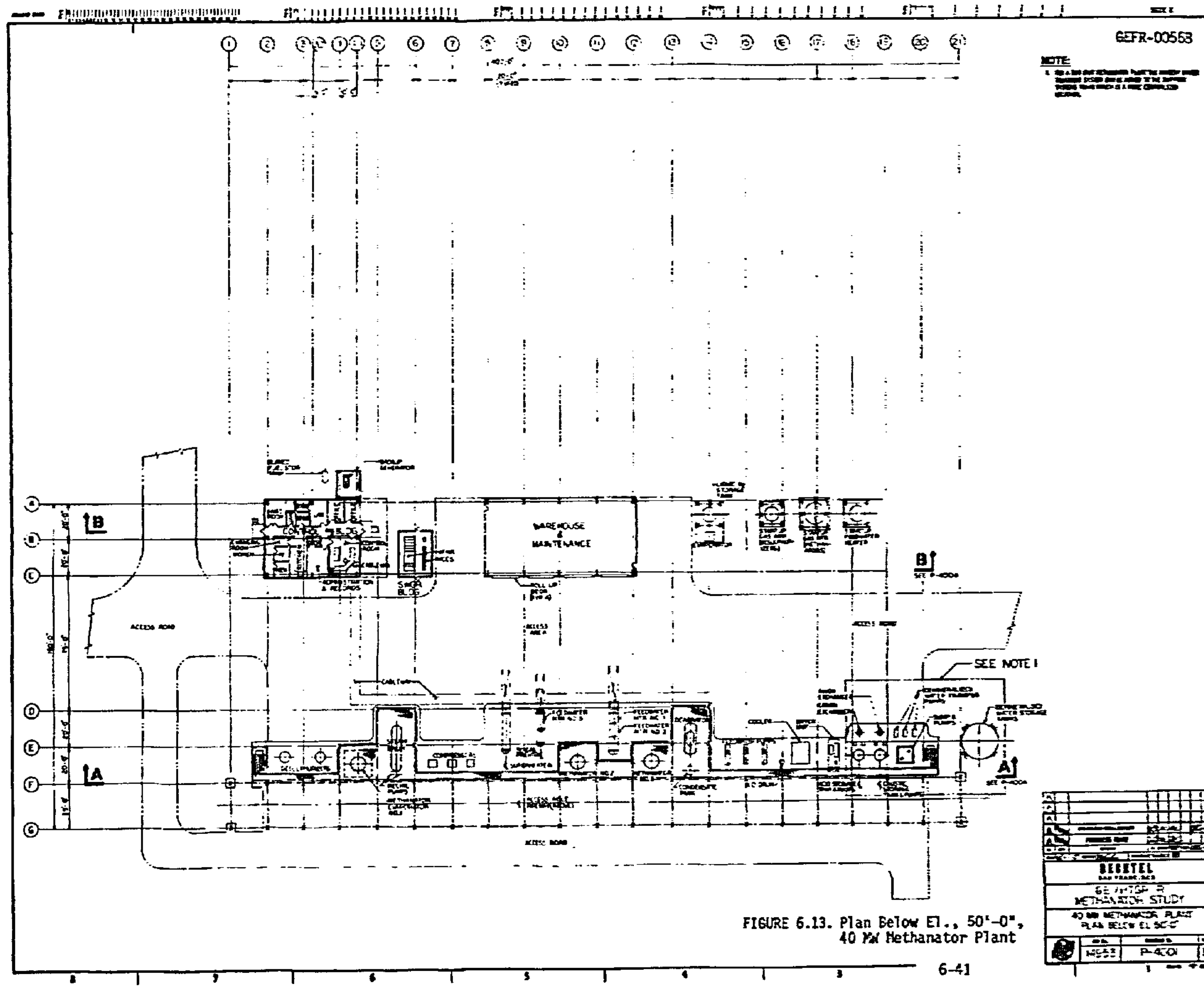

FIGURE 6.13. Plan Below El., 50'-0", 40 MW Methanator Plant

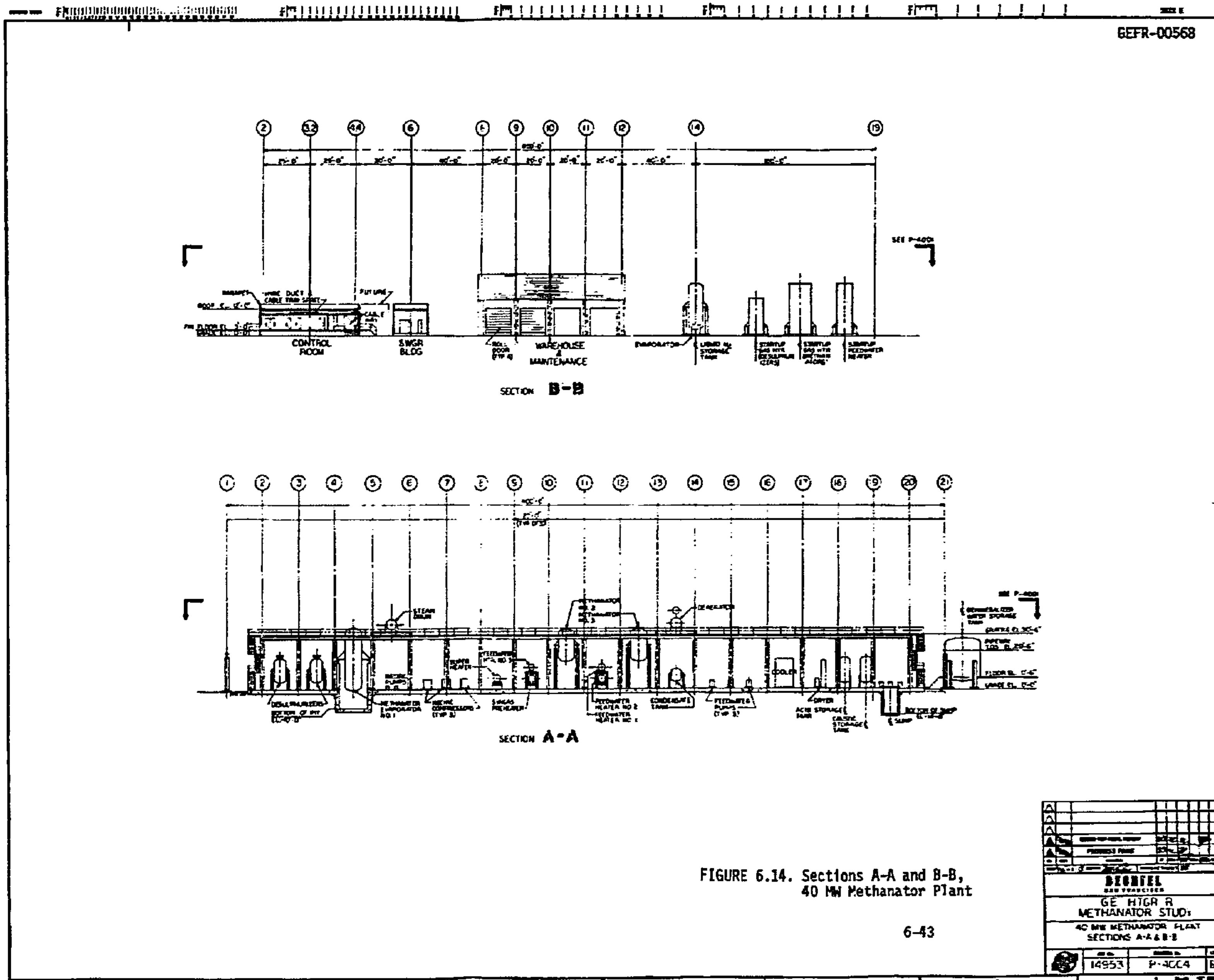

FIGURE 6.14. Sections A-A and B-B, 40 MW Methanator Plant

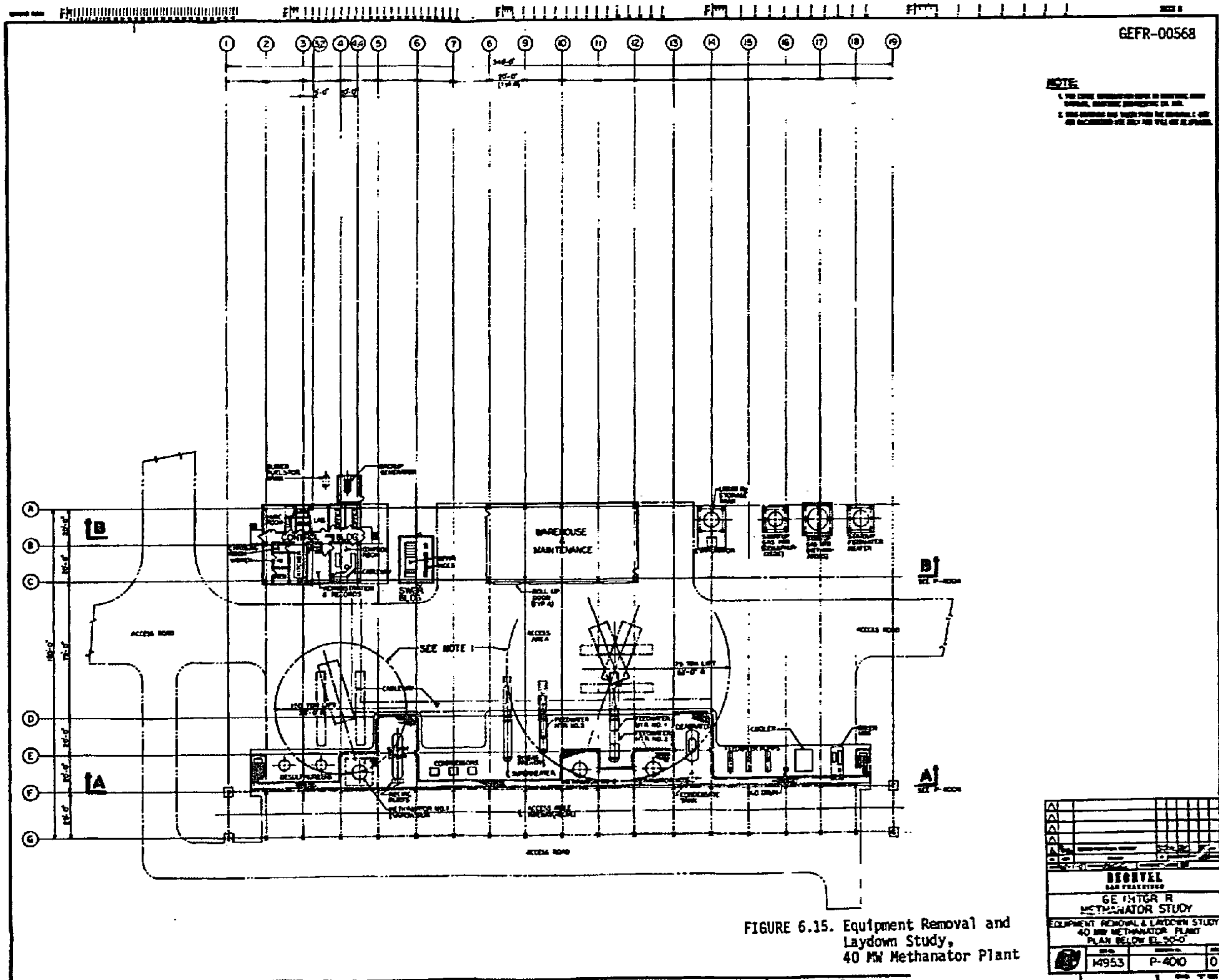

FIGURE 6.15. Equipment Removal and Laydown Study, 40 MW Methanator Plant

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SECTION 6 - REFERENCES

1. "C150 SNG Methanation Catalyst"
United Catalyst Inc., Girdler and CCI Catalyst Manufacturers,
P.O. Box 32370, Louisville, KY 40232.

7.0 PLANT DESIGN FOR SYNFUEL PROCESSES

Three additional methanation systems were designed for coupling to synfuel production processes. One system produces steam for coal gasification and the other two provide heat for retort-gas used in oil shale reduction processes.

7.1 Exxon Coal Gasification Methanation System

A methanation plant with the capacity of 720 MW_t is required for the Exxon Coal Gasification process. Four methanation trains of 180 MW_t capacity each are used. The design of a 180 MW_t train is shown in Figure 7.1. The plant produces 1.44×10^6 lb/hr (3.6×10^5 lb/hr from each train) superheated steam at 960°F and 1,250 psig. Each train is a scale-up of the reference 40 MW_t methanation system. The major components of each methanation train are given in Table 7.1.

7.1.1 Syngas to Methane Conversion

Syngas is supplied from the pipeline at 100°F and 1,285 psi. It passes through three preheaters to raise its temperature to 651°F, after which it flows through desulfurizers to reduce sulfur content to less than 0.1 ppm. the syngas stream is then split into thirds and passed through three combination methanator/evaporators located in parallel. Upon leaving the combination methanators/evaporators the gas (now about 90% methane) is recombined into a single stream and flows through two superheaters and two syngas preheaters all located in series. After the two syngas preheaters a portion of the gas is removed and returned to the feed for the combination methanator/evaporators. This recycle of part of the process gas limits the conversion of syngas to methane and thereby controls the maximum temperature of the combination methanator/evaporator. After leaving the second syngas

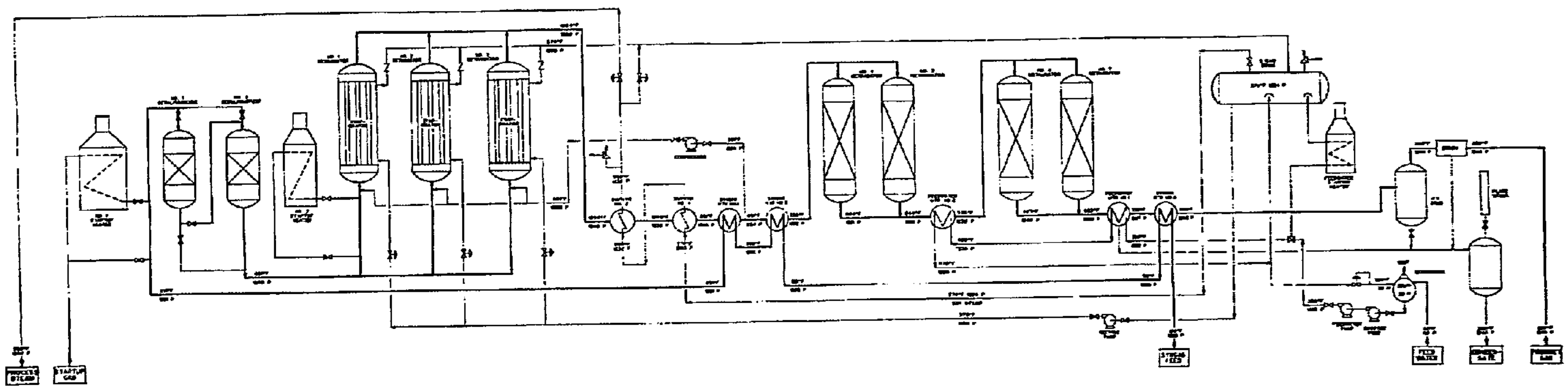

FIGURE 7.1. Flow Sheet,
180 MW Plant,
7 Methanators

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TABLE 7.1

Methanation Plant — Major Components

Item	Quantity	Figure
Combination Methanator/Evaporator	3	7-2
Adiabatic Methanator	4	7-3 & 7-4
Desulfurizer	2	7-5
Superheater	2	7-6 & 7-7
Syngas Preheater	3	7-8, 7-9 & 7-10
Feedwater Heaters	2	7-11 & 7-12
Steam Drum	1	7-13
Feedwater Pump	1	NA*
Recycle Water Pump	1	NA
Recycle Gas Compressor	1	NA
Start-Up Heater	2	NA

*NA: Not Available

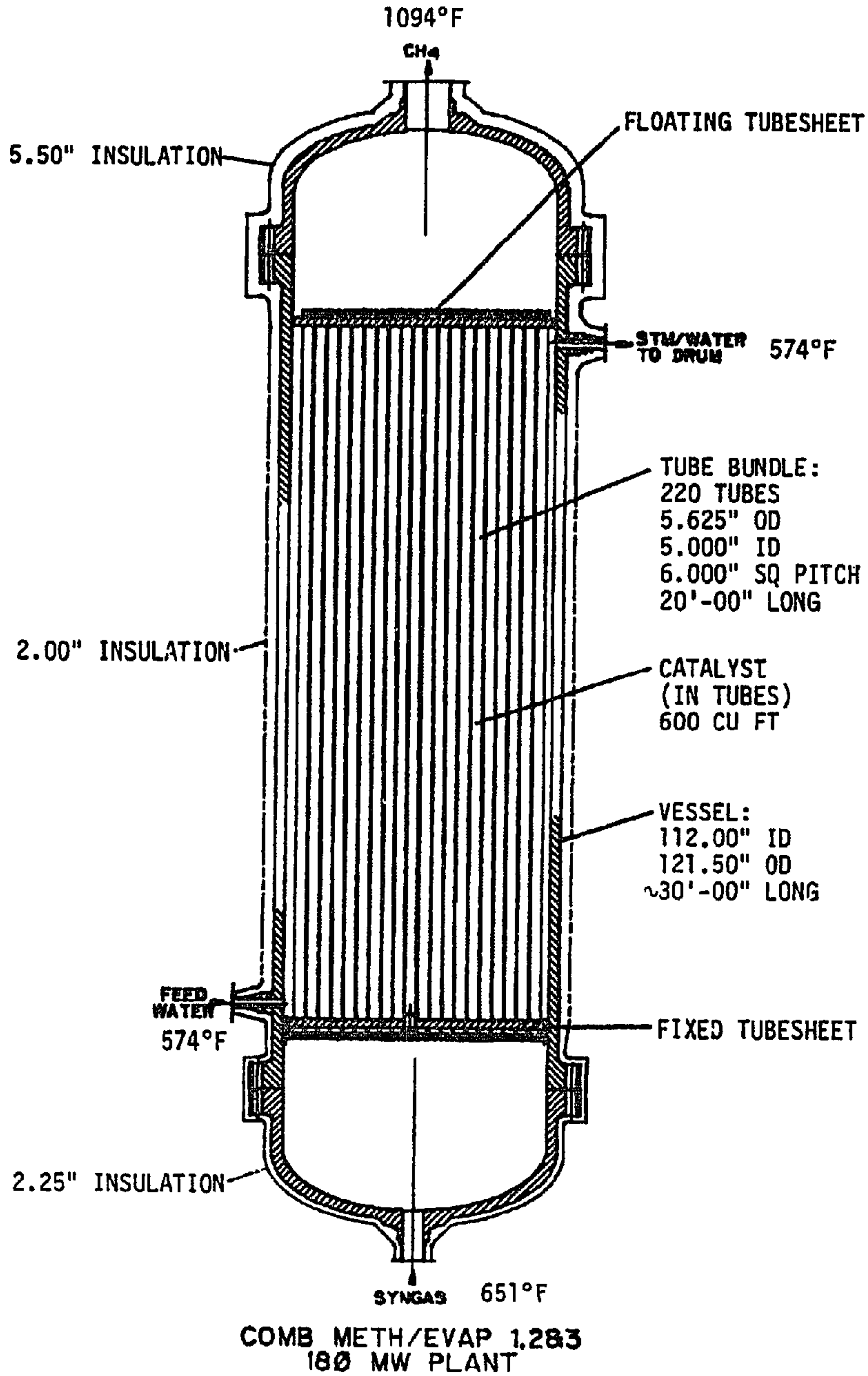

FIGURE 7.2. Comb Meth/Evap 1, 2 & 3 180 MW Plant

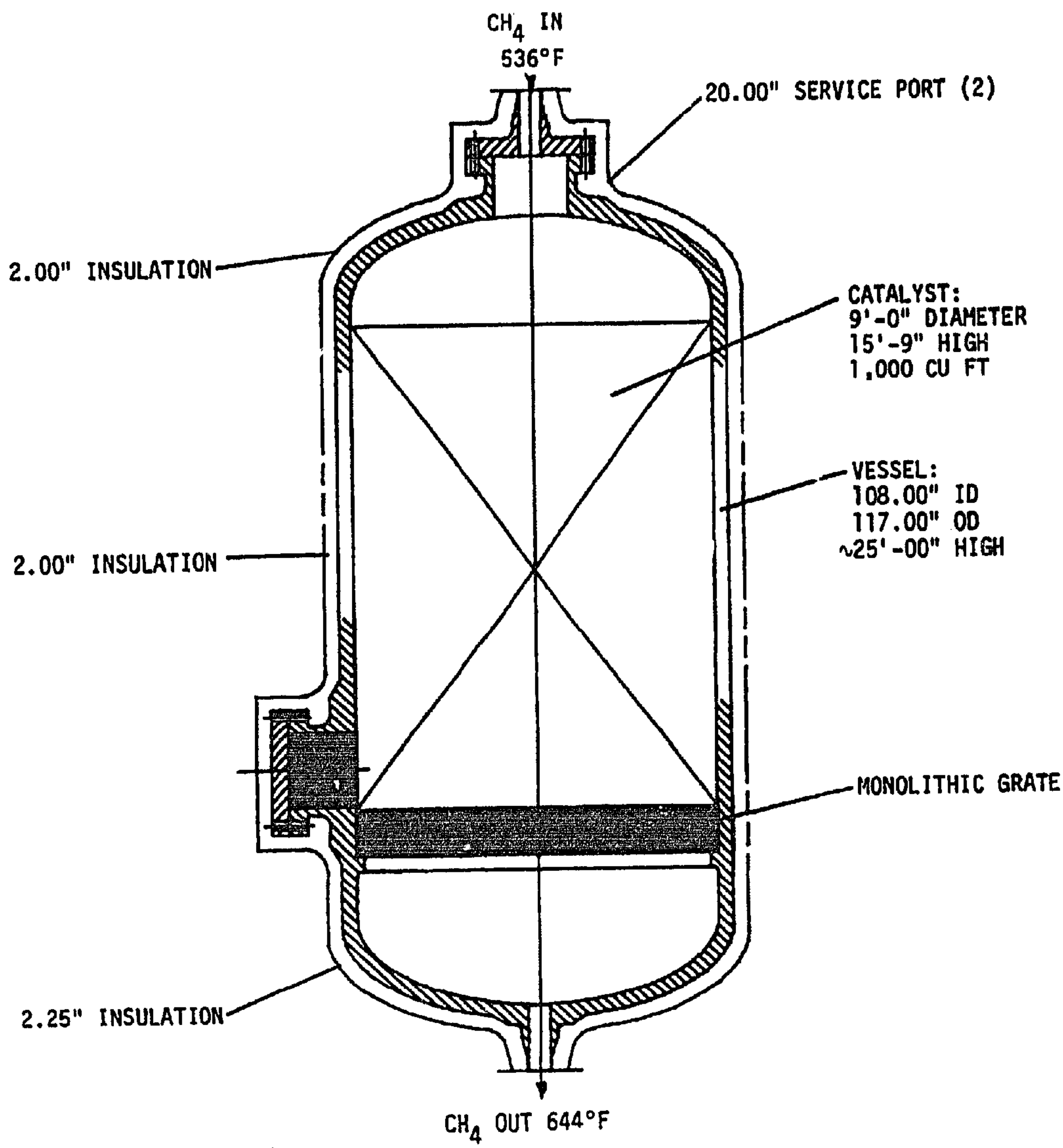

FIGURE 7.3. Methanator 4, 5, 6 & 7 180 MW Plant

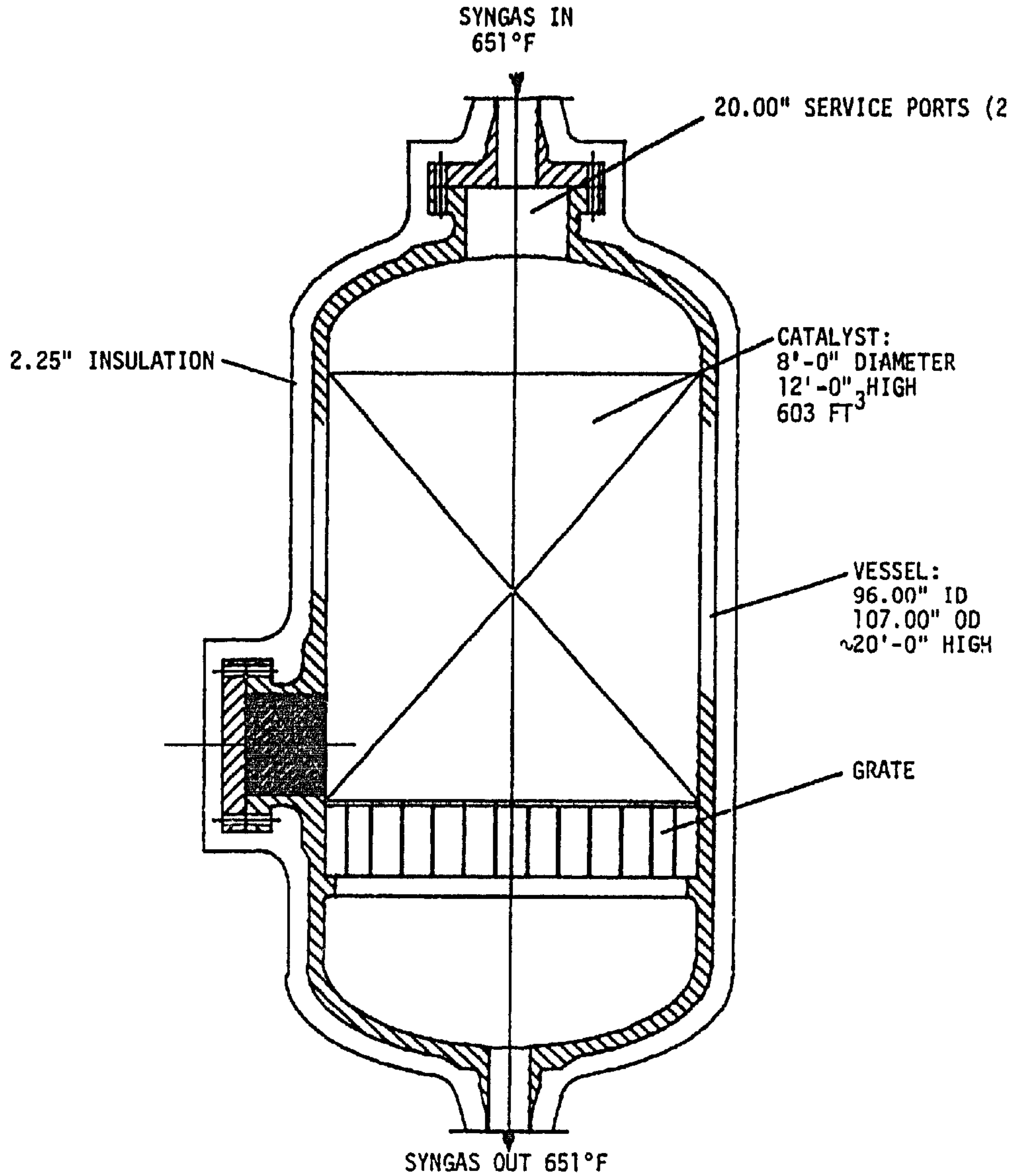

FIGURE 7.4. De-sulfurizer 1 & 2 180 MW Plant

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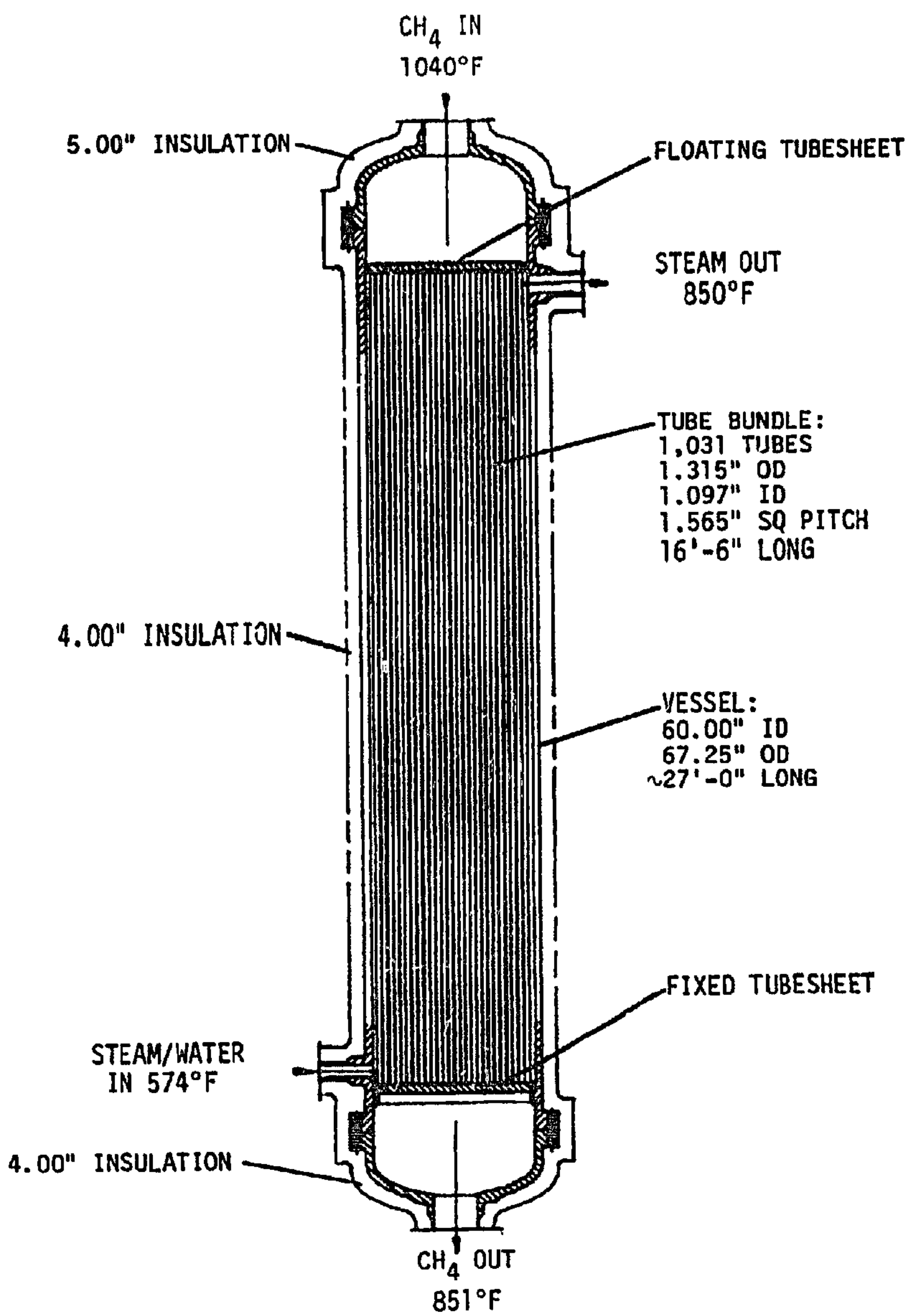

FIGURE 7.5. Superheater No. 1 180 MW Plant

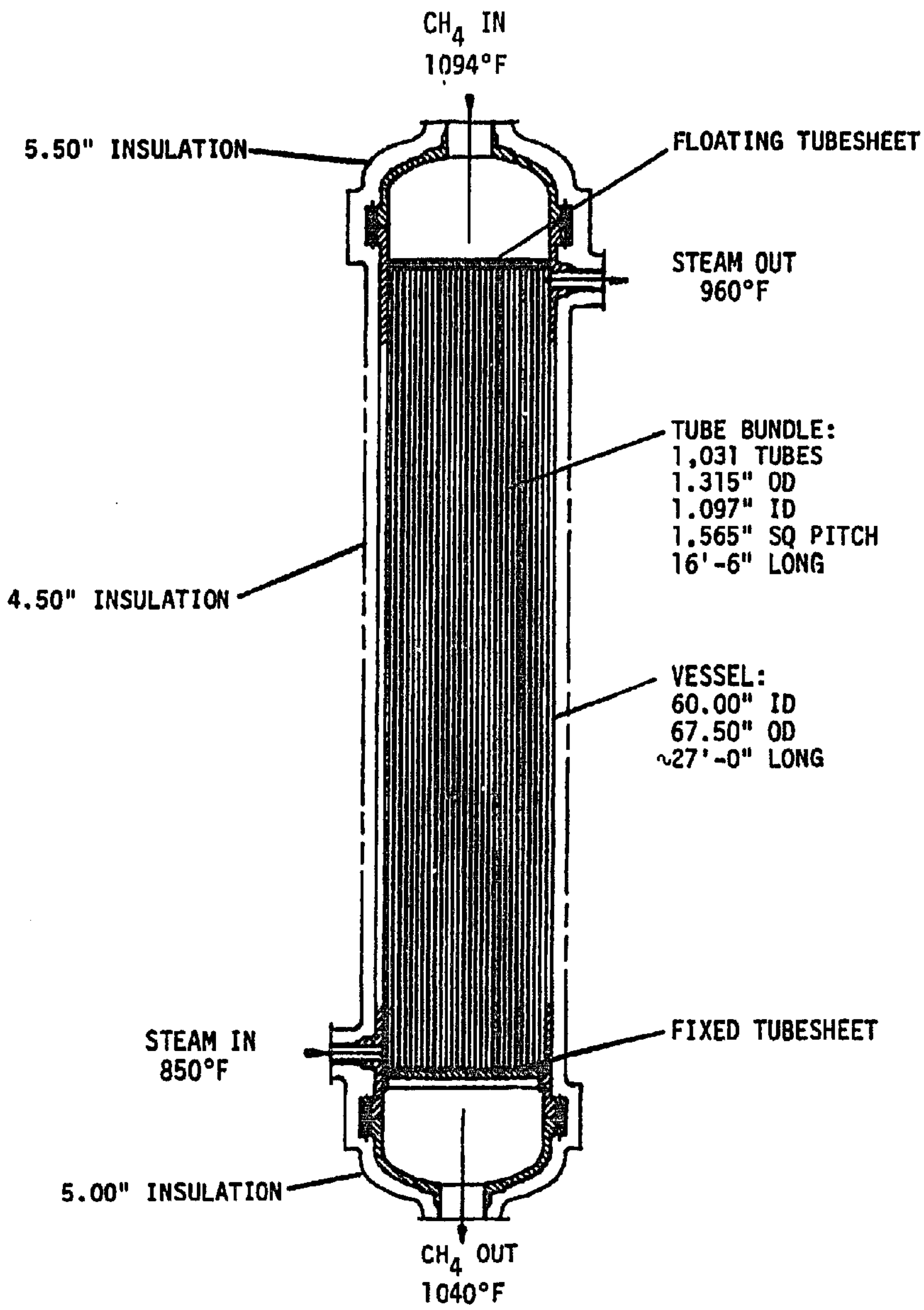

FIGURE 7.6. Superheater No. 2 180 MW Plant

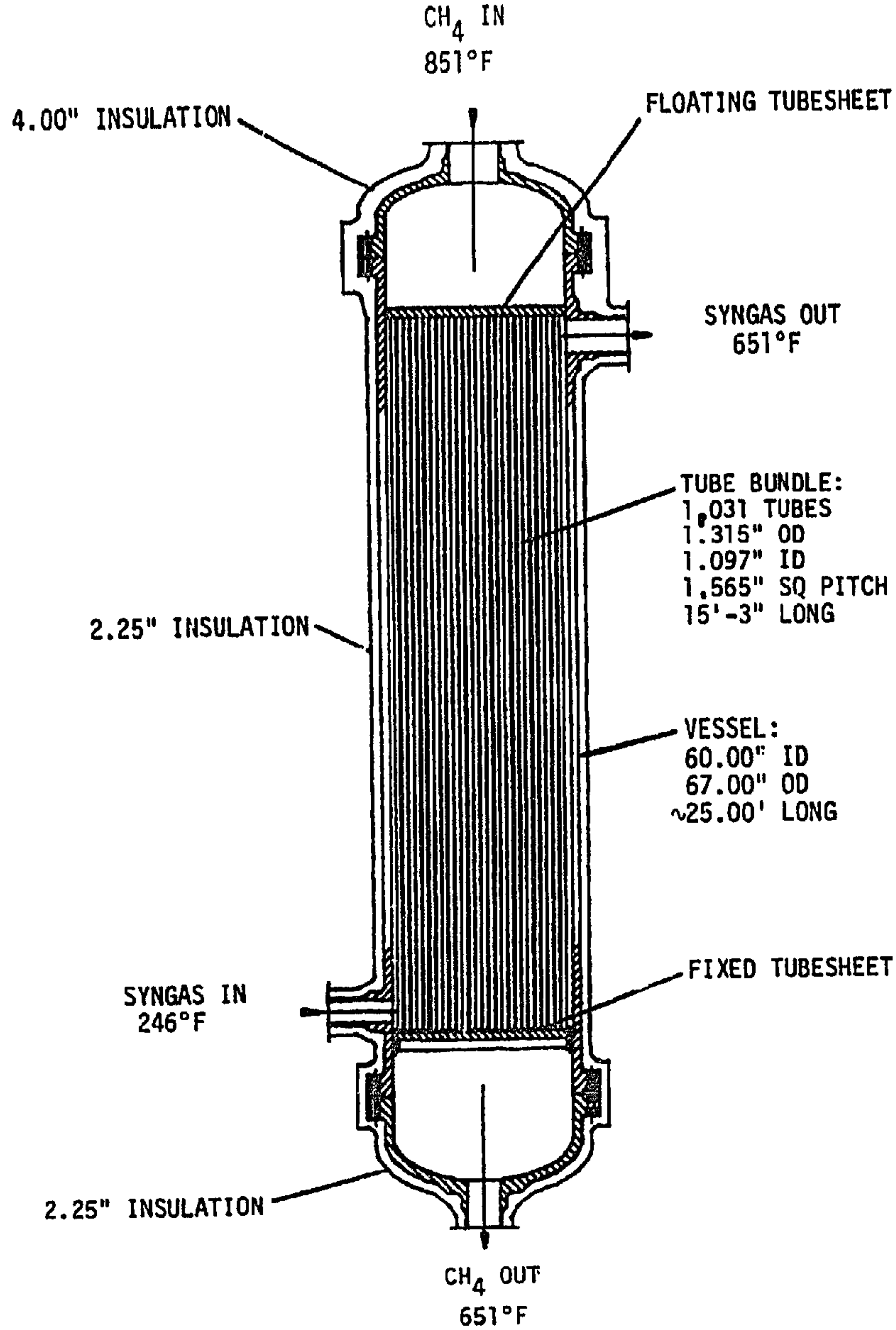

FIGURE 7.7. Syngas Heater No. 1 180 MW Plant

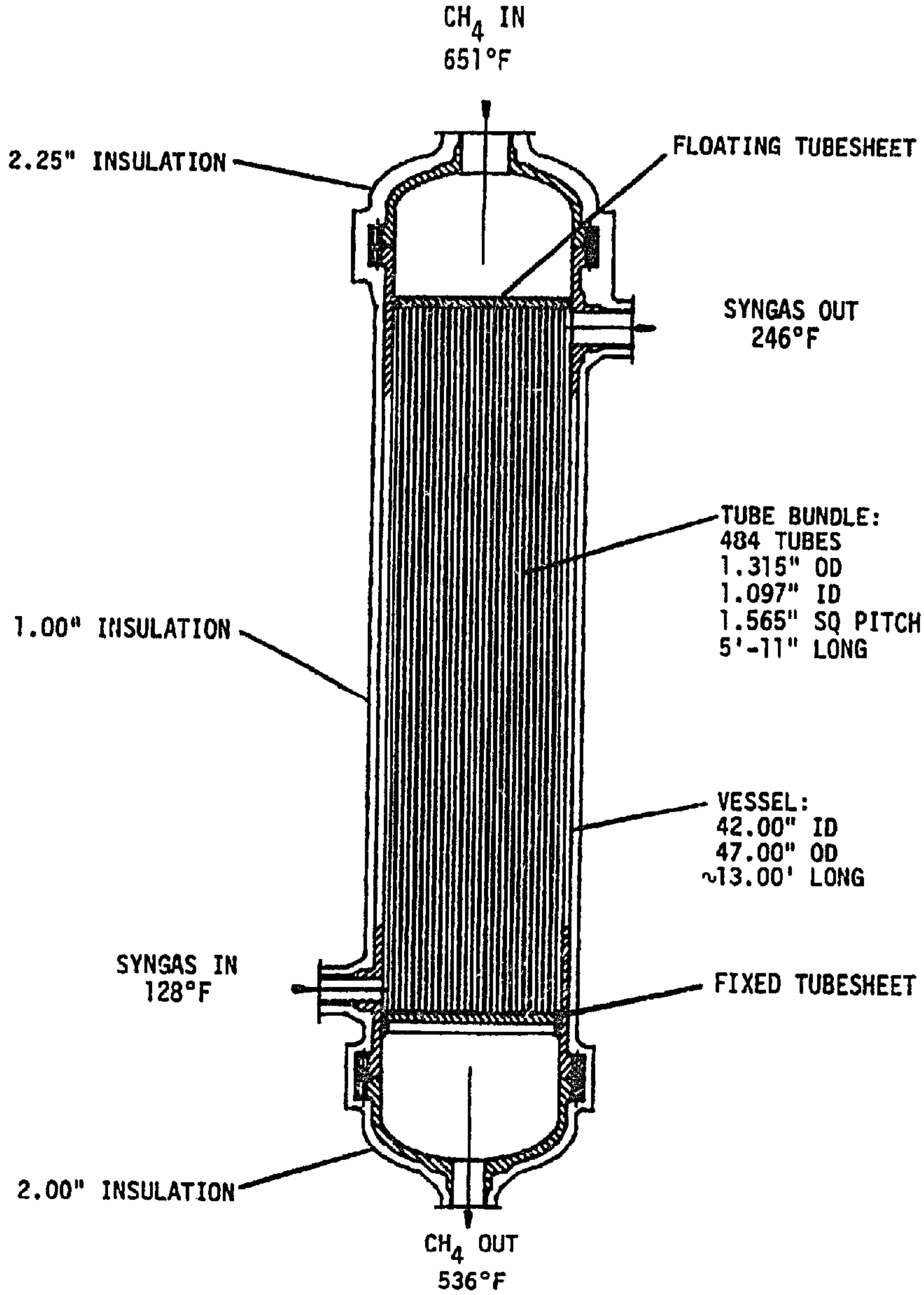

FIGURE 7.8. Syngas Heater No. 2 180 MW Plant

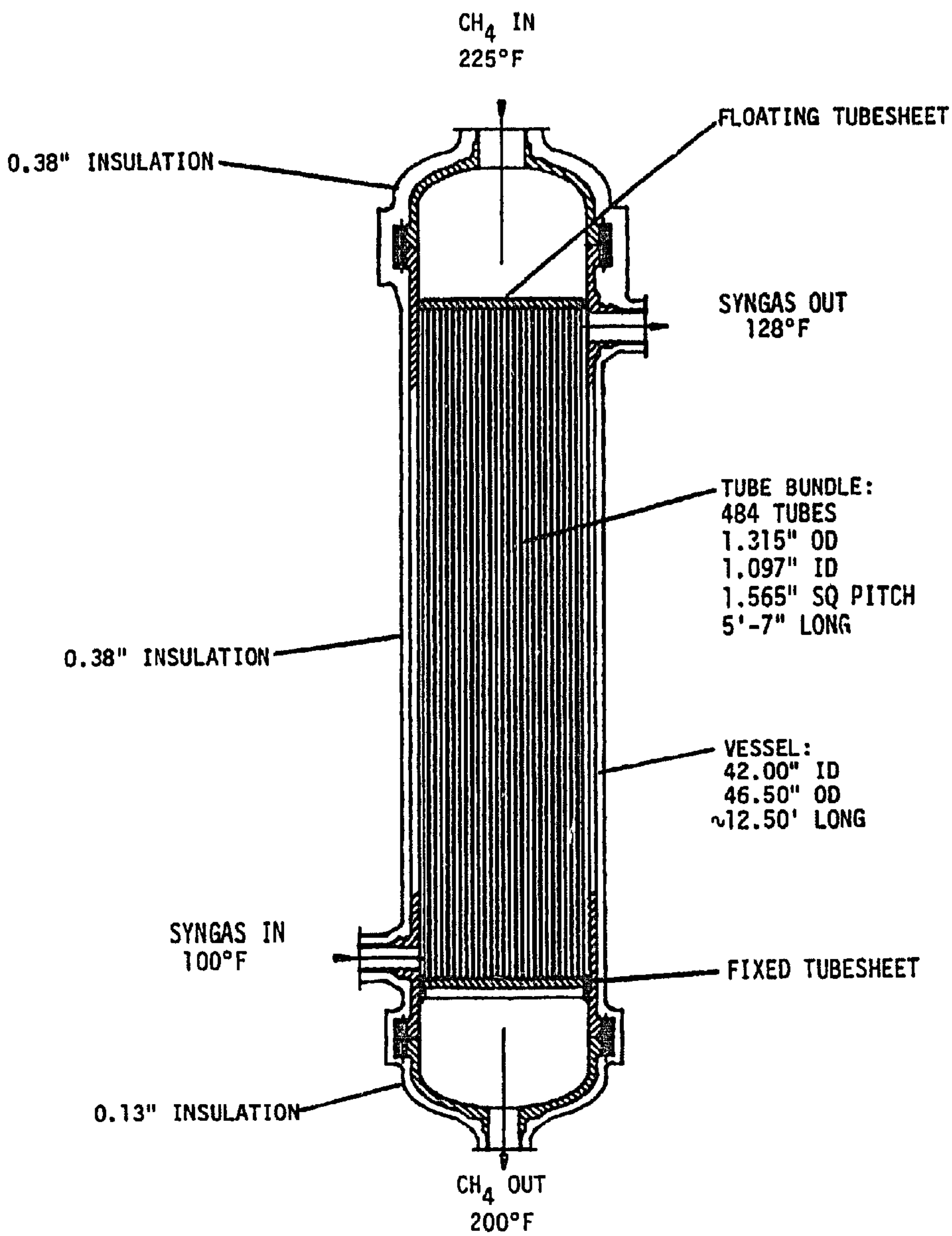

FIGURE 7.9. Syngas Heater No. 3 180 MW Plant

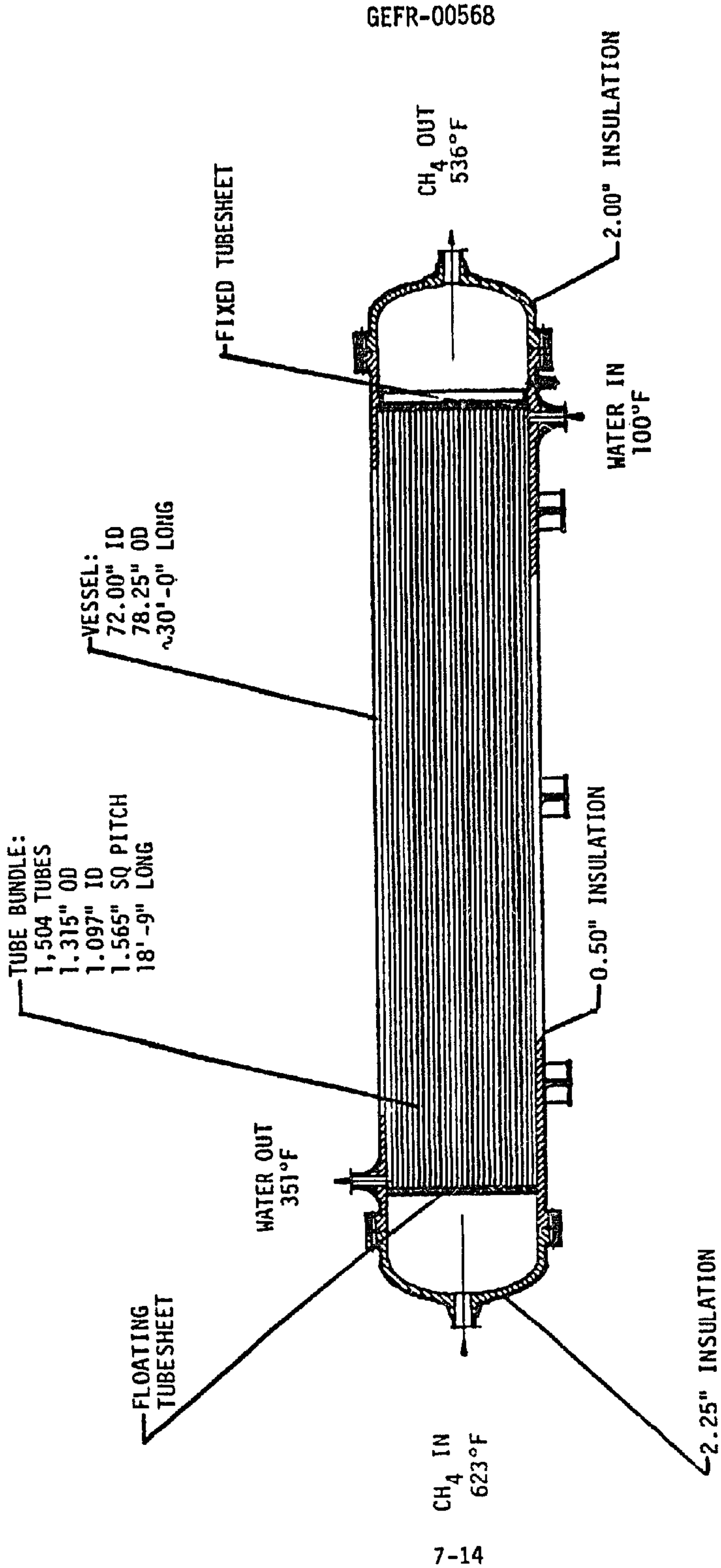

FIGURE 7.10. Feedwater Heater No. 1 180 MW Plant

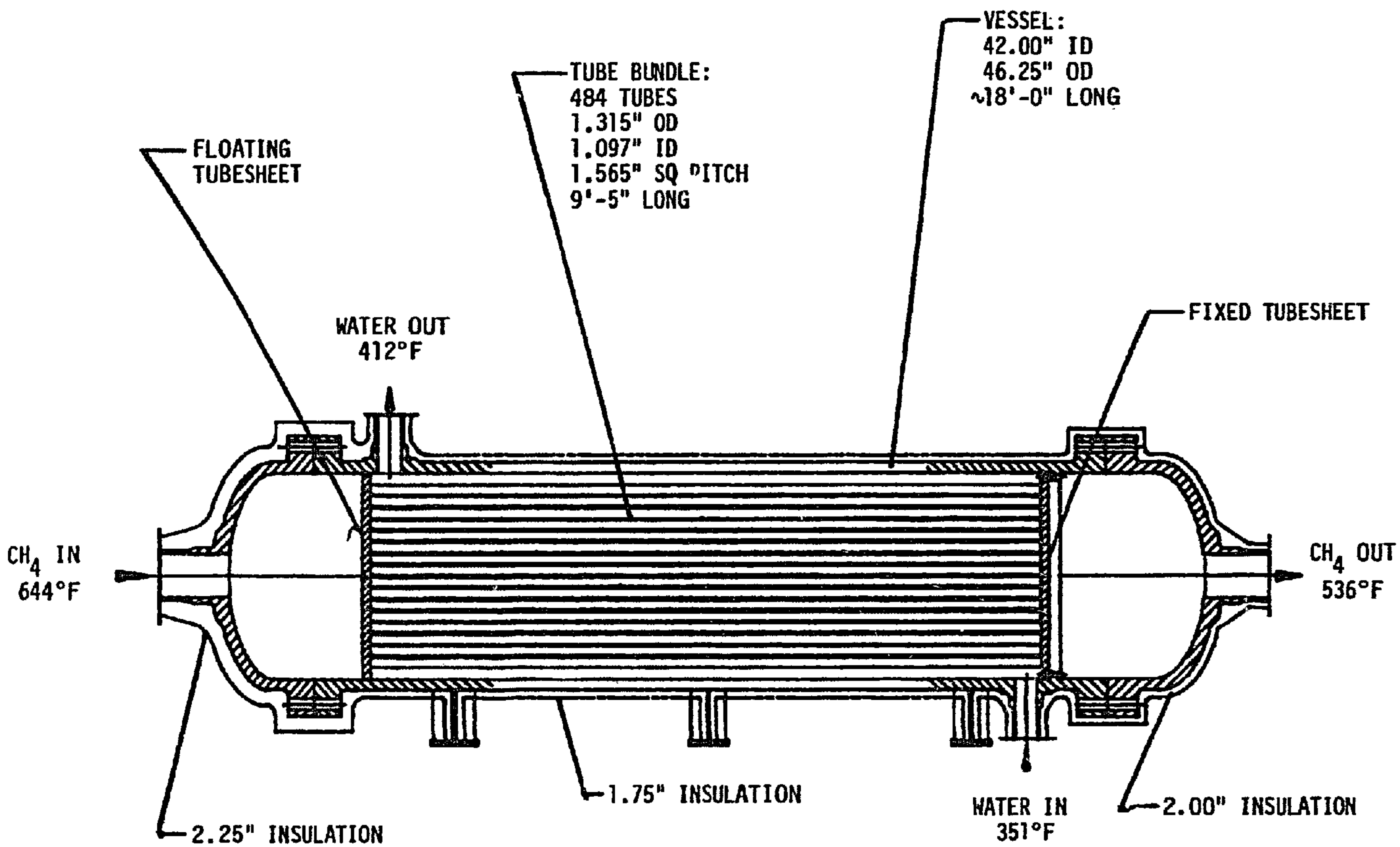

FIGURE 7.11. Feedwater Heater No. 2 180 MW Plant

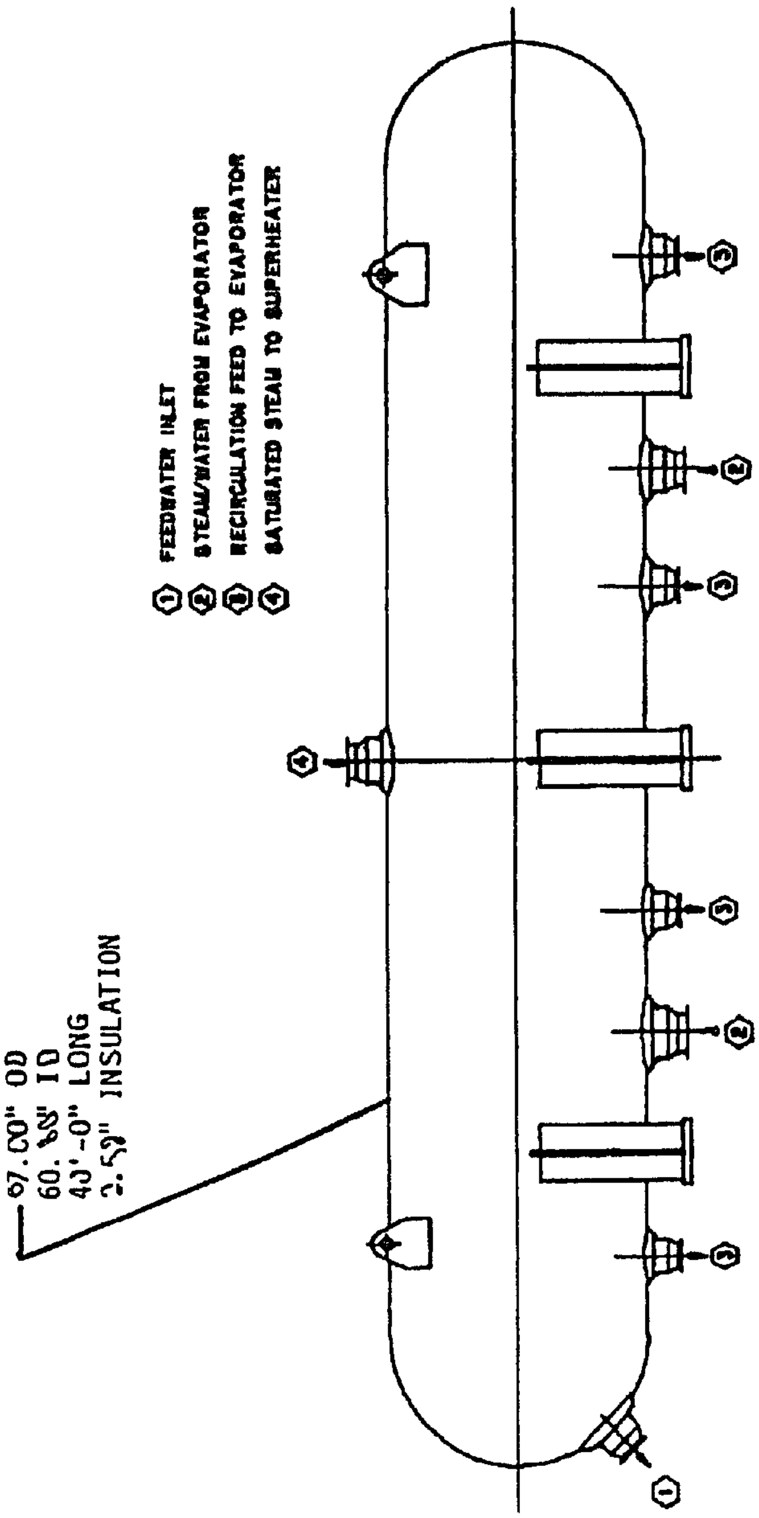

FIGURE 7.12. Steam Drum 180 MW Plant

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preheater the gas stream is split (50/50) and passed through two adiabatic methanators, located in parallel. The gas is then reformed into a single stream and flows through a feedwater heater. Upon leaving the feedwater heater the gas stream is split (50/50) and passed through the last two adiabatic methanators, located in parallel. Again the gas is reformed into a single stream and passed through a feedwater heater and the first syngas preheater, located in series. After leaving the syngas preheater, the gas (now 51% methane and 44% water vapor) is routed through a knock-out drum and dryer to remove the moisture and then to a pipeline for return to the reforming plant. The gas returned is 91.4% methane at a temperature of 200°F and 1,240 psi.

7.1.2 Water to Superheated Steam Conversion

Feedwater is a combination of condensate with a temperature above 100°F and demineralized make-up water with a temperature less than 100°F. It is assumed that the initial water temperature is 100°F at ambient pressure. The feedwater pump increases the water pressure to 1280 psi, after which the water flows through two water heaters and then to a steam drum at 412°F. In the steam drum the water mixes with a small amount of superheated steam and saturated steam to stablilize at a steam/water mixture at 574°F. Water at 574°F is pumped from the steam drum through the combination methanator/evaporators and back to the steam drum. Saturated steam flows from the steam drum through two superheaters and then is delivered to the coal gasification system at 960°F and 1250 psi.

7.1.3 Nozzle and Pipe Sizes

The quantity of syngas received from the pipeline is 325,000 lbm/hr (81,250 lbm/hr/train). The quantity of superheated steam produced is 437,000 lbm/hr (109,250 lbm/hr/train). The required nozzle and pipe sizes for the syngas and methane flow through the methanation system are given in Table 7.2. The nominal velocity is 10,000 FPM taking into consideration temperature and pressure. The required nozzle and pipe size for the water and steam flow are given in Table 7.3. The nominal velocities are:

Water - 400 FPM(1)
High Pressure Sat. Steam - 8,000 FPM(1)
High Pressure S.H. Steam - 12,500 FMP(1)

7.2 Oil Shale Reduction

Conceptual design of methanation systems were completed for coupling to the PARAHO and TOSCO II oil shale reduction processes. In both designs, only adiabatic methanators are used since there is no water. Methanators 1 and 2 of both systems use internal ceramic insulation since the outlet process-gas temperatures are greater than 1500°F.

7.2.1 Methanation Plant for PARAHO

The PARAHO methanation plant receives dry syngas from the HTGR-R plant at the rate of 1.95×10^5 lbm per hour, at a temperature of 60°F and a pressure of 400 psia. This is for a single methanation train rated at 110 MW_t. The plant also receives retort-gas from the retort plant in two streams; one - flows at the rate of 1.54×10^5 lbm per hour and the second at 3.27×10^5 lbm per hour. Both streams enter the methanation plant at a temperature of 216°F and a pressure of 10 psig. The 1.54×10^5 lb/hr stream is returned to the retort plant at a temperature of 1170°F and a pressure of 8 psig and the 3.27×10^5 lb/hr stream is returned at 1300°F and 8 psig.

The PARAHO methanation system uses four adiabatic methanators, two desulfurizers and eight gas-to-gas heat exchangers. The methanation train

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TABLE 7.2

Syngas — Methane Nozzle and Pipe Sizes

Component	Inlet	Outlet
Syngas Heater #3	6" - Schedule 40	6" - Schedule 40
Syngas Heater #2	6" - Schedule 40	8" - Schedule 80
Syngas Heater #1	8" - Schedule 80	10" - Schedule 80
Desulfurizer	10" - Schedule 80	10" - Schedule 80
Combination Methanator/ Evaporator	6" - Schedule 80	8" - Schedule 80
Superheater #2	12" - Schedule 100	12" - Schedule 100
Superheater #1	12" - Schedule 80	12" - Schedule 80
Syngas Heater #1	12" - Schedule 80	10" - Schedule 80
Syngas Heater #2	10" - Schedule 80	10" - Schedule 80
Methanator #4 & 5	6" - Schedule 40	5" - Schedule 40
Feedwater Heater #2	8" - Schedule 80	8" - Schedule 80
Methanator #6 & 7	5" - Schedule 40	5" - Schedule 40
Feedwater Heater #1	8" - Schedule 80	6" - Schedule 40
Syngas Heater #3	6" - Schedule 40	6" - Schedule 40
K. O. Drum, Dryer and Pipeline	6" - Schedule 40	6" - Schedule 40

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TABLE 7.3

Water — Steam Nozzle and Pipe Sizes

Component	Inlet	Outlet
Feedwater Pump	8" - Schedule 20	8" - Schedule 40
Feedwater Heater #1	8" - Schedule 40	10" - Schedule 80
Feedwater Heater #2	10" - Schedule 80	10" - Schedule 80
Recycle Pump	10" - Schedule 80	10" - Schedule 80
Combination Methanator/ Evaporator	6" - Schedule 80	6" - Schedule 80
Superheater #1	8" - Schedule 80	8" - Schedule 80
Superheater #2	8" - Schedule 80	10" - Schedule 80
Steam Drum	(2) 10" - Schedule 80	(1) 10" - Schedule 80 (1) 8" - Schedule 80

is described in Figure 7.13 and the heat exchangers are described in Table 7.4. This figure and table are reproduced from reference (2).

7.2.2 Methanation Plant for TOSCO II

The TOSCO II methanation plant receives dry syngas from the HTGR-R plant at the rate of 1.28 x 10^5 lbm per hour, at a temperature of 100°F and a pressure of 610 psia. This is for a single methanation train rated at 80 MW_t. The plant receives retort-gas from the retort plant at a temperature of 550°F and a pressure of 65 psig. The required retort-gas temperature upon return to the retort plant is 1430°F with a pressure of 62 psig. To achieve this temperature the methanator effluent at the outlet of this first methanator must be about 1665°F and at the outlet of the second methanator the effluent temperature needs to be about 1481°F.

The TOSCO II methanation system uses four adiabatic methanators, two desulfurizers, ten gas-to-gas heat exchangers and two air coolers. The methanation train is described in Figure 7-14.

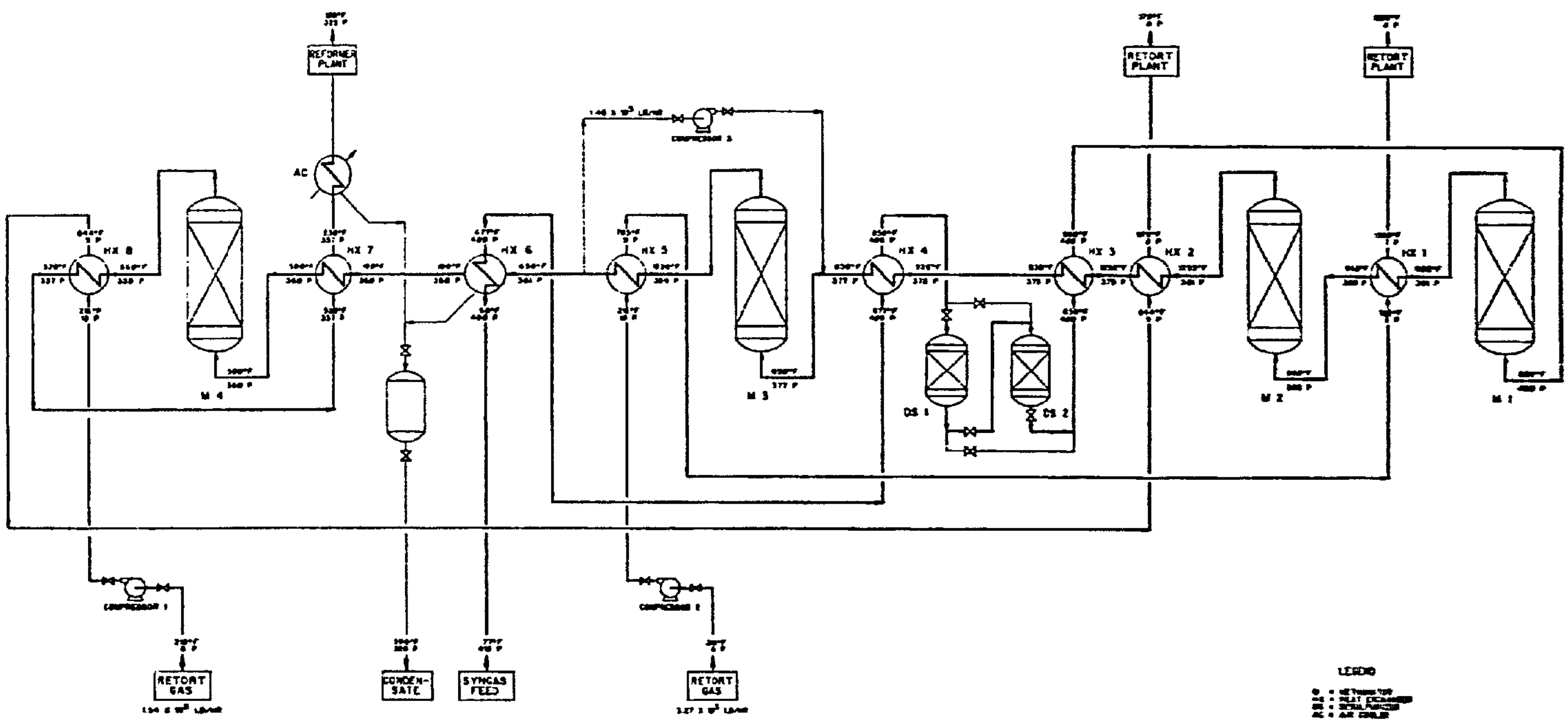

FIGURE 7.13. Flow Sheet Methanation Plant PARAHO

TABLE 7.4

PARAHO Heat Exchange Summary Sheet

	HEAT EXCHANGER NUMBER	1	2	3	4	5	6	7	8
	ENERGY EXCHANGE, MWt	30.2	14	9.6	7.7	27.4	18.2	6.8	9.4
	TYPE*	FT-1	FT-1	FT-1	FT-1	FT-1	FT-1	FT-2	FT-1
	TUBE BUNDLE LENGTH, FT	25	16	12	12	20	18	25	12
	TUBE BUNDLE WIDTH, FT	5	3.5	3.2	2.8	4.5	3.5	5.5	3.5
	TUBE BUNDLE HEIGHT, FT	5.4	2.2	2.9	2.7	2.5	3.6	5.8	2.8
	NUMBER OF TUBES	1510	277	346	274	420	684	1819	520
	TUBE OUTSIDE DIA., INCHES	0.75	1.0	1.0	1.0	0.75	0.75	0.75	0.75
	TUBE WALL THICKNESS, INCHES	0.06	0.06	0.04	0.04	0.04	0.04	0.04	0.04
	TUBE MATERIAL	Alloy 800H	Alloy 800H	304 SST	2-1/4Cr-1Mo	310 SST	2-1/4Cr-1Mo	2-1/4Cr-1Mo	310 SST
	FIN HEIGHT	0.375	0.375	0.375	0.375	0.375	0.375	0.375	0.375
	NO. OF FINS PER INCH	10	10	10	10	10	10	10	10
TUBE SIDE	FLOW, $X10^5$ LB/HR	1.95	1.95	1.95	1.95	3.41	1.95	1.19	1.19
TUBE SIDE	INLET TEMP. °F	1502	1355	1056	836	1036	60	190	860
TUBE SIDE	OUTLET TEMP. °F	860	1056	836	650	650	477	500	520
TUBE SIDE	INLET PRESSURE, PSI	391	381	379	378	364	400	360	358
TUBE SIDE	ΔP, PSI	1.3	2.5	1.0	1.5	2.8	1	0.8	0.7
SHELL SIDE	FLOW, $X10^5$ LB/HR	3.27	1.54	1.95	1.95	3.27	1.95	1.19	1.54
SHELL SIDE	INLET TEMP. °F	785	644	650	477	216	650	520	216
SHELL SIDE	OUTLET TEMP, °F	1300	1170	860	650	785	190	230	644
SHELL SIDE	INLET PRESSURE, PSI	9	9	400	400	10	361	357	10
SHELL SIDE	ΔP, PSI	2.2	0.8	0.1	0.1	1.0	Neg	Neg	0.9

*FT-1 = FINNED TUBE, SINGLE TUBE PASS
FT-2 = FINNED TUBE, DOUBLE TUBE PASS

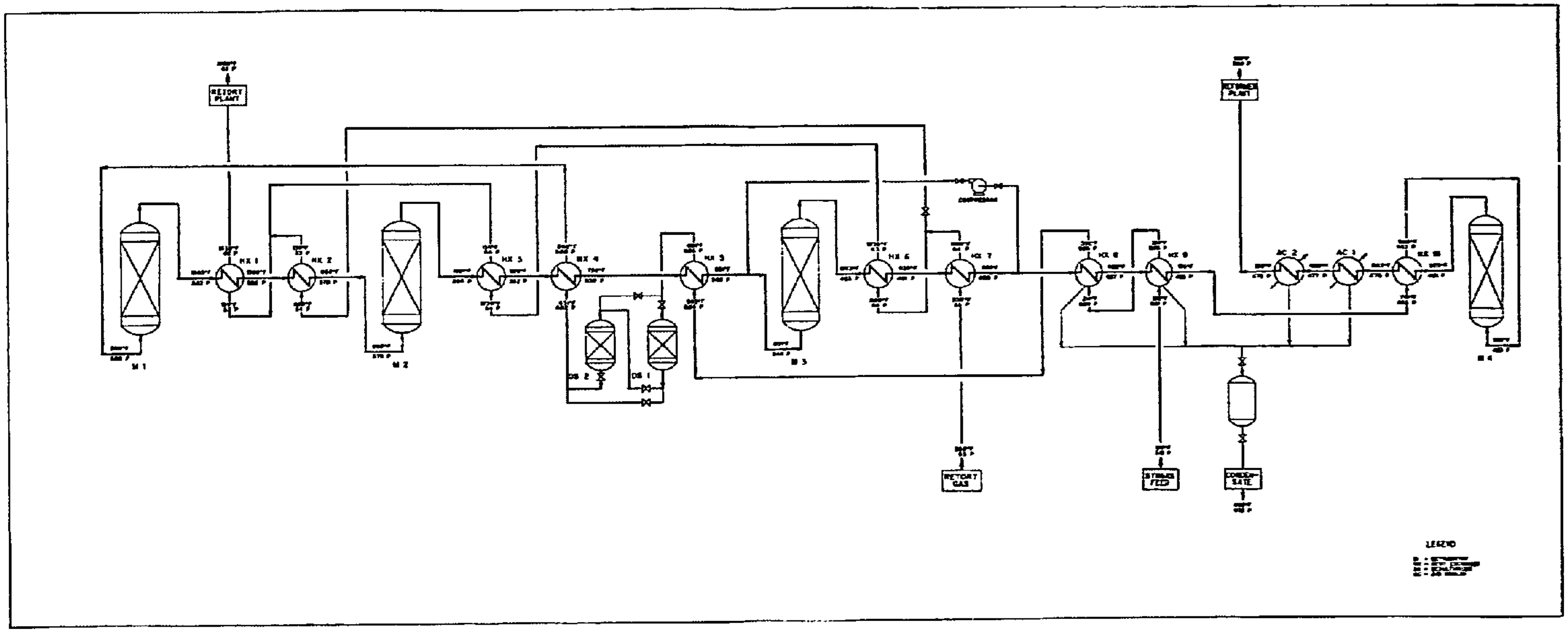

FIGURE 7.14. Flow Sheet, Methanation Plant — TOSCO II

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SECTION 7 - REFERENCES

1. Power Plant Engineering, Third Edition, page 603, by Frederick T. Morse, M.E., D. Van Nostrand Company, Inc., New York.

2. Unnumbered Report, "Conceptual System Design Description of Heat Transfer System Coupling Oil Shale Processes to the HTGR," by T. E. Gleason, dated August 1981, General Electric Company, ARSD, Sunnyvale, California.

8. ADVANCED METHANATION DESIGN CONCEPTS

In parallel with the development of the reference 40 MW_t methanation system, investigations of advanced systems were made. Some of the findings are:

8.1 Isothermal Process Reactors

Investigation[1] found that isothermal process reactors provide, at least in concept, several advantages over an adiabatic process:

a. Product recycle is either reduced or eliminated.

b. Heat exchange duties and cooling costs are lowered.

c. Avoiding low inlet temperatures and chemical-reaction-limited kinetic regimes reduces catalyst requirements.

d. Avoiding extremely high exit temperatures and hot spots prevents sintering of the catalyst, thereby preserving catalyst life.

These are the reasons for developing the Combination Methanator/Evaporator. Isothermal reactor designs have already been used in the synthesis of higher hydrocarbon and in steam reforming of methane.

8.2 Monolithic Catalyst Reactors[1]

Monolithic catalyst reactors are the most promising because of their potential for reducing recycle and catalyst costs. Monolithic catalyst can be used in adiabatic or isothermal processes. They could be manufactured as: coated ceramic monoliths, coated metal monoliths, impregnated Al_2O_3 monoliths, and extruded solid catalyst monoliths. Some considerations of monolithic catalyst are:

a. At temperatures up to 700°C (1300°F) the space velocity can be increased by a factor of 3 to 5. Operation of a monolithic bed at space velocities of up to 50,000 h^{-1} is quite feasible compared to the reference 40 MW_t methanation system which has space velocity of 10,000 h^{-1}. This would permit a corresponding reduction in reactor vessel size and wall thickness which would decrease capital costs. The volume of catalyst would be reduced but the cost savings for catalyst is yet to be determined.

b. At temperatures over 815°C (1500°F) and up to 1100°C (2000°F) which is a temperature range very desirable for synfuel processes a significant R&D effort will be necessary. In view of the obvious potential advantages of monolithic reactors relative to pellets, one would anticipate significant R&D efforts by DOE and industry to test these catalyst on a large scale. Unfortunately such an R&D effort has not materialized in either sector of the R&D community. Thus far, there have been only lab scale tests of monolithic catalysts.

8.3 Fluidized-Bed Reactors

In a recent report[2] it was shown that the rate of attrition of a Ni-Ru/Al_2O_3 catalyst was much too high for economical operation on a large scale. Accordingly the development of an attrition resistant, high temperature catalyst, such as Ni-Ru/SiO_2-Al_2O_3 would be an important prerequisite to further consideration of a fluidized bed process. In addition to being attrition-resistant, the ideal fluidized bed catalyst would one - have to be able to operate at low pressures with high selectivity to methane and two - be cheap enough to throw away. Unfortunately there are no promising candidates which meet all three specifications.

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SECTION 8 - REFERENCES

1. Calvin H. Bartholomew, "Review of Advanced Methanation Technology," Bartholomew Consulting Services, Inc., 400 North 600 East, Orem, Utah 84605, dated July 31, 1981. (Calvin H. Bartholomew is a professor at Brigham Young University and is a recognized authority on catalyst and catalytic reactors).

2. Oscarson, J. L. and Bartholomew, C. H., "Preliminary Calculations and Recommendations for the Design of a Fluidized Bed Methanation System," dated August 19, 1981.

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9. RESEARCH AND DEVELOPMENT REQUIREMENTS

The reference 40 MW_t Methanation system is believed to be a good basic design from which a more cost effective system will evolve. R&D requirements that are needed to produce a viable system in the late 1980's are those which will increase reliability and availability and decrease costs.

9.1 Immediate Engineering Activities (FY-1982 & 1983)

Immediate engineering activities are those which can be completed in FY-1982 or 1983. They would serve to prove the concept of the reference methanation system design and support and hopefully improve the present cost estimate.

9.1.1 Computer Codes

The GE METH Code needs validation and expansion. It was written for adiabatic methanators and can be validated by comparing its predictions with empirical data obtainable from various production processes. Efforts to do this was initiated in the fourth quarter, FY 1981 by subcontracting to United Catalyst, Inc. to provide the data for their catalyst for several operating conditions. The comparison of predictions to results will be made early in FY-1982. The METH Code should also be expanded for use in evaluating the isothermal process and for evaluating monolithic and fluidized-bed catalyst.

The GE HEATEX Code needs to be expanded to predict the performance and the sizing of direct contact condensers and evaporators as well as for conventional tube and shell heat exchangers.

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Using the METH Code, a parametric study of catalyst bed diameters vs bed height needs to be accomplished to determine the feasibility of decreasing vessel diameters.

Two computer codes need to be developed to apply ASME requirements to one - the design of heat exchanger tubesheets and two - the design of flanges for pressure vessels.

9.1.2 Resize Methanation System

The reference 40 MW_t methanation system was successfully scaled to 180 MW_t capacity for application to the Exxon coal gasification process and so it is assumed that it could be scaled to any desired size between 40 and 180 MW_t. However, the 180 MW_t methanation train would be one of five trains in a 720 MW_t methanation plant so that availability would be 100%. The same availability needs to be made available to the industrial steam users. If the steam requirements of a given industrial complex is 40 MW_t, the methanation plant should comprise three 20 MW_t or four 15 MW_t methanation trains. The use of three or four smaller methanation trains would not only provide 100% availability but would also make it easier to follow demand which could vary by a factor of 4 to 5 during a 24 hour day.

The 40 MW_t methanation system should be scaled-down to 15 MW_t and cost estimated. It is anticipated that a smaller size, with possible increased catalyst bed height to bed diameter ratio would result in significant cost decrease.

9.1.3 Other Design Needs

The development of a thermal insulation design for tubesheets is needed in order to improve the reliability and to prolong the life of tubesheets.

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9.2 Short Range Engineering Activities (FY 1984-1985)

Short range engineering activities are those which should be started in FY-1982 but which require more than two years to complete. This would include the planning of tests, the design of a test vehicle, fabrication of the text vehicle and the purchase of support equipment and the identification of testing facility requirements. This program is specifically needed to prove the design and to develop the controls for the isothermal process (combination methanator/evaporator).

9.3 Long Range Engineering Activities (FY 1986-1989)

Long range engineering activities are mostly the development of catalyst to increase the rate of syngas to methane conversion and/or to produce heat at temperatures above 1500°F for sustained periods (longer than one year). The development of catalyst would also include the development of a cost effective catalyst manufacturing process.

9.3.1 Monolithic Catalyst

Currently, there are no R&D activities for developing a monolithic catalyst suitable for a methanation system. However, during the last ten years considerable research and experimentation has been accomplished. Work has been accomplished for both metal and ceramic monolithic catalyst supports. Tests indicate that at sufficiently high pressure, i.e., 1000 psig, the monolithic nickel catalyst is extremely active (i.e., converts greater than 90% CO) at 3-4 times the space velocity of a conventional packed-pellet-bed reactor.[1]. It is believed that only a modest effort would be required to put one or two monolithic catalysts into production. A metal support monolithic catalyst should perform quite well in an isothermal process (catalyst in tubes with heat absorbing fluid outside the tubes) and a ceramic monolithic catalyst could replace pellet-beds in adiabatic reactors.

In both cases the volume of catalyst is reduced by a factor of 2 to 4 and hence the capital cost of pressure vessels and heat exchangers will be reduced.

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10. COST ESTIMATES

Capital cost estimate for most all major components of the methanation system for the reference 40 MW_t plant were developed by G.E. ARSD. Capital cost of purchased components for the methanation system were obtained from vendors by Bechtel. Bechtel also developed the cost estimate for the Methanator Supporting System components and the building construction, including piping, civil structure, electrical and instrumentation. To derive the cost summary, the GE and Bechtel data was combined and Bechtel added values for indirect field labor, engineering services and fee and contingency to determine the total construction cost.

The same procedure was used in determining the capital cost for the 180 MW_t Methanator Plant for Exxon Coal Gasification, and for the two 80 MW_t Methanator Plants proposed for PARAHO and TOSCO II Oil Shale Reduction. G.E. provided the cost for the major components of the methanator systems i.e., methanators, heat exchangers, steam drum, etc., and Bechtel provided the costs for all the other equipment and plant construction.

The Bechtel report is attached hereto as Appendix A-9, so the following will be only an overview of the costs for the various number of 40 MW_t methanator plants.

10.1 Capital Costs of Reference 40 MW_t Methanator Plant

The initial cost estimate for the 40 MW_t methanator plant was based on a 850°C-IDC HTGR-Reforming plant supplying 560 MW_t power per hour, continuously, in the form of syngas. This would require the construction of fourteen 40 MW_t methanator plants, all built in parallel or within the same short period of time. Each of the 40 MW_t plants would also operate at full capacity, continuously.

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A more realistic evaluation of the process steam demand by many steam users in an industrial complex is that 100% of the users operate 8 hours/day, 60-75% of the users operate 16 hours/day and 25 to 35% would operate 24 hours/day. This means that the average demand would be about 2/3rds of the methanator plant's rated capacity, or 26.7 Mw_t per hour instead of 40 MW_t per hour. To satisfy this demand twenty-one 40 MW_t methanator plants must be constructed instead of fourteen, to equal the HTGR-R plant's output of 560 MW_t per hour. In addition 1/3rd of the HTGR-R plant's syngas output per day would need to be stored. The subject and cost of syngas storage was not addressed as part of the Methanator Task.

10.1.1 Methanation System Capital Costs

An equipment list of the methanation system and the estimated cost of the equipment is given in Table 10.1. This is for one plant. For fourteen plants, the savings in production costs of the equipment would be about 15% per plant and if twenty-one plants were constructed in parallel and additional savings of 4% per plant is anticipated.

10.1.2 Support System Capital Costs

An equipment list of the support system and the estimate cost of the equipment is given in Table 10.2. This is for one plant. For fourteen plants the savings in production costs of the equipment would be about 15% per plant and if twenty-one plants were built in parallel and additional savings of 4% per plant is anticipated.

10.1.3 Plant Construction Costs

The estimated costs of plant construction (piping, civil structure, electrical and instrumentation) is given in Table 10.3. Also shown in Table

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TABLE 10.1

Equipment Cost Summary for Supporting System
(40 MWt) – Case 1

($ In 1,000's)

	Number	TOTAL*
DESULFURIZER*	2	101.5
COMBINATION METHANATOR PLANT/EVAP.*	1	620.3
ADIABATIC METHANATOR*	2	339.5
PROCESS STEAM SUPERHEATER*	1	102.6
SYNGAS HEATER*	1	93.0
FEED WATER HEATER*	3	240.1
STEAM DRUM*	1	134.4
AIR COOLER*	1	35.0
CATALYST*	-	824.8
RECYCLE GAS COMPRESSOR	3	527.4
OTHER EQUIPMENT	10***	561.4
TOTAL DIRECT COST (Price & Wage Level, 1Q 1980)	25	3,580.0

* Cost provided by G.E.
** Total direct cost including material and labor
*** Breakdown of number is as follows:

Condensate storage tank	1
K.O. Drum	1
Deaerator	1
Product Gas Dryer	2
Steam Drum Condensate Recycle Dump	2
Feed Water Pump	3
TOTAL	10

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TABLE 10.2

Equipment Cost Summary for Methanator Plant System
(40 MWt) — Case 1

($ In 1,000's)

	Number	TOTAL*
STARTUP HEATER	2	400.9
FEED WATER STARTUP HEATER	1	204.0
STARTUP NITROGEN STORAGE	1	262.5
WATER TREATMENT PLANT	1	138.0
OTHER EQUIPMENT	2**	584.6
TOTAL DIRECT COST	7	1,590.0
(Price & Wage Level, 1Q 1980)		

* Total direct cost including material and labor.

** Breakdown of number is as follows:

K.O. Pot with Flare Stack	1
Emergency Power Supply System	1
TOTAL	2

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TABLE 10.3

Construction Cost Per Plant 40 MWt Methanation System

($ In 1,000's)

Item	NUMBER OF PLANTS 1	14	21
METHANATION SYSTEM EQUIPMENT	3,580	2,910	2,790
SUPPORTING SYSTEM EQUIPMENT	1,590	1,290	1,240
PIPING	1,550	1,490	1,430
CIVIL STRUCTURAL	2,550	2,550	2,550
ELECTRICAL	1,230	1,230	1,230
INSTRUMENTATION	1,030	1,010	990
TOTAL DIRECT COST	11,530	10,480	10,230
INDIRECT FIELD COST	2,470	2,280	2,190
ENGINEERING & FEE	1,700	890	570
CONTINGENCY	3,100	2,350	2,210
TOTAL COST	18,800	16,000	15,200

(Price & Wage Level 1Q 1980)

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10.3 are the costs for indirect labor, engineering services and contingency for one, fourteen and twenty-one plants on a per plant basis.

10.2 Operating Costs for Reference Methanator Plant

The operating cost estimate of the 40 MW_t methanator plant is an update of the operating cost of a process heat plant developed by RMP.[1] Costs for electricity, catalyst replacement and maintenance are updated whereas the cost of operating labor and overhead are the same. See Table 10.4.

10.3 Plant Costs for Synfuel Applications

Using the reference 40 MW_t methanator plant cost estimate as a base, costs of three larger methanator plants were determined by scaling. In all three cases the cost of the major components of the methanation were calculated by GE-ARSD and the cost of the supporting systems and plant construction were determined by Bechtel. In all three cases the methanation plants are dedicated to support a given synfuel process at a single site, so economics of mass production are very limited or non-existent.

10.3.1 180 MW_t Methanator System for Exxon Coal Gasification Process

This system is described in Section 7, above. The estimated cost for this methanation system is scaled from the 40 MW_t plant by applying a 0.75 power factor to the change in plant capacity and assuming the design scope is a larger 180 MW_t plant. See Table 10.5.

10.3.2 80 MW_t Methanator System for PARAHO Oil Shale Reduction Process

See Table 10.5.

TABLE 10.4

Operating Costs 40 MWt Methanation Plant

($/yr)

DIRECT COSTS	COSTS
1. ELECTRIC POWER ($0.04/KWH)	$261,000
2. MAKE-UP WATER ($0.40/1,000 Gal.)	7,000
3. CATALYST REPLACEMENT:	
DESULFURIZERS	20,200
METHANATOR NO. 1	87,200
METHANATORS NO. 2 & 3	116,300
4. OPERATING LABOR (SUPT. + 8 OPERATORS)	231,000
5. MAINTENANCE LABOR (3% CONST. COST)	345,900
INDIRECT COSTS	
6. ADMINISTRATIVE AND SUPPORT LABOR (30% of 4 & 5)	173,100
7. GENERAL AND ADMINISTRATIVE EXPENSE (60% of 4 & 5)	346,200
8. PROPERTY TAXES AND INSURANCE (2.5% of Total Plant Investment)	400,000*
TOTAL OPERATING COST	$1,987,900

*Investment for 1 of 14 Plants

TABLE 10.5

Methanation Systems for Synfuel Process Support

($ In 1,000's)

APPLICATION CAPACITY (MW_t)	80	80	180
METHANATION SYSTEM EQUIPMENT	4,870	6,020	11,080
SUPPORT SYSTEM EQUIPMENT	2,160	2,670	4,920
PIPING	2,600	2,600	4,800
CIVIL STRUCTURAL	4,300	4,300	7,900
ELECTRIC	2,070	2,070	3,800
INSTRUMENTATION	1,730	1,730	3,200
TOTAL DIRECT COST	17,730	19,390	35,700
INDIRECT FIELD COST	3,790	4,150	7,600
ENGINEERING & FEE	2,580	2,860	5,200
CONTINGENCY	4,800	5,200	9,700
TOTAL CONSTRUCTION COST (Price & Wage Level, 1Q 1980)	28,900	31,600	58,200

10.3.3 80 MW_t Methanator System for TOSCO II Oil Shale Reduction Process

See Table 10.5.

SECTION 10 - REFERENCES

(1) "Design of Methanation Plants for Peaking Power Generation and Process Heat", The Ralph M. Parsons Company, Job No. 6065-1, dated June 1980.

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B. A. Hutchins
September 1, 1981

APPENDIX A1

MARKETING INFLUENCE ON BASIS OF DESIGN FOR THE 40 MWt METHANATION PLANT

Estimates were made of the size and characteristics of the industrial process heat market and how this market can affect the characteristics of methanators serving it. The potential industrial market will be composed of thousands of individual plants, ranging from zero to well over 100 MWt of process heat use and geographically distributed throughout the United States. Individual methanators may be dedicated to a single plant or centrally located so as to serve a number of plants within a radius of about 5 miles, the maximum practical distance for transporting steam or direct heat. The local concentration of users and the temperature and pressure requirements of their process heat will determine the sizes and output specifications of the methanators that serve them.

Data on currently operating industrial plants and their locations were used to estimate the ranges in product temperature and output power that will be needed from methanators to satisfy future markets. These ranges were primarily based on results of analyses performed in 1980 by General Energy Associates using their industrial data file IPEP.(1) With over 300,000 plants as a basis, GEA accumulated steam and direct heat demand rates and temperature requirements over specified regions and the total nation. Additional data for these estimates were taken from a survey of large plants performed by ORNL(2) and from studies of the Orange, Texas and Geismar, Louisiana, large industrial sites.(3)

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Figure 1 shows GEA national results for the temperature distribution of process heat, with more than 75 percent of the total at temperatures below 1000°F. The data also indicated that steam was predominantly at the lower temperatures and direct heat in the higher range. Specific large plants were seen to require steam as high as 900°F and 950 psig. Methanator design conditions of 950°F and 950 psig for output steam were chosen as a result of these conditions.

Estimates of the sensitivity of the geographically available steam market with methanator size were made from GEA results for counties within utility service areas. The first step in this estimate was to obtain an approximate distribution of industrial plants in the U.S. as a function of steam demand. Combining GEA results with ORNL results led to the distribution shown in Table 1. This estimated distribution was then superimposed on the geographical distribution of plants for the Public Service Electric and Gas Co., of New Jersey, which appears to be quite representative of a large number of utility service areas in the U.S. An estimate was then made of the steam demand, in general, that might be serviced by one methanator; that is, the average demand within a circle with a 5-mile radius.

Table 2 shows the results when methanators ranging from 1 to 100 MWt steam output are considered. As methanators grow larger, there are more areas in which the demand within 5 miles is too low for the given methanator size, and the available market drops off. This approximate analysis indicates that the potential market that could be served should not be very sensitive to methanator size up to 20 MWt. Even for methanators of 30, 50 and 100 MWt size, the geographic distribution of user plants would reduce the market served by only 10, 15 and 50 percent, respectively. This is largely due to the geographical density of plants increasing approximately as the inverse of the size for plants with steam capacities less than 20 MWt. The numbers of methanators of the given size needed to serve 8 Quads/yr of industrial steam are also given in Table 2.

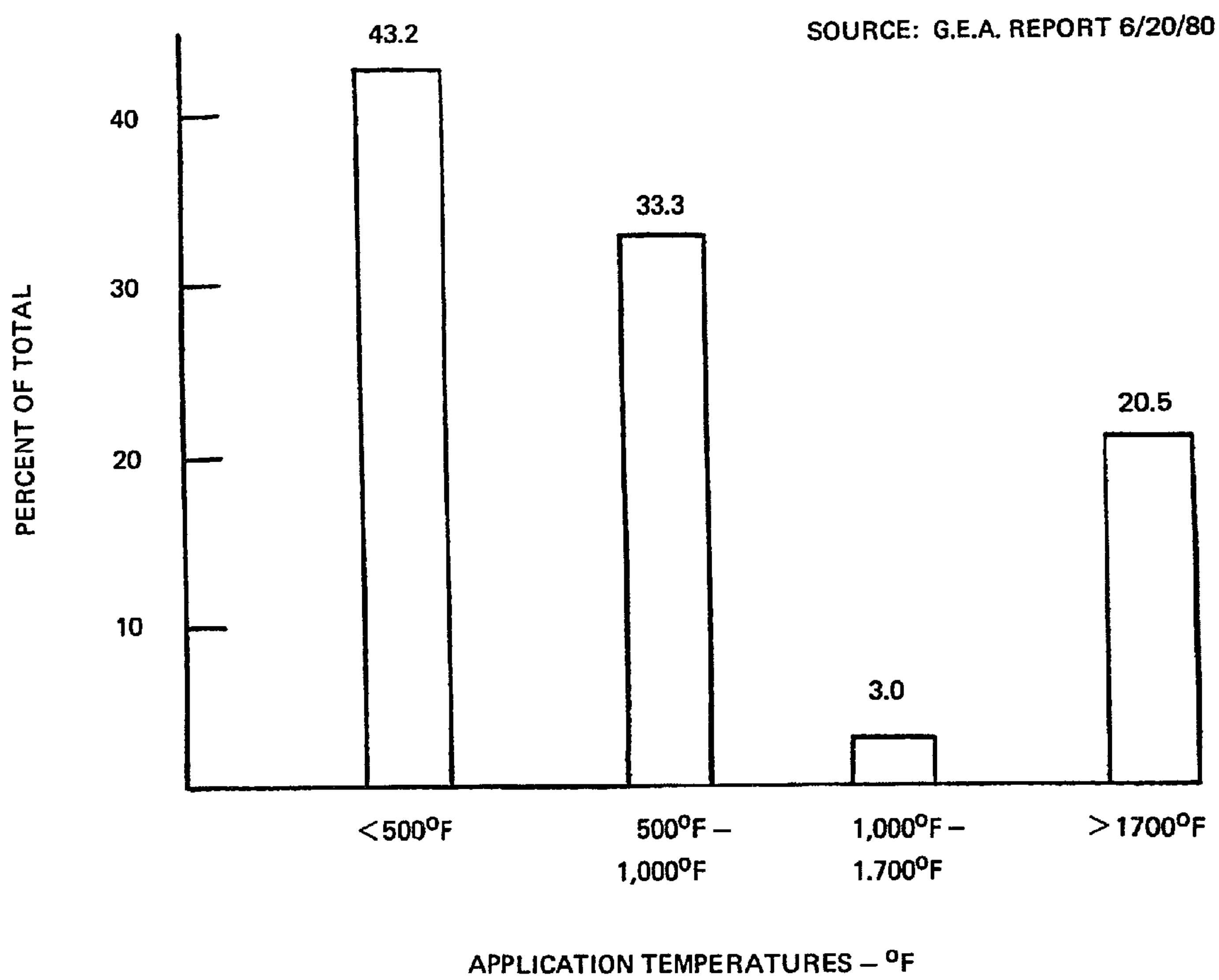

FIGURE 1. U.S. PROCESS HEAT CONSUMPTION DISTRIBUTION BY APPLICATION TEMPERATURE

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TABLE 1

DISTRIBUTION OF INDUSTRIAL PLANTS BY PLANT STEAM USE

Plant Size (MWt Steam Capacity) Range	Plant Size (MWt Steam Capacity) Average	Plant Capacity Factor	No. Plants From GEA	No. Plants From ORNL	Adjusted No. Of Plants	Total Steam Requirements (Ave. MWt)	Total Steam Requirements Quads/Yr
0-0.5	0.25	.35	286,100	-	286,100	25,000	.75
0.5-1.0	0.75	.35	21,100	-	21,100	5,540	.17
1.0-2.5	1.75	.47	15,100	-	15,100	12,400	.37
2.5-5	3.75	.54	5,590	-	5,590	11,300	.34
5-10	7.5	.69	3,450	-	3,450	17,900	.54
10-17.5	13.75	.76	1,490	-	1,490	15,600	.47
17.5-25	21.25	.76	708	-	708	11,400	.34
25-50	37.5	.80	1,060	-	1,060	31,800	.95
50-133	91.5	.80		23	158	11,600	.35
133-176	154.5	.80		61	248	30,600	.92
176-293	234.5	.80		75	202	37,900	1.13
293-410	351.5	.80	779	54	97	27,300	.82
110-586	498	.80		35	44	17,500	.52
586-820	703	.80		18	18	10,100	.30
820	900	.80		12	12	8,640	.26
TOTAL				278	335,400	274,580	8.23

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TABLE 2

SENSITIVITY OF POTENTIAL STEAM MARKET TO METHANATOR SIZE

	Size Of Methanators (MWt Steam)						
	1	5	10	20	30	50	100
Fraction Of Market Available	1	.99	.99	.95	.90	.85	.51
No. Of Methanators For 8 Quads/Yr Steam Demand (1000's)	360	71	36	17.1	10.8	6.1	1.8

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This approximate analysis indicated that the choice of methanator size for design studies should be based primarily on capital costs and could be as large as 50 MWt without incurring a significant penalty in the potential market available. Results being generated during 1981 by GEA, and a more systematic approach to estimating markets will provide a more accurate evaluation of this sensitivity.

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References:

1. Bernard B. Hamel, Harry L. Brown, Bruce A. Hedman, "HTR Multiplex Market Assessment", Revised, General Energy Associates, Inc., November 30 (1980).

2. "The Potential Industrial Market For Process Heat From Nuclear Reactors", ORNL-TM-5516, Oak Ridge National Laboratory, Oak Ridge, Tennessee, July (1976).

3. "Cogeneration Feasibility Study in the Gulf States Utilities Service Area", ORNL/SUB-7317/1, Oak Ridge National Laboratory, Oak Ridge, Tennessee, July (1970).

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YL-221-10090
N. T. Arcilla
August 26, 1981

APPENDIX A2

DESIGN BASIS
40 MWt METHANATION PROCESS

In Figure 1 are the process specifications recommended for equipment design calculations for a 40 MW_t Methanation Plant producing process steam at 900 psia and 900°F.

The design requirements that strongly influenced selection of key process features were:

1. Maximum methanator effluent temperature of 593°C (1100°F),
2. Final methane yield of better than 90%,
3. Maximum of three (3) methanation steps.

The maximum effluent temperature requirement of 593°C largely applies to the first methanator where high process temperatures are expected due to the high percentages of CO and CO_2 (19.4 mole % combined) in the syngas feed. In a simple adiabatic methanator (without provision for in-reactor temperature control), this feed will produce enough heat to raise the process temperature some 450°C above the inlet temperature. In this process, with the inlet at 344°C, an active means of controlling the process temperature in the first methanator would be needed in order not to exceed 593°C.

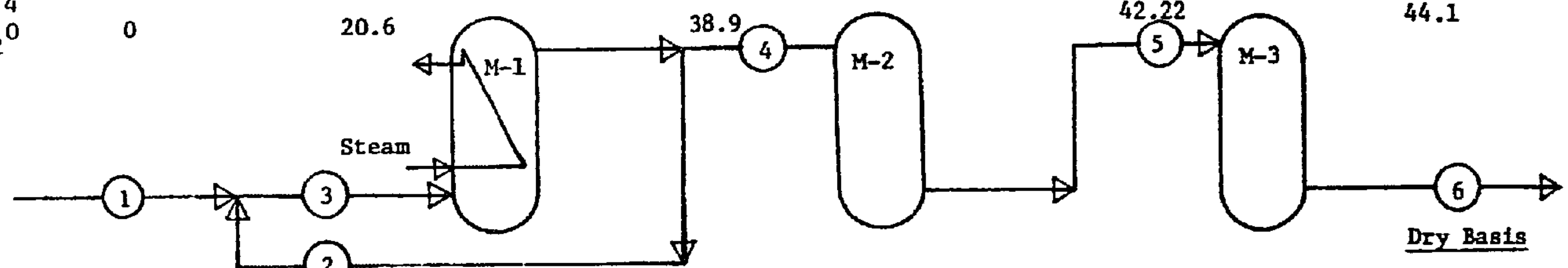
In this process, temperature control in Methanator No. 1 is achieved by internal cooling and product recycle. This methanator will be a tubes-in-shell design. About 56% of the heat generated by the reaction will be removed by the water circulating around the tubes that are packed with the catalyst. At the same time, the syngas feed will be diluted by recycling 63% (mass) of the Methanator No. 1 effluent into the Methanator No. 1 feed stream.

Ninety percent (90%) conversion is achieved in the first methanator. Two (2) additional methanators are required to bring the conversion up to 98% with a methane yield of 91.4%. Methanators No. 2 and 3 will be conventional packed bed designs.

Figure 1. Design Basis, 4C $_t$ Methanation Process

M-1 Internally-Cooled w/Recycle, 56% of Heat in M-1 to Steam
Weight Recycle Ratio = 1.73, Process Pressure = 900 psia

Flow	9.090 kg/sec	24.815 kg/sec	9.090 kg/sec	9.090 kg/sec	9.090 kg/sec
CO	9.9 mole %	4.8 mole %	0.2 mole %	0.01 mole %	0 mole %
CO_2	9.5	5.9	2.7	1.69	0.7
H_2	67.7	37.9	11.4	6.80	4.1
CH_4	12.9	30.8	46.8	49.28	51.1
H_2O	0	20.6	38.9	42.22	44.1



15.725 kg/sec

Dry Basis

CO	0 mole %
CO_2	1.3
H_2	7.3
CH_4	91.4

Heat Produced ,	MW_t	38.6	2.0	1.5
Heat to Steam ,	MW_t	21.6	-	-
$T_{(in)}/T_{(out)}$,	°C	344/546	280/336	280/324
Total Conversion ,	% of Feed	90	94	98
CH_4 in Dry Prod. ,	Mole %	76.6	85.2	91.4
Catalyst Vol. ,	Ft^3	530	500	500
Catalyst		UCI C150-5	UCI C150-4	UCI C150-4
Estimated Average Pressure Drop ,	Psi/Ft	1.0	1.0	1.0
Methanator Design ,		Internally cooled with recycle. Catalyst in tubes Sat'd. water and steam in shell. Gas and coolant in co-current, upward flow.	Adiabatic, Packed Bed	Adiabatic, Packed Bed

p

YL-611-10063
Chi-Kou Fan
May 21, 1981

APPENDIX A3

DESIGN BASIS METHANATION PLANT HEAT TRANSPORT SYSTEM DESIGN

REFERENCES: See Attachment (2)

1.0 SUMMARY

The purpose of this memo is to present the design of the heat transport system for the design basis methanation plant, which generates a total of 42 MW power by chemical reactions, Reference 1. The generated heat will be used to heat up 60°F water to 900°F superheated steam at a pressure of 900 psia, and at a flow rate of 1.0×10^5 lbm/hr, for the process steam users.

The heat transport system design is shown in Figure 1. The size, inlet/outlet fluid temperatures, and the energy removed by each heat exchanger are calculated and listed in Tables 1 and 2.

2.0 DESIGN DESCRIPTION

Figure 1 shows the design of the heat transport system. It consists of a steam drum, a superheater, an evaporator, a syngas preheater, and four feed water preheaters. The system will be operated at a pressure of 900 psia.

The design basis methanation plant contains three methanators, and generates a total of 42 MW thermal power (38.6, 2.0, and 1.5 MW for the No. 1, 2, and 3 methanator respectively).

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2.1 Water/Steam Loop

The feed water comes into the plant at a temperature of 60°F, and a flow rate of 1.0×10^5 lbm/hr. It is heated up to 475°F through the four feed water preheaters and then delivered to the steam drum. The incoming subcooled water mixes with the steam in the steam drum and becomes saturated fluid (532°F). It is then pumped out to the evaporator, that is located inside the first methanator. About 60% of the energy (21.6 MW) generated there will be extracted by evaporating water to saturated steam. The effluent from the evaporator is a mixture of water and steam, and it is then delivered back to the steam drum for separation. The steam coming out of the steam drum will be heated to 900°F by the superheater, for the process steam users.

2.2 Gas Loop

The syngas comes into the plant at a flow rate of 7.2×10^4 lbm/hr, and a temperature of 100°F. It is heated at 651°F through the syngas preheater to feed into the No. 1 methanator. The gas coming out of the methanator is 1015°F. It is then cooled down to 651°F by a superheater and a syngas preheater. About 60% of the gas (1.25×10^5 lbm/hr) coming out of the syngas preheater will be recycled to the No. 1 methanator, the rest will be cooled down to 536°F by the No. 4 water preheater to feed into the No. 2 methanator. The gas comes out at 637°F, and is then cooled down to 536°F by the No. 3 water preheater to feed into the third methanator. The product gas coming out of the third methanator is 615°F. It is cooled down to 100°F by the No. 1 & 2 water preheaters for delivery to the reformer plant.

3.0 HEAT EXCHANGER (HX) CALCULATIONS

The syngas preheater, steam superheater, and water preheaters are all tube-shell type heat exchangers. The evaporator is also similar to a tube-shell type heat exchanger, except inside the tubes are packed with catalyst pellets for speeding up the chemical reactions. All the heat exchangers are designed to have the hot fluid flowing on the tube side, and the cold fluid flowing on the shell side.

3.1 Overall Heat Transfer Coefficient of HXs

The overall heat transfer coefficient of the HXs are calculated by, reference 2.

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$$U_o = \frac{1}{\frac{r_o}{h_i \cdot r_i} + \frac{r_o \ln(r_o/r_i)}{kw} + \frac{1}{h_o} + R_i \cdot \frac{r_o}{r_i} + R_o} \quad (1)$$

The variables in this and the following equations in this section are described in the attachment 1.

3.1.1 Evaporator:

The evaporator is located inside the first methanator. On the tube side of the evaporator, it is packed with catalyst pellets, and the syngas flows through them. The convective heat transfer coefficient for flow through a packed bed can be calculated by, References 4 and 5,

$$h_i = \frac{Kg}{D_p} [2.58(\frac{\rho_g v_s D_p}{\mu g})^{1/3} (Pr)^{1/3} + 0.094(\frac{\rho_g v_s D_p}{\mu g})^{0.8} \cdot (Pr)^{0.4}], \quad (2)$$

where D_p is the equivalent sphere diameter of the catalyst pellet and can be calculated by

$$D_p = (\frac{6 \cdot v_p}{\pi})^{1/3}$$

The superficial velocity, v_s, in equation (2) can be calculated by

$$v_s = \dot{M}g/(\rho_g \cdot A_t \cdot N).$$

On the outer surface of the evaporator tube, water is evaporating. The magnitude of the boiling heat transfer coefficient is on the order of 10^3 to 10^4 Btu/hr·ft^2·°F and is much higher than the heat transfer coefficient of the tube inner surface,

$$\frac{1}{h_o} << \frac{1}{h_i}$$

Therefore, equation (1) becomes

$$U_o = \frac{1}{\frac{r_o}{h_i \cdot r_i} + \frac{r_o \ln(r_o/r_i)}{kw} + R_i \cdot \frac{r_o}{r_i} + R_o} \quad (3)$$

ρ

3.1.2 Other HXs:

The convective heat transfer coefficient for both the inner and outer tube surfaces of the superheater, syngas preheater and water preheaters can be calculated by using the Dittus Boetler correlation, Reference 3,

$$h_{i \text{ or } o} = 0.023 \left(\frac{k}{De}\right) \left(\frac{\rho v De}{u}\right)^{0.8} (Pr)^{0.4} \tag{4}$$

The equivalent hydraulic diameter, De, in this equation for the inner and outer tube surfaces can be calculated by

$$De = Di \tag{5}$$

and

$$De = 4\ As/[\pi \cdot Dv + N \cdot \pi \cdot Do] \tag{6}$$

respectively.

3.2 Thermal Properties

To calculate the convective heat transfer coefficient in equations (2) and (4), it is necessary to know the thermal properties of the fluids. Typically, there are two kinds of fluid, one is water (or steam) and the other is a mixture of various gases. The thermal properties of water can be obtained from the tables in Reference 7 directly, while the thermal properties of a gas mixture can be calculated by using the equations developed in Reference 6,

$$kg = 0.5 \left[\sum_{i=i}^{n} X_i \cdot k_i + \left(\sum_{i=i}^{n} (X_i/k_i)\right)^{-1}\right] \tag{7}$$

$$\mu_g = \sum_{i=i}^{n} (X_i \cdot \mu_i \cdot M_i^{1/2}) / \sum_{i=i}^{n} (X_i \cdot M_i^{1/2}) \tag{8}$$

$$C_{p \cdot g} = \sum_{i=i}^{n} (C_{pi}\ X_i\ M_i) / \sum_{i=i}^{n} (X_i\ M_i) \tag{9}$$

$$\rho_g = \sum_{i=i}^{n} \rho_i \cdot X_i \tag{10}$$

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3.3 Water Flow Rate

In order to calculate the shell side fluid velocity of the superheater and water preheaters in equation (4), it is necessary to know the water flow rate. The following analysis gives the water flow rate.

As specified by the methanation plant requirements, Reference 8, the feed water comes into the plant at a temperature of 60°F, and leaves the plant at 900°F, and 900 psi pressure (superheated steam). The enthalpy of water and steam at these temperatures are

$$H_f\ (60°F) = 28.06\ \text{Btu/lbm}$$

$$H_g\ (900°F,\ 900\ \text{psi}) = 1451.8\ \text{Btu/lbm}$$

The rate of energy to be removed by water is 42 MW (143×10^6 Btu/hr). Therefore, the water flow rate is

$$\dot{M}_{H_2O} = \frac{143 \times 10^6 \quad \text{Btu/hr}}{(1451.8 - 28.06)\ \text{Btu/lbm}} = 1.0 \times 10^5\ \text{lbm/hr}$$

3.4 Energy Removal

The rate of heat removed by each heat exchanger can be calculated by

$$\dot{q} = \dot{M}_g \cdot C_{pg} [T_{pi} - T_{po}], \tag{11}$$

If the secondary fluid is water, we can also calculate the heat transfer rate by

$$\dot{q} = \dot{M}_{H_2O} \cdot [H_o - H_i]. \tag{12}$$

3.5 Sizes of HXs

The total required surface area of the tubes can be calculated by, Reference 2,

$$A = \dot{q}/U_o \cdot \Delta T_{lm} \tag{13}$$

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where ΔT_{lm} is the logarithmic mean value of the temperature difference between the primary (hot) and secondary (cold) fluids, and is defined as

$$\Delta T_{lm} = \frac{(T_{pi} - T_{so}) - (T_{po} - T_{si})}{\ln [(T_{pi} - T_{so})/(T_{po} - T_{si})]} \tag{14}$$

The length of each tube can then be obtained by

$$L = A/(N \cdot \pi \cdot D_o) \tag{15}$$

The detailed calculations for the heat exchangers are in the Engineering Workbook No. 390. Table 1 is a list of the primary and secondary fluid temperatures for all the heat exchangers. The heat transfer coefficient heat transfer rate, and the sizes of the heat exchangers are presented in Table 2.

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ATTACHMENT 1

Nomenclature

A: Total surface area of the HX tubes (ft^2)

A_s: Cross sectional area of the HX shell side (ft^2)

C_p: Specific heat (Btu/ℓbm°F)

C_{pg}: Specific heat of the gas mixture (Btu/ℓbm°F)

C_{pi}: Specific heat of the ith individual gas (Btu/ℓbm°F)

D_i: I.D. of the tube (ft)

D_o: O.D. of the tube (ft)

D_p: Equivalent sphere diameter of the catalyst pellet (ft)

D_e: Equivalent diameter of the tube or shell (ft)

D_v: I.D. of the HX vessel (ft)

H_f: Enthalpy of water (Btu/ℓbm)

H_g: Enthalpy of steam (Btu/ℓbm)

H_o: Enthalpy of water at exit of the HX (Btu/ℓbm)

H_i: Enthalpy of water at entrance of the HX (BTu/ℓbm)

h_o: Heat transfer coefficient on the outer surface of the tube (Btu/hr ft^2°F)

h_i: Heat transfer Coefficient on the inner surface of the tube (Btu/hr ft^2°F)

k: Thermal conductivity (Btu/hr ft°F)

kW: Thermal conductivity of the tube wall (Btu/hr ft°F)

kg: Thermal conductivity of the gas mixture

ki: Thermal conductivity of the individual gas

L: Length of the HX tube (ft)

$\dot{M}_g$: Gas flow rate (ℓbm/hr)

$\dot{M}_{H_2O}$: Water flow rate (ℓbm/hr)

M_i: Molecular weight of the ith individual gas (ℓbm/ℓb-mole)

N: Number of tubes in the HX

Continued...

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ATTACHMENT 1 (Continued)

n: Number of gas species in the gas mixture

P_r: Prandtl number = $C_p \cdot \mu / k$

$\dot{q}$: Heat removed by the HX (Btu/hr)

R_i: Fouling resistance on the inner surface of the tube (0.001 hr·ft²·°F/Btu)

R_o: Fouling resistance on the outer surface of the tube (0.001 hr·ft²·°F/Btu)

r_o: Outer radius of the HX tube (ft)

r_i: Inner radius of the HX tube (ft)

T_{pi}: HX primary fluid entrance temperature (°F)

T_{po}: HX primary fluid exit temperature (°F)

T_{si}: HX secondary fluid entrance temperature (°F)

T_{so}: HX secondary fluid exit temperature (°F)

U_o: Overall heat transfer coefficient of the HX (Btu/hr ft^{2}q)

V_p: Volume of the catalyst pellet (ft^3)

v: Velocity of fluid (ft/hr)

v_s: Superficial velocity of gas in the evaporator tube (ft/hr)

X_i: Mole fraction of the ith individual gas

ρ_g: Density of the gas mixture (ℓbm/ft^3)

ρ_i: Density of the ith individual gas (ℓbm/ft^3)

μ: Viscosity (ℓbm/hr·ft)

μ_g: Viscosity of the gas misture (ℓbm/hr·ft)

μ_i: Viscosity of the ith individual gas (ℓbm/hr·ft)

ΔT_{lm}: Logarithmic mean value of the temperature difference between the primary and secondary fluids of the HX (°F)

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ATTACHMENT 2

References

1) N. Arcilla, "Design Basis, Internally Cooled with Recycle Methanation Plant," April 28, 1981.

2) Rohsenow and Choi, Heat, Mass, and Momentum Transfer, Prentice Hall, Inc., 1961.

3) McAdams, W.H., Heat Transmission, 3rd, Ed., McGraw-Hill, New York, 1954

4) Cooper, K.C., "Steam-Methane Reforming and Intermediate Heat Exchanger Model (PHXSTP) for Nuclear Process Heat," Office Memorandum, Los Alamos Scientific Laboratory (Obtained from Reference 5).

5) J.A. Bond, E.C. Duderstadt, R.G. Frank and B.L. Moor, "Design of a Helium-Heated Duplex Tube Steam-Methane Reformer" General Electric Company Energy Systems and Technology Division, ESTD 76-06, May 1976.

6) Letter YL-611-10023, Chi-Kou Fan to P.T. Hughes, "Thermal Properties of Gas Mixture," March 17, 1981.

7) Vargaftik, N.B., Tables on the Thermalphysical Properties of Liquids and Gases, 2nd Edition, Hemisphere Publishing Corporation, Washington-London, 1975.

8) Letter YL-270-10081, C.R. Davis/P.T. Hughes to Distribution, "Requirements of the Methanation Plant Design," March 31, 1981.

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ATTACHMENT 3

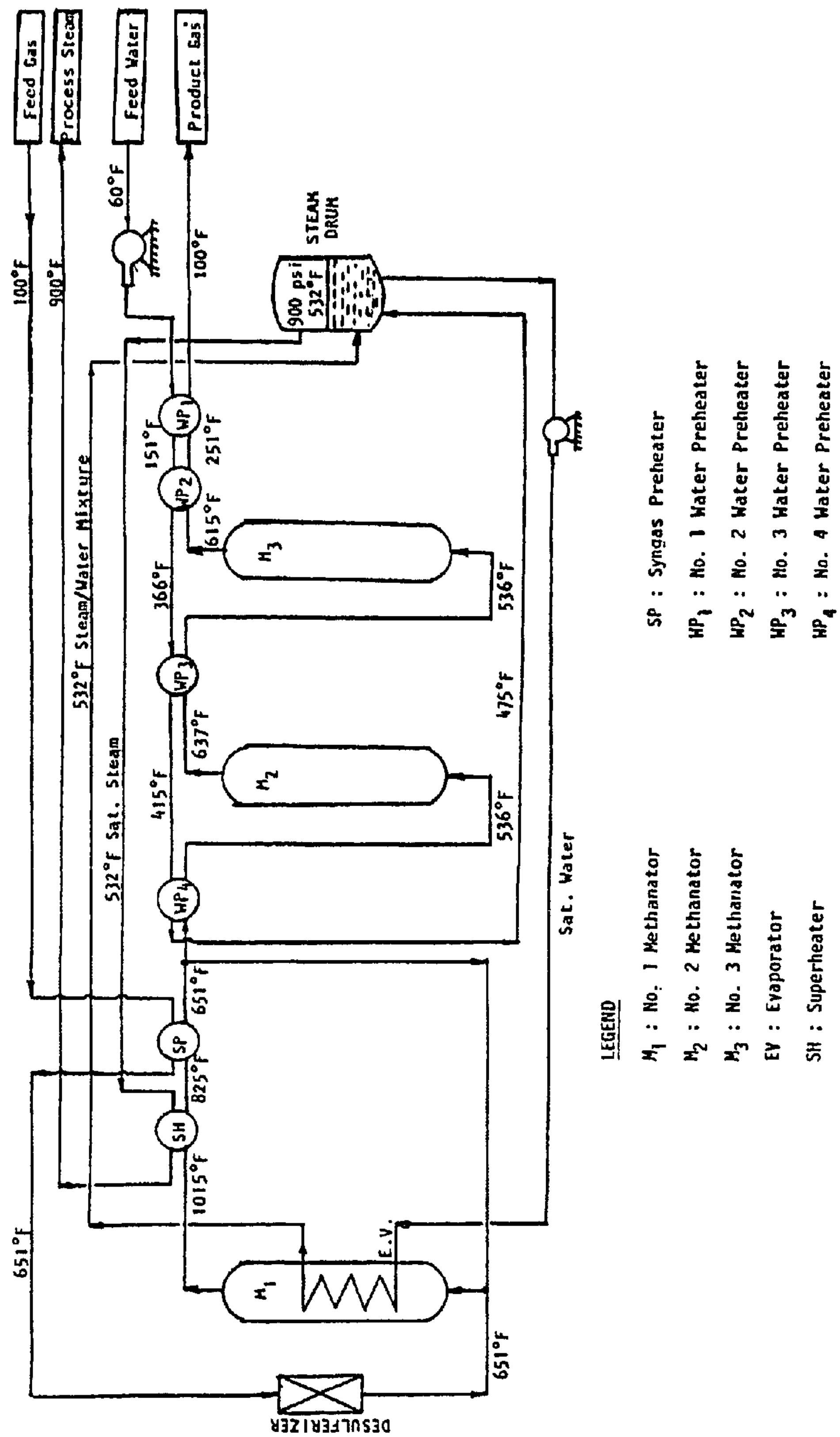

FIGURE 1. DESIGN BASIS METHANATION PLANT HEAT TRANSPORT SYSTEM DESIGN

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ATTACHMENT 4

TABLE 1

Primary and Secondary Fluid Temperatures of the HXs

HXs	Primary Fluid Temp. (°F)	Secondary Fluid Temp. (°F)
Evaporator	651-1015	532
Superheater	1015-852	532-900
Syngas Preheater	852-651	100-651
No. 1 Water Preheater	251-100	60-151
No. 2 Water Preheater	615-251	151-366
No. 3 Water Preheater	637-536	366-415
No. 4 Water Preheater	651-536	415-475

TABLE 2

Heat Exchanger Design

HXs	Overall Heat Transfer Coefficient (Btu/hr ft^2°F)	Heat Transfer Rate (MW)	Required Tube Surface Areas (ft^2)	No. of Tubes*	Tube Length (ft)	HX Vessel I.D. (ft)
Evaporator	117	21.6	2100	(Determined	by the	catalyst volume)
Superheater	31	7.6	3142	500	18.3	3.5
Syngas Preheater	37	9.2	2430	500	3.5	9.2
No. 1 Water Preheater	42	2.7	3280	500	19	3.5
No. 2 Water Preheater	42	6.5	3280	500	19	3.5
No. 3 Water Preheater	52	1.6	540	200	7.8	1.6
No. 4 Water Preheater	57	2.0	810	200	11.8	2.3

* Stainless Steel Schedule 10S 1 inch tube.

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XL-273-10110
S. G. Nagy
September 4, 1981

APPENDIX A4

AIR COOLED HEAT EXCHANGER REQUIREMENTS FOR THE 40 MWt METHANATOR PLANT

The capacity requirements, size and cost of the air cooled heat exchanger are:

Capacity:

Case I. 24,000 $\frac{lb}{hr}$ N_2 every 2-3 yrs. @ 800°F.

Case II. 40,000 $\frac{lb}{hr}$ CH_4 normal operation @ 280°F.

Heat removed is approximately $6 \times 10^6 \frac{BTU}{hr}$

Dimensions:

(5 ft wide) x (12 ft long) x (7½ to 8 ft high).

Requires two 4½ ft diameter fans, rated at 7.5 hp each.

Cost is $25,000 to $30,000.

Manufacturer: Ecodyne, MRM Division

Part No.: MW81-5211

Sales Representative in San Francisco:

Eckels Spaulding Co.
(415) 421-2662

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S. G. Nagy
August 17, 1981

APPENDIX A5

DESIGN CRITERIA FOR STARTUP REQUIREMENTS FOR THE 40 MWt METHANATOR PLANT

I. General

- all heat transfer calculations are based upon $Q = mCp\Delta T$.
- heat losses are assumed to be constant.
- heat requirements are based upon a 150°F/hr heating rate.
- Cp for methanator components assumed to be constant.
- gas-fired feedwater, syngas, and nitrogen heaters assumed to have a 65% thermal efficiency.
- ambient temperature taken at 60°F.
- vessel volumes calculated by assuming cylindrical shape.
- pipe volumes calculated by using the average pipe size and their volume to weight ratio.
- gas compressor/blower work energies considered negligible.

II. Desulfurizer Loop

- components are heated to 650°F.
- heated syngas goes from 700°F to 800°F as it passes through the startup heater (for blower sizing basis).

III. Methanator Loop

- startup heat requirements are based upon the highest demand case, i.e., catalyst reduction at 750°F.
- heated nitrogen goes from 800°F to 900°F as it passes through the startup heater (for blower sizing basis).
- all components that may be heated with water are heated to 532°F with water and then to operating temperature with heated nitrogen.

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Plant Startup Requirements

I. Startup Heater

A. Syngas heater (desulfurizer loop) $4.1 \times 10^6 \frac{BTU}{hr}$

B. Nitrogen heater (methanator loop) $13.3 \times 10^6 \frac{BTU}{hr}$

C. Feedwater heater $9.4 \times 10^6 \frac{BTU}{hr}$

II. Heated Gas Flow Rates	$\dot{m}\left(\frac{lb}{hr}\right)$	$\dot{V}\left(\frac{SCF}{hr}\right)$
A. Syngas (desulfurizer loop)	51,333	1.78×10^6
B. Nitrogen recycle (methanator loop)	492,593	6.32×10^6

III. Gas Volumes

A. N_2 17,000 ft^3 at standard conditions.

B. CH_4 147,000 ft^3 at standard conditions.

C. H_2 290,000 ft^3 at standard conditions.

Recommended Plant Capacities

I. Startup Heaters

A. Syngas heater (desulfurizer loop) $5 \times 10^6 \frac{BTU}{hr}$

B. Nitrogen heater (methanator loop) $15 \times 10^6 \frac{BTU}{hr}$

C. Feedwater heater $10 \times 10^6 \frac{BTU}{hr}$

II. Type of Gas Storage

A. N_2 - high pressure (no compressor needed)

B. CH_4 - high pressure (no compressor needed)

C. H_2 - undetermined, if greater than 900 psi no compressor needed

III. Gas Storage Volumes

A. N_2 1×10^6 SCF

B. CH_4 6.5×10^5 SCF

C. H_2 5×10^5 SCF

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YL-492-100253
G. J. Licina
June 15, 1981

APPENDIX A6

MATERIALS RECOMMENDATIONS FOR METHANATOR PLANT

Reference 1 provided materials recommendations for a 40 MW_t methanation plant with a normal operating temperature of 900°F with "hot spots" to 1020°F. A May 15, 1981, meeting with the addressees indicated that the design outlet temperature for the unit should be 1020°F with peak temperatures to 1400°F. Outline drawings of the two vessel designs and a P&ID for the plant were provided at that time. (Figures 1-4)

Materials considerations for a 1020°F methanation vessel are significantly different than for a 900°F case. Table 1 of Reference 1 illustrates this point most dramatically. At 900°F, allowables for a 2¼Cr-1Mo or 1¼Cr-½Mo steel plate are as high as those for Type 304 stainless steel; at 1020°F, however, the stainless steel has an S_o twice that of the ferritics, but water-side stress corrosion cracking considerations limit the usefulness of the stainless steel for much of the unit. For the methanator/evaporator unit shown in Figures 2 and 3, 2¼Cr-1Mo (normalized and tempered) is recommended for the vessel and nozzles, and upper and lower tubesheets. Tubesheet forgings should be clad with type 304 stainless steel on the process gas side. To facilitate tube-to-tubesheet welding, a minimum cladding thickness of .125 should be specified. A 1¼ Cr-½Mo (SA 335 Gr P11) or lower alloyed steel (e.g. SA 106B) piping should be used for water/steam piping including the feedwater train. The steam drum may utilize carbon steels (e.g. SA533B) while 2¼Cr-1Mo is recommended for feedwater heaters and the superheater.

Type 304H stainless steel is recommended for tubes in the number 1 methanator (methanator/evaporator) and for piping from the methanator to the superheater. These areas are the highest duty regions of the plant. The strength and corrosion resistance of the Type 304H will be required for these applications. The No. 2 and No. 3 methanators may be fabricated from 2¼Cr-1Mo (SA387 Gr.22 Cl2), clad with Type 321 or (preferably) 347 stainless steel for resistance to corrosion by acids of sulfur. A 1¼ Cr-½ Mo steel is recommended for process gas piping downstream of the last feedwater heater and upstream of the desulphurizers.

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Conclusions:

A 1020°F maximum temperature (vs. 900°F) strongly influences the selection of stainless steel over a low alloy Cr-Mo steel. The Cr-Mo steel should be used for steam/water containment whenever possible to preclude the possibility of through-wall stress corrosion cracking. As design temperature decreases to 900°F and below, the incentive for the use of stainless steels disappears. Alloys with a minimum of ½% molybdenum should be specified for all vessels and piping containing methanator process gases to avoid hydrogen embrittlement. The use of clad Cr-Mo plate for large vessels, a common practice in petrochemical applications [2], provides a significant cost savings on material as well as eliminating the concern over through wall SCC.

p

REFERENCES:

1. "Preliminary Materials Recommendations for Methanator Plant", letter G. J. Licina to C. R. Davis/P. T. Hughes, 4-28-81, letter No. YL-492-10236.

2. Telephone Conversations, G. J. Licina to William Erwin, Chevron Oil Co., 4-27-81, and 6-10-81.

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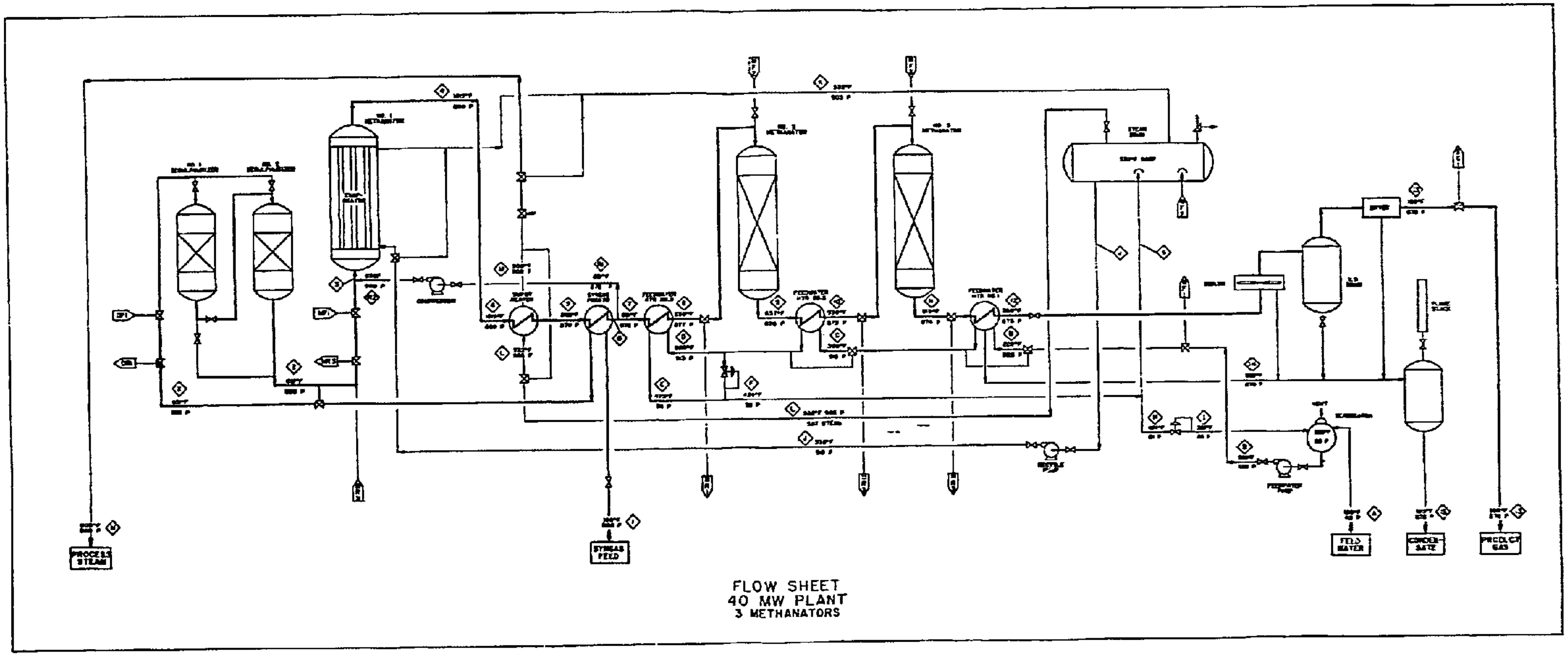

FIGURE 1. FLOW SHEET, 40 MW PLANT, 3 METHANATORS

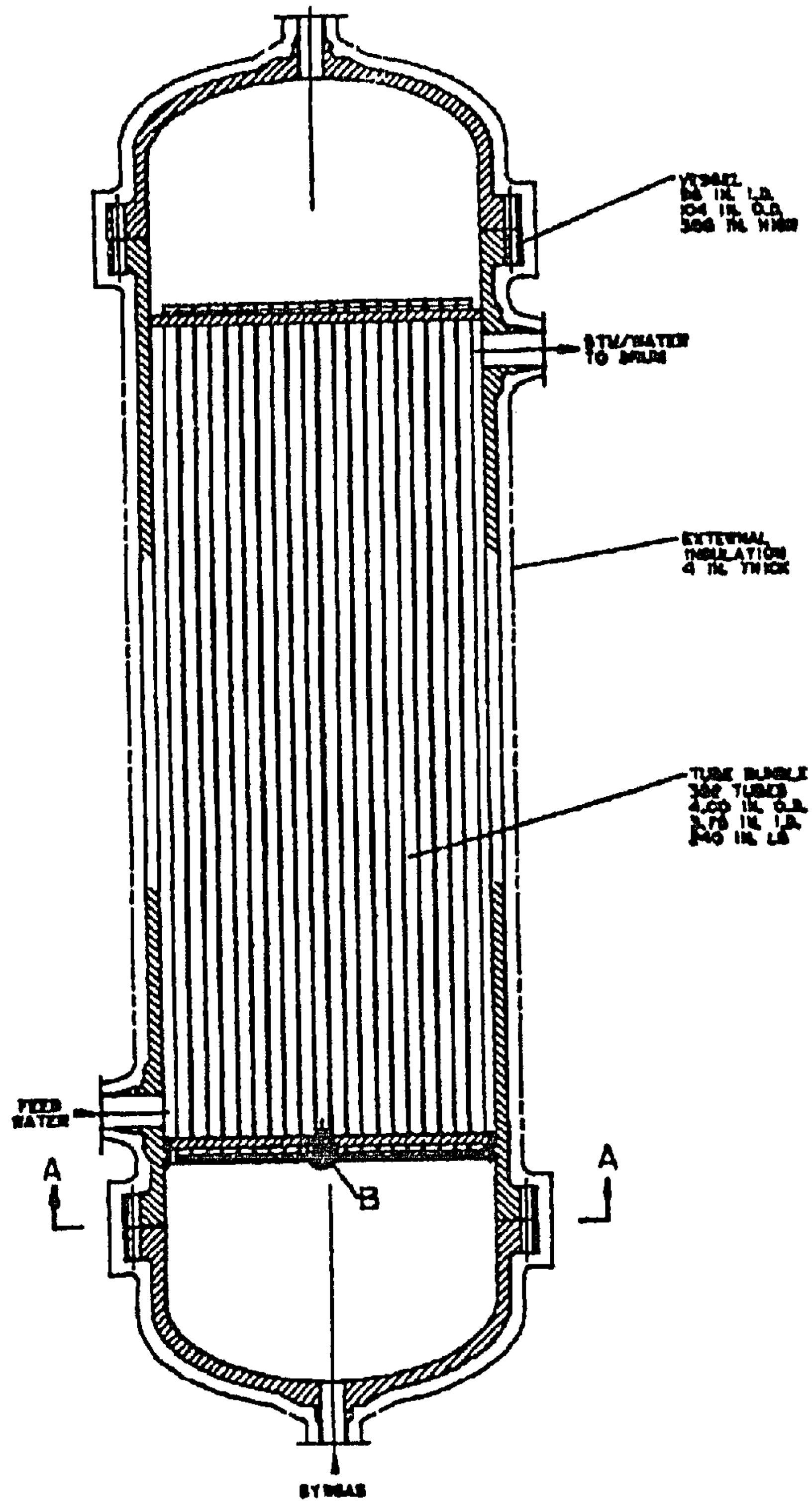

FIGURE 2. METHANATOR NO. 1

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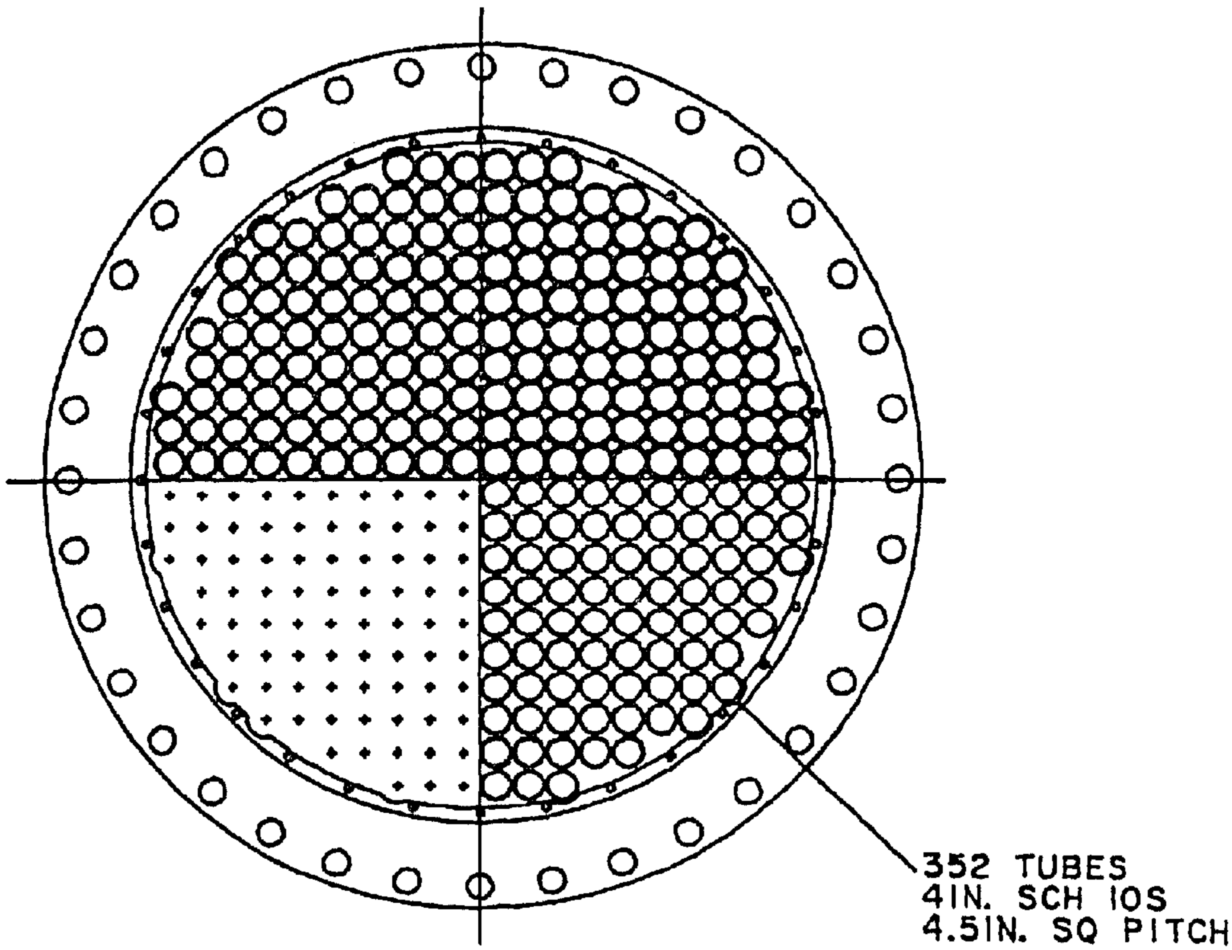

FIGURE 3. METHANATOR NO. 1 — SECTION A-A

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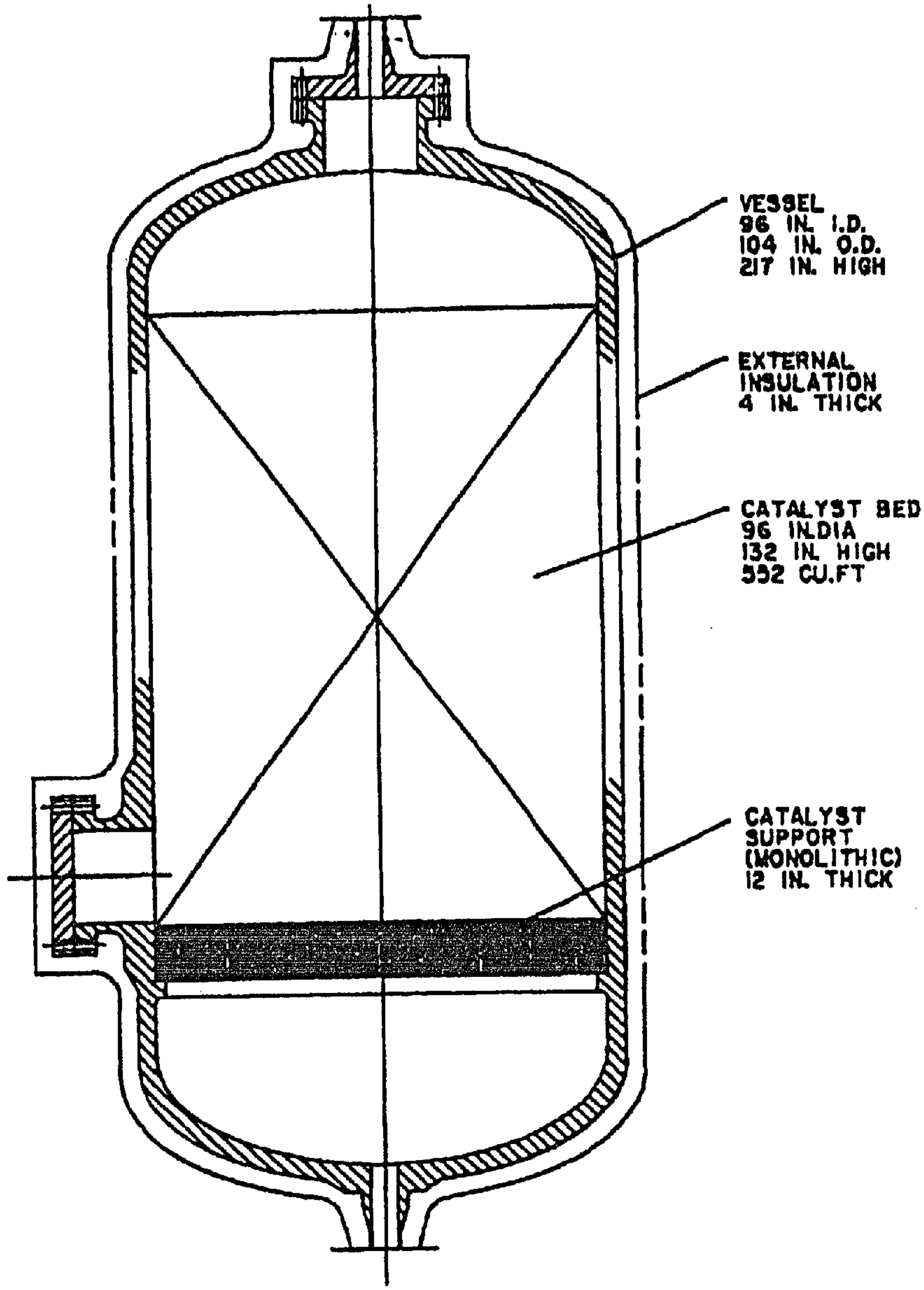

FIGURE 4. METHANATOR 2 & 3

YL-600-10123
J. B. Noriel
August 28, 1981

APPENDIX A7

PRELIMINARY ANALYSIS OF THE METHANATOR/EVAPORATOR AND SUPERHEATER TUBESHEETS

REF:

1. YL-611-10063, Chi-Kou Fan to P. T. Hughes, "Design Basis Methanation Plant Heat Transport System Design," 5/21/81.

2. YL-492-100253, G. J. Licina to C. R. Davis/P. T. Hughes, "Revised Materials Recommendations for Methanator Plant," 6/15/81.

The purpose of this memo is to summarize the results of the sizing and fatigue analysis done on the methanator and superheater tubesheets.

TUBESHEET THICKNESS

Preliminary analysis and sizing of the tubesheets, done in accordance with ASME Code, Section III Article A-8000, 1980, produced the following minimum tubesheet thickness for the two given conditions:

	(TUBESHEET THICKNESS)	
	ΔP_{normal} (in.)	$\Delta P_{emergency}$ (in.)
Methanator/Evaporator	9.18	33.67
Superheater	3.15	12.52

see Attachments 1 and 2.

It was to my understanding that the thicknesses generated by the emergency conditions were excessive, therefore, the following additional analysis is based on the normal condition thicknesses.

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FATIGUE EVALUATION

Approximate methods were used to evaluate the fatigue effects on the tubesheets. This method neglected any thermal loads and is based solely on mechanical loads. Using 50 applied cycles during the expected lifetime, the analysis showed some limitations in the amount of thermal loading which the tubesheets could sustain. The fatigue damage fraction for the mechanical loadings follow:

METHANATOR/EVAPORATOR	0.031	(See pg. 6 of Attachments
SUPERHEATER	0.020	1 and 2)

An approximation of the thermal load effects was evaluated to determine the sensitivity of the tubesheets to temperature differentials. The equations used in this approximation established the sensitivity of the peak stress intensity as a function of the difference between the mean tubesheet temperature, T_m, and the temperature at the tubesheet surface, T_s, $\Delta T = T_m - T_s$. To maintain fatigue damage fractions equal to 1, the respective tubesheets could sustain the following temperature differentials:

METHANATOR/EVAPORATOR	$\Delta T = 190°F$
SUPERHEATER	$\Delta T = 148°F$

The graphs on pg. 6 of Attachments 1 and 2 show the relationship between fatigue damage fraction and temperature differential. The fatigue damage fraction should represent cumulative effects of all events. Our results reflect fatigue contributions of only one event. In order to insure the integrity of the tubesheets, the damage fraction contribution of this one event must be kept low enough to insure sufficient margins for other, as yet, undefined events and still maintain a total fatigue damage fraction <u>less</u> than 1 for total lifetime.

It appears that the mechanical load contribution is sufficiently low but the actual ΔT's present problems. Since the number of tubes running through the tubesheet is large, it is reasonable to assume that the tubesheet will experience a mean temperature equal to the syngas temperature. Under this assumption our $\Delta T = T_m - T_s$ would be the difference between the sygas temperature and the surface temperature and exceeds the limits established above. This fact produces fatigue damage fractions exceeding 1. Additionally, our analysis does not consider the effects of creep and plastic deformations, therefore, compounding the problem.

RECOMMENDATIONS

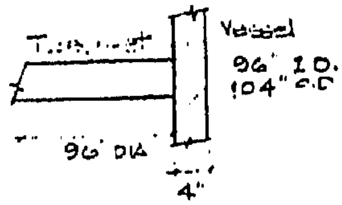
Due to the significant effect the temperature differential has on the fatigue damage fraction, the use of thermal barriers and insulation of the tubesheets should be considered. By applying these, the ΔT across the tubesheet could be minimized. In doing so, the stress allowables would increase. Minimizing the temperature effects allows larger margins to compensate for creep, plastic deformations, and other undefined events. Alternatively, an increase in the tubesheet thickness would produce a proportional decrease in the primary stresses thereby increasing the margin for thermal load contribution to the secondary stresses. Assuming that the thermal gradients are low, the stresses caused by the structural interaction between the tubesheet & vessel walls would be low thereby keeping our secondary stresses intensities to a minimum. This in turn allows more room for the ΔT contribution to the fatigue damage fraction.

These recommendations may be further analyzed in detail when the actual operating load conditions are provided.

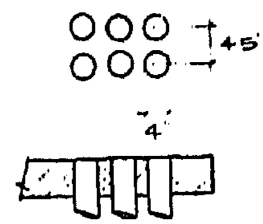
p

COMPUTATION/RECORD SHEET

Prepared By: [illegible] Checked By: Date: 8/18/81	GENERAL ELECTRIC COMPANY Advanced Reactor Systems Department METHANATOR / EVAPORATOR TITLE TUBESHEET ANALYSIS	Page No: 1 of 6 Attachment 1



336 Holes
4 in O.D.
3.624 in. I.D.
4.5 in SQ Pitch



$d = 4"$
$p = 4.5"$
$h = p - d = 0.5"$

Ligament Efficiency $\quad \eta = \frac{h}{p} = \frac{0.50}{4.50} = 0.111$

$\frac{E^*}{E} = 0.02 \qquad \nu^* = 0.66$

Tubesheet Effective Radius $\quad R^* = r_o + \frac{1}{2}(p-h)$

$r_o = 42.75 \sim 43"$

$R^* = 42.75 + \frac{1}{4}(4.5 - 0.5) = 43.75" \sim 44"$

Effective Diameter : 88"

$\boxed{\text{DESIGN FOR } \Delta P_{emergency}}$

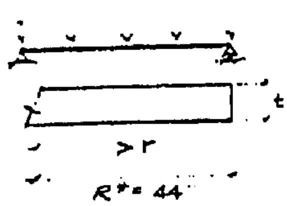
Worst Condition : $\Delta p = 910 - 0 = 910$ psi
Assume 20 hr. duration
Material $2\frac{1}{4}$ Cr - 1 Mo (Properties from N47 High Temp Rules Class I)

$S_t = 15.36$ ksi $= 15360$ psi
$S_{mt} = 14.94$ ksi $= 14940$ psi
$S_m = 15.27$ ksi $= 15270$ psi
} @ 1015°F

Note: The following analysis/calculations based on assumptions that a 910 psi pressure differential of the worst case condition constitutes an emergency condition. The assumed duration is 20 hrs. with any damage being repairable as specified by the emergency classification. The calculations will be based on a pinned connection between tubesheet and vessel wall under the assumption that the tubesheet thickness will be large relative to the vessel wall thickness. Analysis done under ASME Code Section III Div. I Appendix A. Article A-8000.

UNIFORMLY DISTRIBUTED PRESSURE LOADS.

Pinned Condition :



$$\sigma_r = \frac{3}{8}\left(\frac{R^*}{t}\right)^2 (3+\nu^*)\left[1 - \left(\frac{r}{R^*}\right)^2\right] \Delta P$$

$$\sigma_\theta = \frac{3}{8}\left(\frac{R^*}{t}\right)^2 \left[(3+\nu^*) - (1+3\nu^*)\left(\frac{r}{R^*}\right)^2\right] \Delta P$$

p

COMPUTATION/RECORD SHEET

Prepared By: Joby Hamel	GENERAL ELECTRIC COMPANY	Page No: 2 of 6
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Determination of Minimum t:

Allowable Stresses : $P_L + P_b \leq K_b S_t = 1.25\ S_t$; $P_L \cdot P_b \leq 1.80\ S_m$ } minimum governs

@ $r = 0$ $\Delta p = 910$ psi

$\sigma_r = \sigma_\theta \Longrightarrow \frac{\sigma_r}{\sigma_\theta} = \frac{\sigma_\theta}{\sigma_r} = 1.00$

$-1 \leq \beta \leq 1$ ∴ $\beta = 1.00$ $K = 1.00$

$S_I = K\left(\frac{P}{h}\right)\sigma_{ave} = 1.00\left(\frac{4.50}{0.50}\right)\sigma_{ave} \leq$ $1.25(15360) = 19200$ psi ; $1.80(15270) = 27486$ psi (Material Properties N47)

$K\left(\frac{P}{h}\right)\sigma_{ave} = 19200$

$1.00\left(\frac{4.50}{0.50}\right)\sigma_{ave} = 19200$

$\sigma_{ave} = 2133$ psi

$\sigma_r = \sigma_\theta = \frac{3}{8}\left(\frac{44}{t}\right)^2(3+0.66)\left[1-\left(\frac{0}{44}\right)^2\right]910 = 2133$ psi

$t = 33.67$ inches minimum

Check:

@ $r = R^* = 44"$ $t = 33.67"$ $\Delta p = 910$ psi

$\sigma_\theta = \frac{3}{8}\left(\frac{44}{33.67}\right)^2\left[(3.66)-(1+1.98)\left(\frac{44}{44}\right)^2\right]910$

$= 396.28$ psi

$\sigma_r = \frac{3}{8}\left(\frac{44}{33.67}\right)^2(3.66)[1-1]910$

$= 0$ psi

$\frac{\sigma_r}{\sigma_\theta} = \frac{0}{396.28} = 0$ $\beta = 0$ $K = 1.15$

$S_I = K\left(\frac{P}{h}\right)\sigma_{ave} = 1.15\left(\frac{4.50}{0.50}\right)396.28 = 4102$ psi < 19.2 ksi

S_I (@ $r=0$) $>$ S_I (@ $r = R^*$)
controls

⟶ Minimum tubesheet thickness required to satisfy this emergency condition is 33.67 inches.

p
p
p
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p
p
p
p
p
p

COMPUTATION/RECORD SHEET

Prepared By: John N[illegible]	GENERAL ELECTRIC COMPANY	Page No: 3 OF 6
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Date: July 7 1981	METHANATOR/EVAPORATOR TITLE TUBESHEET ANALYSIS	

$\boxed{\text{DESIGN FOR } \Delta P_{normal}}$

Δp = 905 - 880 = 25 psi

15 years = 131400 hrs. T = 1015°F

Material = 2 1/4 Cr - 1 Mo (properties from High Temp Rules N47 Class I)

S_t @ 1015°F 100 000 hrs = 5.82 ksi

S_t @ 1015°F 300 000 hrs = 4.64 ksi

$\therefore$ S_t @ 1015°F 131400 hrs = 5.67 ksi = 5670 psi

$\left.\begin{array}{l} S_m = 15.27 \text{ ksi} \\ S_m = 15270 \text{ psi} \end{array}\right\}$ @ 1015°F

Note: The following calculations are based on service load conditions with a maximum temperature differential of 25 psi under normal operating conditions. Again calculations are based on pinned conditions because it produces maximum values for stresses.

Determination of minimum t:

Allowable Stresses: $\left.\begin{array}{l} P_L+P_b \le K_t S_t = 1.25\, S_t = 1.25(5670) = 7088 \text{ psi} \\ P_L+P_b \le 1.5\, S_m = 1.5(15270) = 22905 \text{ psi} \end{array}\right\}$ minimum

$P_L+P_b \le 1.5\, S_o = 1.5\,(7200) = 10800$ psi

@ $r = 0$ $\Delta p = 25$ psi

$\sigma_r = \sigma_\theta \longrightarrow \frac{\sigma_r}{\sigma_\theta} = \frac{\sigma_\theta}{\sigma_r} = 1.00$ $\beta = 1.00$ $K = 1.00$

$S_I = K\left(\frac{P}{h}\right)\sigma_{ave} = 1.00\left(\frac{4.5}{0.55}\right)\sigma_{ave} \le 7088$ psi

$\sigma_{ave} = 787.56$ psi

$\sigma_r = \sigma_\theta = \frac{3}{8}\left(\frac{44}{t}\right)^2(3+0.66)\left[1 - \left(\frac{0}{44}\right)^2\right]25 = 787.56$ psi

$t = 9.18$ inches.

$\longrightarrow$ For $\Delta p = 25$ psi @ 1015°F for 131400 hrs. the minimum tubesheet thickness should be 9.18 inches.

@ $r = R^*$, $\sigma_r = 0$ and σ_θ is < $\sigma_{\theta(r=0)}$ by inspection.

p

COMPUTATION/RECORD SHEET

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Date: 8/12/81	TITLE METHANATOR EVAPORATOR TUBESHEET ANALYSIS	

FATIGUE CONSIDERATIONS

Section A-8142.2 Combined Mechanical and Thermal Effects

Note: This calculation based on mechanical loadings no specific data given on thermal loads, normal operating condition.

a.) Stress Intensity Range (A-8142.2 Eq. 19)

$K = 1.00$ (@ $r=0$)
$p = 4.50$
$d = 4.00$
$h = 0.50$

$$S_{II} = K\left(\frac{p}{h}\right)\sigma_1 \leq S_y\,(avg) \qquad (T\text{-}1323 \ N47)$$

$S_y = 30.0$ ksi
S_y @ 1015°F = 15.94 ksi
$S_y\,(avg) = 22.97$ ksi

S_{II} = stress intensity based on stresses average across minimum ligament width but not through thickness of plate.

$$S_{II} = 1.00\left(\frac{4.50}{0.50}\right)\sigma_1 \leq 22.97 \text{ ksi}$$

$$\sigma_1 = 2552 \text{ psi}$$

b.) Peak Stress Intensity: (@ $r=0$) Thermal Loads Neglected:

$$S_{V} = Y_{max}\left(\frac{p}{h}\right)\sigma_1 + P_s \qquad (A\text{-}8142.2 \ Eq.\ 20)$$

$\sigma_1 = \sigma_2 \longrightarrow \frac{\sigma_2}{\sigma_1} = 1 \quad \therefore \beta = 1 \quad \therefore \quad Y_{max} = 1.30$ (Fig. A-81422)

P_s = pressure @ surface = 25 psi
$\sigma_1 = \sigma_1$ from (a.) above = 2552 psi
$\sigma_1 = P_L + P_b$ (mechanical)
$\sigma_1 = [P_L + P_b + Q]$ Limit

Assuming $Q = 0$, no thermal loads given:

$$S_{V} = 1.30\left(\frac{4.50}{0.50}\right)(2552) + 25$$

$$S_{V} = 29\,893 \text{ psi}$$

S_{V} = peak stress intensity due to all loadings with fatigue considerations.

p

COMPUTATION/RECORD SHEET

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d.) Peak Stress Intensity @ Outermost Hole:

$S_{rim} = K_r \sigma_{rim} + P_s$ (A-8142-2 Eq. 22)

K_r = stress multiplier (Fig. A-8142-6)

$\rho / b = \frac{2.00}{2.00} = 1.00$

$\therefore K_r \approx 1.50$

b = thickness of outside ring

ρ = tube radius

Assume: $\sigma_{rim} = \sigma_\theta \; (r = R^*)$

$t = 3.18$ inches

$\Delta p = 25$ psi

$\sigma_\theta = \frac{3}{8}\left(\frac{44}{3.18}\right)^2 \left[3.66 - (1 + 3(0.66))(1)^2\right] 25$

$\sigma_\theta = 146.45$ psi

$\therefore \quad S_{rim} = 1.50(146.45) + 25$

$S_{rim} = 245$ psi

e.) Peak Stress Intensities, Thermal Skin Stresses Included:

$\sigma_r = \sigma_\theta = K_{skin} \frac{E\alpha}{(1-\nu)} (T_m - T_s)$ (A-8152 Eq. 24)

K_{skin} = stress ratio (Fig. A-8153-1)

$$\left.\begin{array}{l} p = 4.50 \\ h = 0.50 \\ \frac{h}{p} = \frac{1}{9} = 0.111 \end{array}\right\} \therefore K_{skin} = 0.07$$

E, α, ν = unmodified material properties @ 1015°F

T_m = mean temperature of plate °F

T_s = surface temperature of plate °F

$\sigma_r = \sigma_\theta = 0.07 \frac{(22.61 \times 10^6)(7.832 \times 10^{-6})}{(1 - 0.30)} (T_m - T_s)$

$\sigma_r = \sigma_\theta = 17.71 \; \Delta T$ Sensitivity

p

COMPUTATION/RECORD SHEET

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STRESS RANGE EVALUATION (Approximation)

a.) Using the stress S_{III} :

$\Delta p = 25$ psi
$\bar{T} = 1015°F$
t = 15year lifespan
n_i = 50 cycles
$S_{III} = 29883$ psi
$E = 22.61 \times 10^6$ psi
Q = thermal loading = 0

$E = \frac{\sigma}{\epsilon} = \frac{S_{III}}{\epsilon}$

$\epsilon = \frac{S_{III}}{E} = \frac{29883}{22.61 \times 10^6}$

$\epsilon = 0.00132$

From strain range plot Fig. T-1430-1C (N47):

$N_d \sim 1611$ cycles $\Longrightarrow$ Allowable Cycles

$\Sigma \frac{n_i}{N_d} = \frac{50}{1611} = 0.031$ (T-1400 N47) Fatigue Damage Fraction

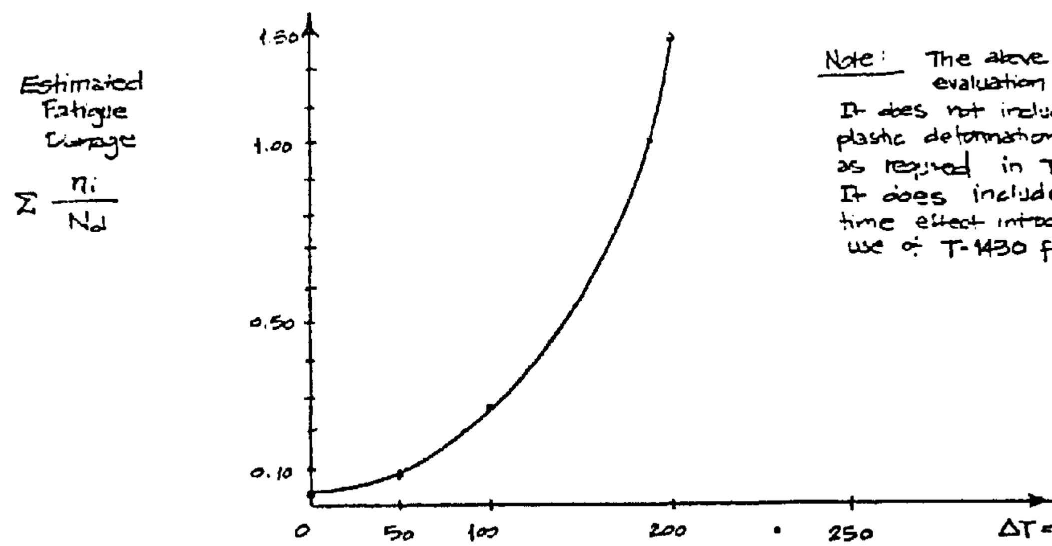
b.) Assuming a thermal load in order to determine sensitivity:
From Previous section : $\sigma_T = 17.71\ \Delta T$

Given the relationship for the thermal skin stress determined by the above relationship, the thermal skin effect can be added to the mechanical stresses σ_i to get an estimation of total fatigue damage. These results are given in the following graph:

$\sigma_\Sigma = \sigma_1 + \sigma_2 = 2552 + 17.71\,(T_m - T_S)$

$S_\Sigma = 1.30\left(\frac{4.50}{0.50}\right)\sigma_\Sigma + 25$

$\epsilon = \frac{S_\Sigma}{E}$



Note: The above fatigue evaluation is approximate. It does not include creep or plastic deformation considerations as required in T-1400 Eq 7. It does include the hold time effect introduced by the use of T-1430 fatigue curve.

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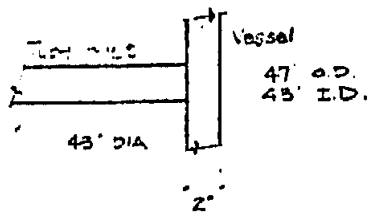
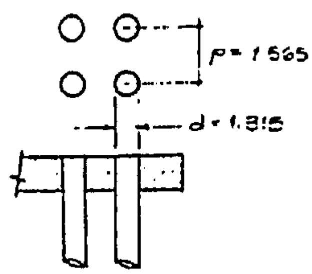




500 Tubes
1 inch SCH 10S

$d = 1.315$
$p = 1.565$
$h = 0.25$

Ligament Efficiency : $\eta = \frac{h}{p} = \frac{0.25}{1.565} = 0.16$

$\frac{E^*}{E} = 0.08 \qquad \nu^* = 0.56$

Tubesheet Effective Radius $\qquad R^* = r_o + \frac{1}{4}(p-h)$

$r_o = 19.56"$

$R^* = 19.56 + \frac{1}{4}(1.565 - 0.25) = 19.89 \sim 20"$

Effective Diameter $\quad 40"$

DESIGN FOR $\Delta P_{emergency}$

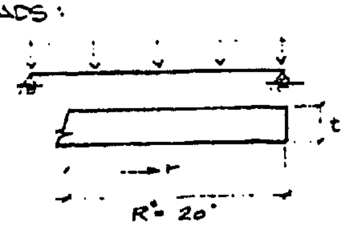
Worst Condition : $\Delta p = 900 - 0 = 900$ psi
Assume 20 hr. duration
Material $2\frac{1}{4}Cr - 1Mo$ (Properties from N47 High Temp Rules Class I)

$S_t = 15360$ psi
$S_{rt} = 14940$ psi
$S_m = 15270$ psi

Note : The following calculations/analysis based on assumptions that a 900 psi pressure differential of the worst case condition constitutes an emergency condition. The assumed duration is 20 hrs. with any damage being repairable. The calculations will be based on a pinned connection between tubesheet and vessel wall under the assumption that the tubesheet thickness will be large relative to the vessel wall thickness. Analysis done under ASME Code Section III Div. I, Appendix A, Article A-8000.

UNIFORMLY DISTRIBUTED PRESSURE LOADS :

Pinned Condition :



$$\sigma_r = \frac{3}{8}\left(\frac{R^*}{t}\right)^2 (3+\nu^*)\left[1-\left(\frac{r}{R^*}\right)^2\right]\Delta p$$

$$\sigma_\theta = \frac{3}{8}\left(\frac{R^*}{t}\right)^2 \left[(3+\nu^*) - (1+3\nu^*)\left(\frac{r}{R^*}\right)^2\right]\Delta p$$

p

COMPUTATION/RECORD SHEET

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Determination of Minimum t :

Allowable stresses : $P_L + P_b \leq K_t S_t \leq 1.25 S_t$ } minimum governs
$P_L + P_b \leq 1.8 S_m$ }

@ r = 0 $\Delta p = 900$ psi

$\sigma_r = \sigma_\theta \longrightarrow \frac{\sigma_r}{\sigma_\theta} = \frac{\sigma_\theta}{\sigma_r} = 1.00 = \beta$ $-1 \leq \beta \leq 1$ (Material Properties N47)

$\therefore \beta = 1.00$ $K = 1.00$

$S_I = K\left(\frac{P}{h}\right)\sigma_{ave} = 1.00\left(\frac{1.565}{0.25}\right)\sigma_{ave} \leq$ 1.25 (15360) = 19200 psi
1.8 (15270) = 27486 psi

$1.00\left(\frac{1.565}{0.25}\right)\sigma_{ave} = 19200$

$\sigma_{ave} = 3067$ psi

$\sigma_r = \sigma_\theta = \frac{3}{8}\left(\frac{20}{t}\right)^2 (3+0.56)\left[1+\left(\frac{0}{20}\right)^2\right] 900 = 3067$ psi

$t = 12.52$ inches minimum

Check :
@ $r = R^* = 20''$ $t = 12.52''$ $\Delta p = 900$ psi

$\sigma_\theta = \frac{3}{8}\left(\frac{20}{12.52}\right)^2\left[(3.56) - (1+1.68)\left(\frac{20}{20}\right)^2\right] 900$

$= 757.9$ psi

$\sigma_r = \frac{3}{8}\left(\frac{20}{12.52}\right)^2 (3.56)\left[1-1\right] 900$

$= 0$ psi

$\frac{\sigma_r}{\sigma_\theta} = \frac{0}{757.9} = 0$ $\beta = 0$ $K = 1.15$

$S_I = K\left(\frac{P}{h}\right)\sigma_{ave} = 1.15\left(\frac{1.565}{0.25}\right) 757.9 = 5456$ psi

S_I (@ r = 0) > S_I (@ r = R*)
controls

⟶ Minimum tubesheet thickness required to satisfy this emergency condition is 12.52 inches.

p

COMPUTATION/RECORD SHEET

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DESIGN FOR ΔP_{normal}

ΔP = 900 - 879 = 21 psi

15 yrs. = 131400 hrs.

Material 2¼ Cr - 1Mo T = 1015°F

S_t @ 1015°F = 5670 psi (Material Properties N47)

S_m = 15270 psi @ 1015°F

Note: The following calculations are based on service load conditions with a maximum temperature differential of 21 psi under normal operating conditions. Calculations are based on a pinned connection between tubesheet and vessel wall because it produces maximum stress values.

Determination of Minimum t:

Allowable Stress: $\left.\begin{matrix} P_L + P_b + K_t S_t \le 1.25 S_t \\ P_L + P_b \le 1.5 S_m \end{matrix}\right\}$ minimum

$1.25 S_t = 1.25(5670) = 7088$ psi (Material Properties N47)

$1.00 S_m = 1.50(15270) = 22905$ psi

@ r = 0 Δp = 21 psi

$\sigma_r = \sigma_\theta \longrightarrow \frac{\sigma_r}{\sigma_\theta} = \frac{\sigma_\theta}{\sigma_r} = 1.00 \quad \therefore \beta = 1.00 \qquad K = 1.00$

$S_I = K\left(\frac{P}{h}\right)\sigma_{ave} = 1.00\left(\frac{1.565}{0.25}\right)\sigma_{ave} = 7088$ psi

$\sigma_{ave} = 1132.27$ psi

$\sigma_r = \sigma_\theta = \frac{3}{8}\left(\frac{20}{t}\right)^2(3+0.56)\left[1+\left(\frac{0}{20}\right)^2\right]21 = 1132.27$ psi

$t = 3.15$ inches

⟶ For Δp = 21 psi @ 1015°F for 131400 hrs. the minimum tubesheet thickness should be 3.15 inches.

@ $r = R^*$ $\sigma_r = 0$ $\sigma_\theta < \sigma_{\theta (r=0)}$ by inspection.

p

COMPUTATION/RECORD SHEET

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FATIGUE CONSIDERATIONS

Section A-8142·2 Combined Mechanical and Thermal Effects

Note: This calculation based on mechanical loadings, no specific data given on thermal loads, normal operating condition.

a) Stress Intensity Range (A-8142.2 Eq. 19)

$K = 1.00 \ (@ \ r=0)$
$p = 1.565$
$h = 0.25$

$S_{II} = K\left(\frac{p}{h}\right)\sigma_1 \leq S_y \ (avg)$ (T-1323 N47)

$S_y = 30.0$ ksi
S_y @ 1015°F = 15.94 ksi
S_y (avg.) = 22.97 ksi

S_{II} = stress intensity based on stresses averaged across minimum ligament width but not through thickness of plate.

$S_{II} = 1.00\left(\frac{1.565}{0.25}\right)\sigma_1 \leq 22970$

$\sigma_1 = 3669$ psi

b) Peak Stress Intensity: (@ r=0) Thermal Loads Neglected:

$S_{III} = Y_{max}\left(\frac{p}{h}\right)\sigma_1 + P_S$ (A-8142-2 Eq. 20)

$\sigma_1 = \sigma_2 \longrightarrow \frac{\sigma_2}{\sigma_1} = 1 \ \therefore \ \beta = 1 \ \therefore \ Y_{max} = 1.20$ (Fig. A-8142-2)

P_S = pressure @ surface = 21 psi
$\sigma_1 = \sigma_1$ from (a) above = 3669 psi
$\sigma_1 = P_L + P_b$ (mechanical)
$\sigma_1 = [P_L + P_b + Q]$ Limit

Assuming $Q = 0$, no thermal loading:

$S_{III} = 1.20\left(\frac{1.565}{0.25}\right)(3669) + 21$

$S_{III} = 27583$ psi

S_{III} = peak stress intensity due to all loadings with fatigue considerations.

p

COMPUTATION/RECORD SHEET

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d) Peak Stress Intensity @ Outermost Hole :

$S_{rim} = K_r \sigma_{rim} + P_s$ (A-8142-2 Eq. 22)

K_r = stress multiplier (Fig. A8142-6)

$\rho/b = \frac{0.65"}{1.500} = 0.438$

$\therefore K_r \cong 1.50$

b = thickness of outside ring

ρ = tube radius

Assume $\sigma_{rim} = \sigma_\theta$ @ $(r = R^*)$

t = 3.15 inches

Δp = 21 psi

$\sigma_\theta = \frac{3}{8}\left(\frac{20}{3.15}\right)^2 [3.56 - (1 + (3)(0.56)(1)^2] \, 21$

σ_θ = 279.37 psi

$\therefore S_{rim} = 1.50(279.37) + 21$

S_{rim} = 440 psi

e.) Peak Stress Intensity , Thermal Skin Stresses Included :

$\sigma_r = \sigma_\theta = K_{skin} \frac{E\alpha}{(1-\nu)} (T_m - T_s)$ (A 8152 Eq. 24)

K_{skin} = stress ratio (Fig. A-8153-1)

$$\left.\begin{array}{l} p = 1.565 \\ h = 0.25 \\ h/p = \eta = 0.160 \end{array}\right\} \therefore K_{skin} = 0.14$$

E, α, ν = unmodified material properties @ 1015°F

T_m = mean plate temperature

T_s = plate surface temperature

$$\sigma_r = \sigma_\theta = 0.14 \frac{(22.61 \cdot 10^6)(7.832 \times 10^{-6})}{(1 - 0.30)} (T_m - T_s)$$

$\sigma_r = \sigma_\theta$ = 35.42 ΔT Sensitivity

p

COMPUTATION/RECORD SHEET

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Date: 9/20/81	TITLE SUPERHEATER TUBESHEET ANALYSIS	

$\boxed{\text{STRESS RANGE EVALUATION}}$ (Approximation)

a) Using the stress S_{II}:

$\Delta p = 21$ psi
$T = 1515°F$
$t = 15$ years
$n_i = 50$ cycles
$S_{II} = 27583$ psi
$E = 22.61 \times 10^5$ psi
Q = thermal loads = 0

$$E = \frac{\sigma}{\epsilon} = \frac{S_{II}}{\epsilon}$$

$$\epsilon = \frac{S_{II}}{E} = \frac{27583}{22.61 \times 10^6}$$

$$\epsilon = 0.00122$$

From strain range plot Fig. T-1430-1C (N47):

$N_d = 2462$ allowable cycles

$$\Sigma \frac{n_i}{N_d} = \frac{50}{2462} = 0.02$$

(T-1400 N47)
Fatigue Damage Fraction

b) Assuming a thermal load in order to determine sensitivity

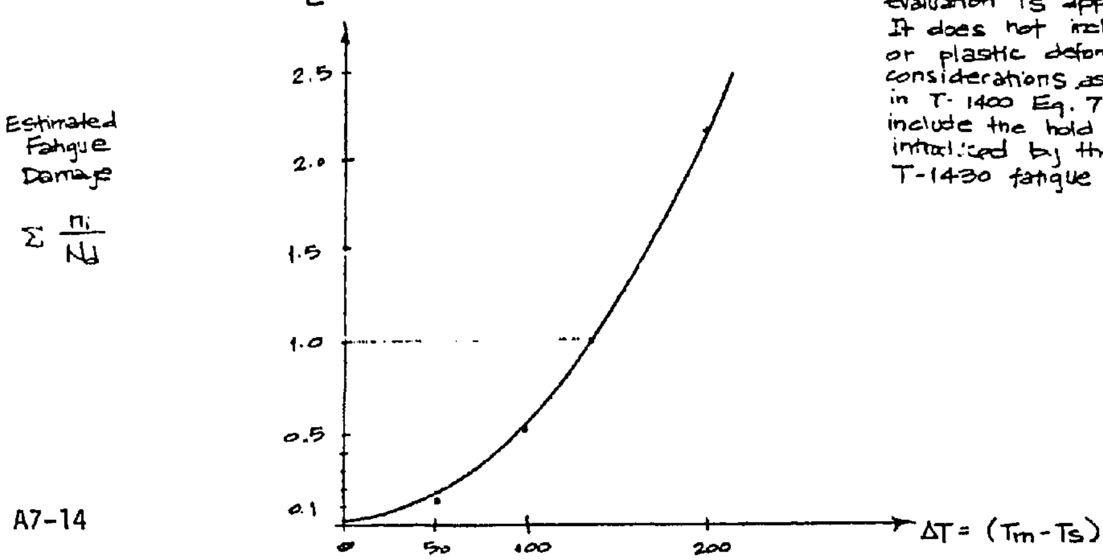
From Previous Section $\sigma_r = 35.42\,\Delta T$

Given the relationship for the thermal skin stress determined by the above relationship, the thermal skin effect can be added to the mechanical stresses σ_1 to get an estimation of total fatigue damage. These results are given in the following graph:

$$\sigma_\Sigma = \sigma_1 + \sigma_r = 3669 + 35.42\,\Delta T$$

$$S_\Sigma = 1.26\left(\frac{1.565}{0.25}\right)\sigma_\Sigma + 21$$

$$\epsilon = \frac{S_\Sigma}{E}$$



Note: The above fatigue evaluation is approximate. It does not include creep or plastic deformation considerations as required in T-1400 Eq. 7. It does include the hold time effects introduced by the use of T-1430 fatigue curve.

p

YL-600-10128
J. B. Noriel
September 9, 1981

APPENDIX A8

PRELIMINARY SIZING AND STRESS ANALYSIS OF THE METHANATOR/EVAPORATOR AND SUPERHEATER HEAD TO VESSEL FLANGE CONNECTION

REF: 1. YL-611-10063, Chi-Kou Fan to P. T. Hughes, "Design Basis Methanation Plant Heat Transport System Design," 5/21/81.

2. YL-492-100253, G. J. Licina to C. R. Davis/P. T. Hughes, "Revised Material Recommendations for Methanator Plant," 6/15/81.

SUMMARY

The purpose of this memo is to transmit the results of the analysis of the head to vessel flange connection with respect to bolt connections, flange thicknesses, and stresses. It has been discussed that the Methanator/Evaporator and Superheater analysis could be governed by ASME Code Section VIII. For purposes of our analysis, the methods of ASME Code Section III, Article XI 3000, RF Flanges have to be used. With geometry as provided in Ref. 2, a flange thickness of 30.35 inches for the Methanator/Evaporator and 16.30 inches for the Superheater were found to meet minimum code requirements.

BOLT CONNECTIONS

Dimensions for the bolt circle diameter, bolt size, and number of bolts were specified. This data produced a required bolting area which was compared to the provided actual bolting area. Allowable bolt stress was established under the assumption that the bolting material would be of comparable strength to the flange material. Table I-14.12 of code case N-47 provides allowable bolt stresses for various materials and temperatures. Based on this assumption, the allowable bolt stress was established at 30 ksi. Summary of results follow:

	Bolt Load (kips)	Bolt Circle Diameter (in.)	Bolt Nominal Diameter (in.)	Bolt Area $(in)^2$	No. Bolts	Total Bolt Area $(in.)^2$	Required Bolt Area $(in.)^2$	Status
METHANATOR/ EVAPORATOR	9834	114	3 5/8	10.32	48	495.36	327.8	ok
SUPERHEATER	2542	57	3 5/8	10.32	48	495:36	84.73	ok

p

FLANGE THICKNESS

Using the methods of Article XI-3000, minimum flange thicknesses were determined for the given pressure and temperature conditions.

	LOAD CONDITIONS		Minimum Flange Thickness (in.
	Pressure (psi)	Temperature (°F)	
METHANATOR/EVAPORATOR	880	1015°	30.35
SUPERHEATER	880	1015°	16.30

The methanator/evaporator having a larger vessel diameter produces a larger flange thickness under the same conditions.

FLANGE STRESSES

The stresses produced by these thicknesses are tabulated on page 4 of Attachments 1 and 2. With respect to the stress limits of Section XI-3250 of the code, the flanges meet the minimum stress requirements.

HYDROSTATIC TEST

Article XII 1000 supplies a provision for the hydrostatic test. This test is used to determine the tightness of the flange joint. The vessel is subjected to a hydrostatic test pressure equal to 1 1/2 times the design pressure and the test is performed at room temperature. The actual design pressure has not been established, but for this analysis has been assumed equal to the operating pressure. Under the test condition, the flanges must meet the stress requirements of Code Case N-47-3226. Results follow:

	S_y @ RT (psi)	Allowable Code Stress ($P_L+P_b=1.35S_y$ @ RT (psi))	Maximum Stress combination (psi)	Status
METHANATOR/EVAP.	30,000	40,500	16,947	ok
SUPERHEATER	30,000	40,500	16,988	ok

See page 5 of Attachments 1 and 2.

Since a large portion of the flange stresses are primarily in bending, by inspection, the primary membrane stress limit $P_m \leq 0.9 S_y$ @ RT is also met. Results show that the flanges meet both the normal operating pressure and hydrostatic test pressure conditions stress requirements. Some of the actual to allowable stress margins are small, but the flange thicknesses represent minimum values only.

This concludes planned analysis on Methanator Structural Support Work Package 113.

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COMPUTATION/RECORD SHEET

Prepared By: J.B Nonal	GENERAL ELECTRIC COMPANY	Page No: 1 of 5
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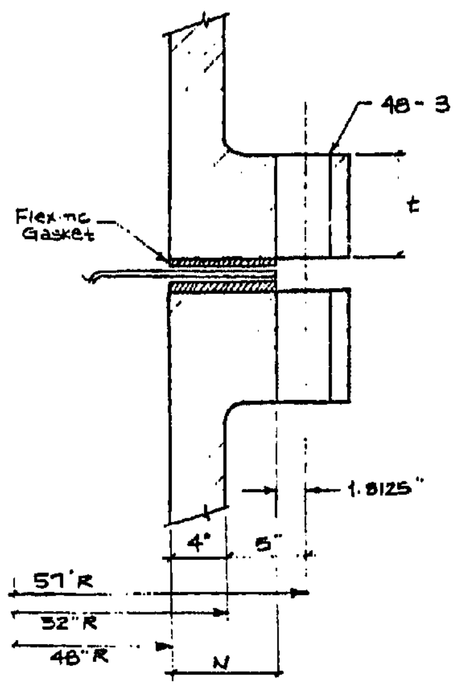
Date: July 15, 1981	TITLE MET... ...R / E... FLANGE ANALYSIS	

REF: ASME Code Section III XI-3200 RF Flanges 1980



Known:

Bolts Size 3 5/8" DIA 48 bolts

$T = 1015°F$

$P = 880$ psi

Gasket: Spiral wound metal, asbestos filler, stainless steel $m = 3.00$, $y = 4500$ psi (Tbl. XI-3221.1-1)

$N = (9" - 1.8125") = 7.1875"$

$b_o = \frac{N}{2} = 3.595" > 1/4"$

$b = \frac{\sqrt{b_o}}{2} = 0.948"$

$t = 30.35"$

Bolt Load For Operating Conditions:

G = diameter location @ gasket load reaction = $d_{o_g} - 2b$

$G = 110.38 - 2(0.948) = 108.48"$

Minimum required bolt load = $W_{m1} = H + H_p = 0.785 G^2 P + 2b(3.14 G m P)$ (XI-3221.1 Eq 1)

$W_{m1} = 0.785(108.48)^2(880) + 2(0.948)(3.14)(108.48)(3)(880)$

$W_{m1} = 9834261.52$ lbs. $= 9834$ K

Minimum required bolt load for gasket seating $= W_{m2} = 3.14 G y b$

$W_{m2} = 3.14(108.48)(4500)(0.948)$

$W_{m2} = 1453115.64$ lbs. $= 1453^{k}$

Total Required and Actual Bolt Area:

$$A_m = \text{maximum of} \begin{cases} W_{m1}/S_b \\ W_{m2}/S_a \end{cases}$$

Assume a minimum allowable stress for the bolts as 30.0×10^3 psi (N47 Tbl I14.12)

$$A_m = \frac{9834}{30} = 327.8 \quad \text{or} \quad \frac{1453}{30} = 48.43$$

$A_m = \frac{327.8 \text{ in}^2}{48 \text{ bolts}} = 6.829 \text{ in}^2/\text{bolt}$

Using 3 5/8" DIA. bolts, $A_b = 10.32$ in²

$A_b \geq A_m$ Satisfies code.

p

COMPUTATION/RECORD SHEET

Prepared By: J.B Konel	GENERAL ELECTRIC COMPANY	Page No: 2 of 5
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Date: July 15, 1981	TITLE Flange Analysis	

Bolt Loads: W =

Operating Conditions $W_{m1} = 9834^{k}$

Gasket Seating $W = \frac{(A_m + A_b) S_a}{2} = 206^{k}$

Flange Moments:

Operating Conditions:

$$M_o = M_D + M_T + M_G$$

$$= H_D h_D + H_T h_T + H_G h_G$$ (XI-3230 (b))

$$= 0.785 B^2 P h_D + (H - H_D) h_T + (W - H) h_G$$

$p = 880$ psi

$B = 96''$

$$h_D = R + 0.5 g_1 = \frac{C-B}{2} - g_1 + 0.5 g_1 = \frac{114-96}{2} - 4 + \frac{1}{2}(4) = 7''$$

$$h_G = \frac{C-G}{2} = \frac{114 - 108.48}{2} = 2.76$$ (Tbl. XI-3230-1)

$$h_T = \frac{R + g_1 + h_G}{2} = \frac{5 + 4 + 2.76}{2} = 5.88$$

$$M_o = 0.785(96)^2(880)(7) + \left[0.785(108.48)^2(880) - 0.785(96)^2(880)\right] 5.88 +$$

$$+ \left[(9834000 - 0.785(108.48)^2(880))\right] 2.76$$

$M_o = 5.96 \times 10^7$ in-lbs

Gasket Seating:

$$M_o = W \frac{(C-G)}{2} = 206000 \frac{(114 - 108.48)}{2}$$ (XI-3221.2 Eq. 2)

$M_o = 568560$ in-lbs

p

COMPUTATION/RECORD SHEET

Prepared By: J.B. Noriel	GENERAL ELECTRIC COMPANY	Page No: 3 of 5
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Date: July 15, 1981	TITLE Flange Analysis	

FLANGE STRESSES

Design Loads $T = 1015°F$
$P = 880$ psi

Code Criteria : Limits $P_m \leq S_o$
$P_L + P_b \leq 1.5 S_o$

For 2¼ Cr - 1Mo : Allowable Design Stress $S_o = 5663$ psi @ 1015°F 132000 hr
(Properties from High Temp Rules N-47)

1. Longitudinal Hub Stress $S_H \leq \left.\begin{matrix} 1.5 S_t \\ 1.5 S_m \end{matrix}\right\}$ minimum

$\therefore \; S_H \leq 1.5(5663) = 8495$ psi (XI-3250)

2. Radial Flange Stress $S_R < S_t = 5663$ psi

3. Tangential Flange Stress $S_T < S_t = 5663$ psi

4. $\dfrac{S_H + S_R}{2} \leq S_t$ and $\dfrac{S_H + S_T}{2} \leq S_t = 5663$ psi

LONGITUDINAL HUB STRESS

$$S_H = \frac{f M_o}{L g_1^2 B} \qquad \text{(XI-3240 Eq. 6)}$$

$$\left.\begin{matrix} g_o = 4'' \\ g_1 = 4'' \end{matrix}\right\} \frac{g_1}{g_o} = 1 \qquad \therefore \quad f = 1 \;\; \text{(Fig. XI-3240-6)}$$

$$L = \frac{te+1}{T} + \frac{t^3}{d} = 6.007$$

$$d = \frac{U}{V} h_o g_o^2 = \frac{10.44}{0.550103}(19.6)(4)^2 = 5951.58$$

$$\left.\begin{matrix} U = 10.44 \\ V = 0.550103 \end{matrix}\right\} \text{(Ref. Fig. XI-3240-1 and XI-3240-3)}$$

$$h_o = \sqrt{B g_o} = \sqrt{96(4)} = 19.60$$

$T = 1.83$ (Fig. XI-3240-1)

$F = 0.908920$ (Fig XI-3240-2)

$$e = \frac{F}{h_o} = \frac{0.908920}{19.6} = 0.046$$

$$K = \frac{A}{B} = \frac{118}{96} = 1.23$$

Assuming $A = 118''$

p

COMPUTATION/RECORD SHEET

Prepared By: JON	GENERAL ELECTRIC COMPANY	Page No: 4 of 5
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Date: 8/24/81	TITLE METHANATOR / EVAPORATOR FLANGE ANALYSIS	

$$S_H = \frac{1.00(5.96 \times 10^7)}{6.007(4.00)^2(96)}$$

$$S_H = 6460 \text{ psi}$$

RADIAL FLANGE STRESS

$$S_R = \frac{(1.33te + 1)M_o}{Lt^2B} \qquad (\text{XI-3240 Eq. 7})$$

$$S_R = \frac{(1.33(30.35)(0.046)+1)(5.96 \times 10^7)}{6.007(30.35)^2(96)}$$

$$S_R = 321 \text{ psi}$$

TANGENTIAL FLANGE STRESS

$$S_T = \frac{YM_o}{t^2B} - ZS_R \qquad (\text{XI 3240 Eq. 8})$$

$Y = 9.50 \qquad Z = 4.90 \qquad$ (Fig. XI 3240-1)

$$S_T = \frac{9.50(5.96 \times 10^7)}{(30.35)^2(96)} - 4.90(321)$$

$$S_T = 4832 \text{ psi}$$

	Actual Stress	Allowable Code Stress	Status
S_H	6460 psi	8495 psi	Pass
S_R	321	5663	Pass
S_T	4832	5663	Pass
$\left(\frac{S_H+S_R}{2}\right)$	3391	5663	Pass
$\left(\frac{S_H+S_T}{2}\right)$	5646	5663	Pass

The given stresses are produced by the minimum flange thickness to satisfy code.

p

COMPUTATION/RECORD SHEET

Prepared By: VBN	GENERAL ELECTRIC COMPANY	Page No: 5 of 5
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Date: 8/20/81	TITLE Flange Analysis	

HYDROSTATIC TEST

$P_{test} = 1.50\ P_{design}$ (XII-1000 (d))

$P_{test} = 1.50\,(880) = 1320$ psi

$P_{Test} = 1320$ psi

T_{test} = room temperature

Bolt Load @ Test:

$W = 0.785(108.48)^2(1320) + 2(0.948)(3.14)(108.48)(8)(1320)$

$W = 14751^K$

Flange Moments:

$M_o = M_D + M_T + M_G$

$M_o = 0.785(96)^2(1320)(7) + [0.785(108.48)^2(1320) - 0.785(96)^2(1320)]\ 5.88 +$
$[14751000 - 0.785(108.48)^2(1320)]\ 2.76$

$M_o = 8.945 \times 10^7$ in-lbs

Flange Stresses:

$$S_H = \frac{1.00\,(8.945)\times 10^7}{6.007\,(4.0)^2(96)} = 9695 \text{ psi}$$

$$S_R = \frac{[1.33(30.35)(0.046) + 1](8.945\times 10^7)}{6.007\,(30.35)^2\,(96)} = 481 \text{ psi}$$

$$S_T = \frac{9.50(8.945\times 10^7)}{(30.35)(96)} - 4.90(481) = 7252 \text{ psi}$$

Code Criteria:

$(P_L + P_b) \leq 1.35 S_y$ @ RT S_y (@ RT) = 30 ksi (N47)

$(P_L + P_b) \leq 1.35(30) = 40.50$ ksi (N47-3226)

Maximum Stress Combination = $(S_H + S_T) = 16947 \sim (P_L + P_b)$

$(P_L + P_b) = 16947 \leq 40500$ psi ok

Since a majority of the stresses are bending, the primary membrane stresses are minimal:

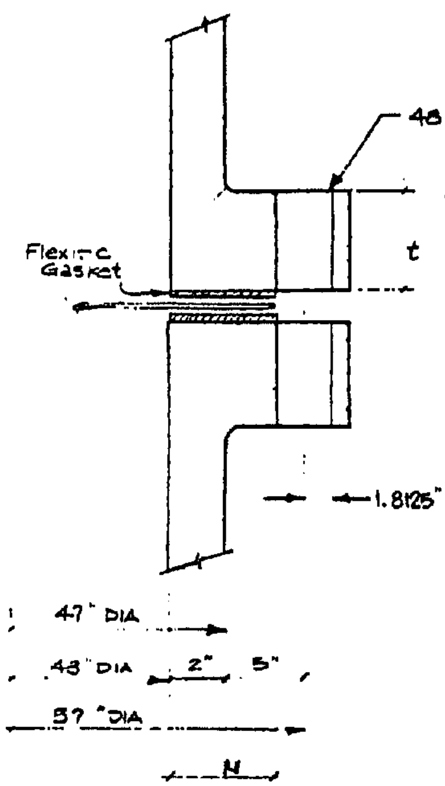
$P_m \leq 0.9\, S_y$ is okay by inspection. (N47-3226)

p

COMPUTATION/RECORD SHEET

Prepared By: J.B. Norel	GENERAL ELECTRIC COMPANY	Page No: 1 of 5
Checked By:	Advanced Reactor Systems Department	Attachment 2
Date: July 15, 1981	TITLE SUPERHEATER FLANGE ANALYSIS	

REF: ASME Code Section III
XI 3200 RF Flanges 1980



Known:

$T = 1015°F$

$P = 880$ psi

Gasket: Spiral wound metal, asbestos filler, stainless steel $m = 3.00$

$y = 4500$ psi (Tbl. XI 3221.1-1)

Bolts 48 - 3 5/8" DIA

$N = (7 - 1.8125) = 5.1875"$

$b_o = \frac{N}{2} = \frac{5.1875}{2} = 2.594" > 1/4"$

$b = \sqrt{\frac{b_o}{2}} = \sqrt{\frac{2.594}{2}} = 0.805"$

$t = 16.30"$

Bolt Load for Operating Conditions.

$G = d_{og} - 2b = 53.38 - 2(0.805) = 51.77$

W_{m1} (minimum bolt load) $= H + H_p = 0.785G^2P + 2b(3.14)Gmp$

$W_{m1} = 0.785(51.77)^2(880) + 2(0.805)(3.14)(51.77)(3)(880)$

$W_{m1} = 2542^k$ (XI 3221.1 Eq. 1)

W_{m2} (req'd bolt load for gasket seating) $= 3.14Gyb$ (XI 3221.2 Eq. 2)

$W_{m2} = 3.14(51.77)(4500)(0.805)$

$W_{m2} = 589^k$

Total Required Actual Bolt Area:

$$A_m = \text{maximum of} \begin{cases} W_{m1}/S_b \\ W_{m2}/S_a \end{cases}$$

Assuming a minimum allowable bolt stress of 30×10^3 psi. (N47 Tbl. I 14.12)

$$\left.\begin{array}{l} A_m = \frac{2542}{30} = 84.73 \\ \text{or} \\ \frac{589}{30} = 19.63 \end{array}\right\} \quad A_m = \frac{84.73 \text{ in}^2}{48 \text{ bolts}} = 1.765 \text{ in}^2/\text{bolt}$$

Using 3 5/8" bolts $A = 10.32 \text{ in}^2$

$\therefore A_b \geq A_m$ ok

p

COMPUTATION/RECORD SHEET

Prepared By: J.E. Nonel	GENERAL ELECTRIC COMPANY	Page No: 2 of 5
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Date: July 15, 1981	TITLE Flange Analysis	

Flange Moments :

Operating Conditions

$M_o = M_D + M_T + M_G$ (XI-3230 (b.))

$= H_D h_D + H_T h_T + H_G h_G$

$= 0.785B^2 p h_D + (H - H_D) h_T + (W - H) h_G$

$p = 880$ psi

$B = 43$ "

$h_D = R + 0.5g_1 = \frac{C-B}{2} - g_1 + 0.5g_1 = \frac{57-43}{2} - 2 + \frac{1}{2}(2) = 6$ "

$h_G = \frac{C-G}{2} = \frac{57-51.77}{2} = 2.62$ " (Tbl. XI 3230-1)

$h_T = \frac{R + g_1 + h_G}{2} = \frac{5+2+2.62}{2} = 4.81$ "

$M_o = 0.785(43)^2(880)(6) + [0.785(51.77)^2(880) - 0.785(43)^2(880)]4.81 +$
$+ [2542000 - 0.785(51.77)^2(880)]2.62$

$M_o = 1.22 \times 10^7$ in-lbs.

Flange Stresses :

Design Loads $T = 1015°F$
$P = 880$ psi

For $2\frac{1}{4}Cr - 1Mo$: Allowable Design Stress $S_o = 5663$ psi @ 1015°F 132000 hrs. (N47)

Code Criteria :

1. Longitudinal Hub Stress $S_H \leq \left.\begin{matrix}1.5S_t \\ 1.5S_m\end{matrix}\right\}$ minimum

$\therefore S_H = \leq 1.5(5663) = 8495$ psi

2. Radial Flange Stress $S_R < S_t = 5663$ psi (XI 3250)
3. Tangential Flange Stress $S_T < S_t = 5663$ psi
4. $\frac{S_H + S_R}{2} \leq S_t$ and $\frac{S_H + S_T}{2} \leq S_t$

p

COMPUTATION/RECORD SHEET

Prepared By: J.B Norel	GENERAL ELECTRIC COMPANY	Page No: 3 of 5
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Date: July 15, 1981	TITLE Flange Analysis	

LONGITUDINAL HUB STRESS

$$S_H = \frac{f M_o}{L g_1^2 B} \qquad (XI\text{-}3240 \;\; Eq. 6)$$

$$\left.\begin{array}{l} g_o = 2 \\ g_1 = 2 \end{array}\right\} \quad \frac{g_o}{g_1} = 1 \quad \therefore \; f = 1 \qquad (Fig. \; XI \; 3240\text{-}6)$$

$$L = \frac{te + 1}{T} + \frac{t^3}{d} = 11.731$$

$$K = \frac{A}{B} = \frac{61}{43} = 1.42$$

Assuming $A = 61''$

$$\left.\begin{array}{l} U = 6.27 \\ V = 0.550103 \end{array}\right\} (Fig \; XI\,3240\text{-}1 \text{ and } XI\,3240\text{-}3)$$

$$T = 1.75 \qquad (Fig. \; XI\,3240\text{-}1)$$

$$d = \frac{U}{V} h_o g_o^2 = \frac{6.27}{0.550103}(9.27)(2)^2 = 422.63$$

$$h_o = \sqrt{Bg_o} = \sqrt{43(2)} = 9.27$$

$$F = 0.908920 \qquad (Fig. \; XI\,3240\text{-}2)$$

$$e = \frac{F}{h_o} = \frac{0.908920}{9.27} = 0.098$$

$$\frac{f M_o}{\left[\frac{te+1}{T} + \frac{t^3}{d}\right] g_1^2 B} = S_H \leq 8495 \text{ psi}$$

$$\frac{1.00\,(1.22 \times 10^7)}{\left[\frac{t(0.098)+1}{1.75} + \frac{t^3}{422.63}\right](2)^2(43)} \leq 8495$$

$$S_H = \frac{1.00(1.22 \times 10^7)}{11.731\,(2.00)^2(43)^2}$$

$$S_H = 6046 \text{ psi}$$

p

COMPUTATION/RECORD SHEET

Prepared By: JBH	GENERAL ELECTRIC COMPANY	Page No: 4/5
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Date: 8/24/91	TITLE Flange Analysis	

RADIAL FLANGE STRESS

$$S_R = \frac{(1.33te + 1)M_o}{Lt^2B} \qquad (\text{Fig. XI 3240 Eq. 7})$$

$$S_R = \frac{(1.33(16.30)(0.098)+1)(1.22 \times 10^7)}{(11.731)(16.30)^2(43)}$$

$$S_R = 284 \text{ psi}$$

TANGENTIAL FLANGE STRESS

$$S_T = \frac{YM_o}{t^2B} - ZS_R \qquad (\text{XI 3240 Eq. 8})$$

$$Y = 5.70 \qquad Z = 2.97 \qquad (\text{Fig. XI-3240-1})$$

$$S_T = \frac{5.70(1.22 \times 10^7)}{(16.30)^2(43)} - 2.97(284)$$

$$S_T = 5242 \text{ psi}$$

	Actual Stress	Allowable Code Stress	Status
S_H	6046 psi	8495 psi	Pass
S_R	284	5663	Pass
S_T	5242	5663	Pass
$\left(\frac{S_H+S_R}{2}\right)$	3165	5663	Pass
$\left(\frac{S_H+S_T}{2}\right)$	5644	5663	Pass

The given stresses are produced by the minimum flange thickness to satisfy code.

COMPUTATION/RECORD SHEET

Prepared By: JBW	GENERAL ELECTRIC COMPANY	Page No: 5/5
Checked By:	Advanced Reactor Systems Department	
Date: 8/28/81	TITLE Flange Analysis	

HYDROSTATIC TEST

$P_{Test} = 1.50\, P_{design}$ (XII - 1000 (d.))

$P_{test} = 1.50(880)$

$P_{test} = 1320$ psi

T_{test} = room temperature

Bolt Load @ Test:

$W = 0.785(51.77)^2(1320) + 2(0.805)(3.14)(51.77)(3)(1320)$

$W = 3814^K$

Flange Moments:

$M_o = M_D + M_T + M_G$

$M_o = 0.785(48)^2(1320)(6) + [0.785(51.77)^2(1320) - 0.785(48)^2(1320)]\, 4.81 +$

$[3814000 - 0.785(51.77)^2(1320)]\, 2.62$

$M_o = 1.836 \times 10^7$ in-lbs

Flange Stresses:

$$S_H = \frac{1.00(1.836 \times 10^7)}{11.731(2.00)^2(45)} = 3099 \text{ psi}$$

$$S_R = \frac{(1.33(16.3)(0.098)+1)(1.836 \times 10^7)}{(11.731)(16.5)^2(45)} = 428 \text{ psi}$$

$$S_T = \frac{5.70(1.836 \times 10^7)}{(16.30)^2(49)} - 2.97(428) = 7889 \text{ psi}$$

Code Criteria:

$(P_L + P_b) \leq 1.35\, S_y$ @ RT S_y (@ RT) = 30.0 ksi (N47)

$(P_L + P_b) \leq 1.35(30) = 40.50$ ksi

Maximum Stress Combination = $(S_H + S_T) = (P_L + P_b) = 16988$ psi

$(P_L + P_b) = 16.988 \leq 40.50$ ksi ok

A majority of the stresses are bending therefore the primary membrane stress is minimal:

$P_m \leq 0.9\, S_y$ is okay by inspection. (N47-3226)

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Enclosure (2)

APPENDIX A9

BECHTEL JOB 14953

FOR

GENERAL ELECTRIC COMPANY

ADVANCED REACTOR SYSTEMS DEPARTMENT

METHANATOR PLANTS

FOR

THE HTGR - REFORMER

THERMOCHEMICAL PIPELINE SYSTEM

Conceptual Cost

Estimate and Scheaule

Ecomomic Evaluation
September 1981

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JOB 14953
GENERAL ELECTRIC COMPANY
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PROJECT DATA

CLIENT: General Electric Company

LOCATION: Hypothetical site near Willmington, Delaware

SCOPE: Conceptual construction cost estimate and schedule for the engineering, procurement ano construction of methanator plant with 40 MWT capacity.

As alternative cases, the construction cost of 80 and 180 MWT facilities also evaluated and also the concurrent construction of multiple 40 MWT units.

PURPOSE OF ESTIMATE: To provide the client a conceptual construction cost estimate and engineering, procurement, and construction (EPC) schedule for a 40, 80, and 180 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: After notice to proceed (NTP) foı 40 MWT Plant(s).

	MONTHS AFTER NTP
NTP and Start of Engineering	0
Start Construction	7
Control Building Complete	14
Start Plant Testing	17
Complete Construction	20

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SECTION 1

INTRODUCTION

1.1 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 or more methanator plants of various 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 either to industrial process steam or to directly heat process streams, e.g., the retorting gas required in the oil shale process.

1.2 TECHNICAL SCOPE

The scope of this study is limited to the methanator plants. The thermal capacity of the several plants investigated and their application is as follows:

CASE	CAPACITY	DESCRIPTION
1 (base)	40 MWT	Plant designed to generate steam at 532° F, 905 psig.
2	80 MWT	Plant designed to raise the temperature of retorting gas from 216° F to 1,300° F for oil shale application.

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CASE	CAPACITY	DESCRIPTION
3	80 MWT	Identical to Case 1 except equipment and systems have been sized larger to provide a capacity of 80 MWT.
4	80 MWT	Two identical trains similar to Case 1.
5	180 MWT	Identical to Case 1 except equipment and systems have been sized larger to provide a capacity of 180 MWT.
6	560 MWT	Fourteen identical 40 MWT methanator plants matching the total capacity of the reformer: 560 MWT. The estimate has been prepared on the assumption that these plants are to be built simultaneously and so located that construction facilities can be shared.

The construction cost of the base case (40 MWT) has been estimated and its basis is included in this report. The other cases for the 80 MWT plants and 180 MWT plant, have been scaled based on the 40 MWT plant and the major equipment cost provided by G. E.

For Case 6, certain economies are realized through duplication, i.e., reduced engineering and construction indirects, and purchasing discounts.

A brief description of the base case is as follows:

The syngas is delivered to the combination methanator/evaporator at 651°F, 898 psig after it passes through the desulfurizer for residual sulfur cleanup. The product gas from the evaporator is then sent to the packedbed type adiabatic methanators 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 feed water 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 psig generated from the evaporator is sent to the steam orum where the saturated steam 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.

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CONSTRUCTION COST ESTIMATE

The construction cost of the base case methanator plant (40MWT) has been estimated and included in this report based on the following brief discussion:

2.1 ESTIMATE BASES

The estimates are based on the conceptual design and engineering information prepared by G.E. for the study in the form of engineering drawings, outline specifications, and equipment lists. Estimating methods consistent with the conceputal 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 G.E. 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 responsibility for engineering, procurement, and construction (EPC).

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2.3.1 Direct Field Cost

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

Mechanical Equipment

The mechanical equipment estimate is based on an equipment list - see Enclosure 1. The cost of major equipment including combination of methanator and evaporator, adiabatic methanator, steam drum, heat exchanger and catalyst were provided by G.E. These costs include field installation and appropriate contingency. These major equipment costs represent approximately 50 percent of all equipment cost. The cost of other equipment was obtained through direct vendor contacts, they include:

- Flare Stack
- 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

The piping requirements for the process has been estimated based on quantities derived from engineering diagrams and sketches.

The piping requirements for the utility system have been estimated based on discussion with engineers, and experience with similar facilities.

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The resulting quanties are summarized below:

- Process Piping

10" & Under, Alloy (A/G*)	4,770 LF
8" & Under, C/S (A/G*)	4,140 LF

- Utility Piping

8" Under, C/S (A/G*)	10,650 LF
6" Under, C/S (U/G**)	2,200 LF
8" Under, VCP (U/G**)	800 LF

*A/G: Above Ground **U/G: Under Ground

The piping estimate includes the cost of fittings, insulation, hangers, supports, testing and freight.

Material pricing was based on informal vendor contacts, recent Bechtel experience, vendor catalogues and national pricing bulletins.

Other Bulk Materials

Other bulk materials include the following:

- Civil/Structural
- 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.

Subcontracts

Subcontracts are not included as such in the cost estimate. Components and systems that normally would be estimated as single subcontract cost entries (e.g., mass earthwork, roofing and siding, paving, fencing, stacks, and insulation) are divided into, and entered as, direct materials

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and direct labor costs. This procedure was adopted to ensure a comprehensive accounting of all field labor manhours. An accurate estimate of field manhours is required for determining the construction schedule. Labor that is normally subcontracted cannot be identified if its cost is included with material costs, overhead costs, and profit in a single subcontract cost entry. This method, which in effect assumes all field labor is performed by the prime contractor's work forces, results in essentially the same total estimated construction cost as that based on some subcontract work.

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.

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.

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

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- 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 12 percent of field construction costs has been used.

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 G.E. based on G.E.'s statement that these costs include appropriate contingency.

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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 power will be adequate for the requirements of this project.
- Construction site is flat and nominal effort is requireo 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

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2.8 ESTIMATE BASIS FOR OTHER CASES

The estimate basis for Case 1 (40 MWT) has been described above. The estimate basis for the other five cases is briefly described below:

CASE	CAPACITY	ESTIMATE BASIS
2	80 MWT	Estimated based on the assumption that all supporting equipment and bulks are identical with Case 3; however, methanation equipment costs have been replaced with the costs provided by G.E.
3	80 MWT	Scaled from Case 1 by applying a 0.75 power factor to the change in plant capacity, and assuming the design scope is a larger 80 MWT plant.
4	80 MWT	Scaled from Case 1 by applying 0.95 power factor to the change in plant capacity, and assuming the design scope is two 40 MWT plants.
5	180 MWT	Scaled from Case 1 by applying a 0.75 power factor to the change in plant capacity, and assuming the design scope is a larger 180 MWT plant. G.E. also provided costs for the methanation equipment.
6	560 MWT	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 . . . 20 percent and 10 percent discount for equipment and bulk material cost respectively have been included in the estimate to reflect larger quantity procurment. • 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.

CASE	CAPACITY	ESTIMATE BASIS
		• 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 on some plants that can be utilized on other plants.

2.9 ESTIMATE TABLES AND FIGURE

2.9.1 Estimate Tables

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

Table 2-1	Construction Cost Summary for Single 40 MWT Plant	CASE (1)
Table 2-2	Equipment Cost Summary for Methanation System (40 MWT)	CASE (1)
Table 2-3	Equipment Cost Summary for Supporting System (40 MWT)	CASE (1)
Table 2-4	Construction Cost Comparison (40, 80, 180 and 560 MWT)	

2.9.2 Figure

The result of the common based estimate for Cases 1, 4 and 6 was plotted on log-log coordinates as shown on the following figure:

Figure 2-1 Total cost showing the effect of the number of 40 MWT Plants.

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Table 2-1

CONSTRUCTION COST SUMMARY
FOR
40 MWT METHANATOR PLANT - CASE 1

($ In 1,000's)

	Equipment & Materials	Labor	TOTAL
METHANATOR PLANT SYSTEM, EQUIPMENT	3,470	110	3,580 [1]
SUPPORTING SYSTEM, EQUIPMENT	1,470	120	1,590 [2]
PIPING	1,070	480	1,550 [3]
CIVIL STRUCTURAL	1,300 [4]	1,250	2,550
ELECTRICAL	630 [5]	600	1,230 [6]
INSTRUMENTATION	630 [7]	400	1,030
TOTAL DIRECT COST	8,570	2,960	11,530
INDIRECT FIELD COST			2,470
TOTAL FIELD COST			14,000
ENGINEERING SERVICES & FEE			1,700
SUBTOTAL			15,700
CONTINGENCY			3,100
TOTAL CONSTRUCTION COST (Price & Wage Level, 1Q 1980)			18,800

1 For Detail Breakdown, See Table 2-2
2 For Detail Breakdown, See Table 2-3
3 Estimate based on process piping quantity takeoff and allowance for utility piping.
4 Estimate based on the mean percent derived from a 58 job study with a range of 10-57% of equipment cost.
5 Estimate based on the mean in a range of 6-18% of equipment cost.
6 Estimate based on a $/KW basis for less than a 5 MW plant. The estimated cost range is $800-1,200/KW.
7 Estimate based on the mean in a range of 7-20% of equipment cost.

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Table 2-2

EQUIPMENT COST SUMMARY
FOR
METHANATOR PLANT SYSTEM (40 MWT) - CASE 1

($ In 1,000's)

	Number	TOTAL**
DESULFURIZER*	2	101.5
COMBINATION METHANATOR PLANT/EVAP.*	1	620.3
ADIABATIC METHANATOR*	2	339.5
PROCESS STEAM SUPERHEATER*	1	102.6
SYNGAS HEATER*	1	93.0
FEED WATER HEATER*	3	240.1
STEAM DRUM*	1	134.4
AIR COOLER*	1	35.0
CATALYST*	-	824.8
RECYCLE GAS COMPRESSOR	3	527.4
OTHER EQUIPMENT	10 ***	561.4
TOTAL DIRECT COST (Price & Wage Level, 1Q 1980)	25	3,580.0

* Cost provided by G.E.

** Total direct cost including material and labor.

*** Breakdown of number is as follows:

Condensate storage tank	1
K.O. Drum	1
Deaerator	1
Product Gas Dryer	2
Steam Drum Condensate Recycle Dump	2
Feed Water Pump	3
TOTAL	10

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Table 2-3

EQUIPMENT COST SUMMARY
FOR
SUPPORTING SYSTEM (40 MWT) - CASE 1

($ In 1,000's)

	Number	TOTAL*
STARTUP HEATER	2	400.9
FEED WATER STARTUP HEATER	1	204.0
STARTUP NITROGEN STORAGE	1	262.5
WATER TREATMENT PLANT	1	138.0
OTHER EQUIPMENT	2 **	584.6
TOTAL DIRECT COST	7	1,590.0
(Price & Wage Level, 1Q80)		

* Total direct cost including material and labor.

** Breakdown of number is as follows:

K.O. Pot with Flare Stack	1
Emergency Power Supply System	1
TOTAL	2

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Table 2-4

CONSTRUCTION COST COMPARISON

($ In 1,000's)

STUDY CASE	1	2	3	4	5	6
CAPACITY, MWT	40	80	80	80	180	*
EQUIPMENT	5,170	7,030	8,690	9,980	16,000	4,200
PIPING	1,550	2,600	2,600	2,990	4,800	1,490
CIVIL STRUCTURAL	2,550	4,300	4,300	4,900	7,900	2,550
ELECTRIC	1,230	2,070	2,070	2,370	3,800	1,230
INSTRUMENTATION	1,030	1,730	1,730	1,990	3,200	1,010
TOTAL DIRECT COST	11,530	17,730	19,390	22,230	35,700	10,480
INDIRECT FIELD COST	2,470	3,790	4,150	4,750	7,600	2,280
TOTAL FIELD COST	14,000	21,520	23,540	26,980	43,300	12,760
ENGINEERING & FEE	1,700	2,580	2,860	3,300	5,200	890
SUBTOTAL	15,700	24,100	26,400	30,280	48,500	13,650
CONTINGENCY	3,100	4,800	5,200	5,920	9,700	2,350
TOTAL CONSTRUCTION COST (Price & Wage Level, 1Q80)	18,800	28,900	31,600	36,200	58,200	16,000

* The cost represents that for one 40 MWT plant when 14 such plants are to be built simultaneously at locations so that construction facilities can be shared.

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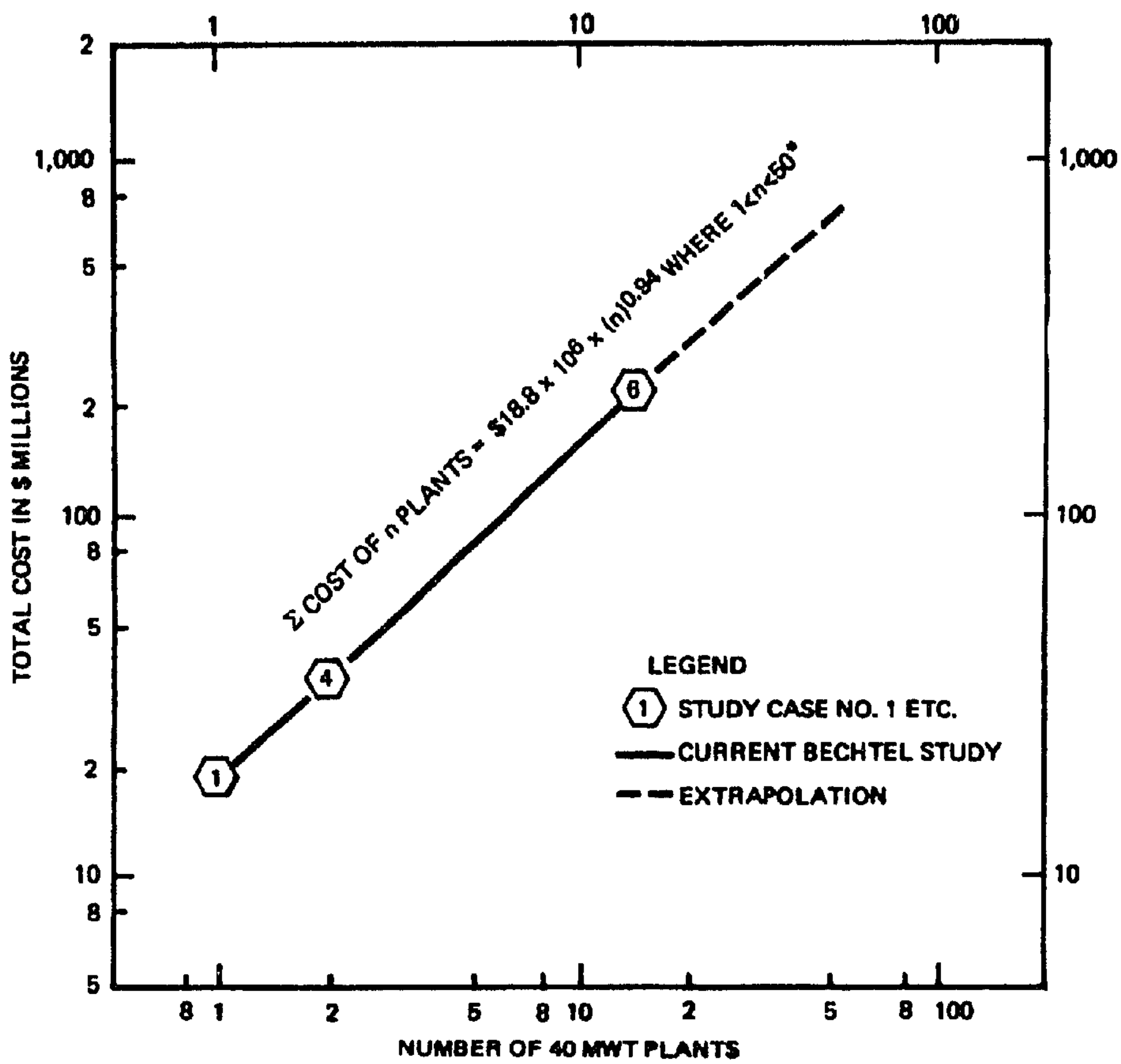

*IT HAS BEEN JUDGED THE CORRELATION CAN BE EXTRAPOLATED UP TO n=50 WHERE THE CONSTRUCTION EFFICIENCY CAN BE REASONABLY MAINTAINED. HOWEVER, FURTHER VERIFICATION IS NEEDED.

Figure 2-1 TOTAL COST SHOWING THE EFFECT OF THE NUMBER OF 40 MWT METHANATION PLANTS

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SECTION 3

ENGINEERING, PROCUREMENT AND CONSTRUCTION SCHEDULE

3.1 GENERAL

The construction sequence and methods used for commercially developed process plants generally follow an approach which permits procurement and construction to commence before engineering is complete thus shortening the overall schedule.

The milestones for the 40 MWT Methanation Plant are as follows:

	MONTHS AFTER NTP
Notice to Proceed (NTP) and Start of Engineering	0
Start Construction	7
Control Building Complete	14
Start Plant Testing	17
Complete Construction	20

The schedule is developed by using a months after Notice to Proceed concept, as requested by General Electric. The elapsed times are based on Bechtel experience for process units of similar size and complexity.

3.2 CRITICAL PATH

The critical path for the project is as follows:

- Engineering
- Procurement (Switchgear, Water Treatment, And Combined Methanator/Evaporator No.1)

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- Switchgear Building
- Waste Water Treatment Building
- Methanator System
- Complete Construction
- Testing

As this schedule is of a conceptual nature, a detailed evaluation of the critical path cannot be developed until near the end of preliminary engineering when the key project documents (i.e., Plot Plans, P&ID's, Single Lines, etc.) are available.

3.3 QUALIFICATIONS/ASSUMPTIONS

- It is assumed that the preparation and review of the environmental reports will not affect the schedule
- The owner may issue pruchase orders for long-lead items prior to the award of the construction contract
- The schedule of field activities is predicated on a normal 5 day, 40 hour work week with two-shift operations allowed when the work schedule is tight
- A source of craft labor is available in the vacinity in amounts required to meet the construction manpower demands
- Support utilities will be available during the initial phases of construction
- Off site preassembly and modularzation may improve the construction duration shown on the EPC schedule

3.4 EPC SCHEDULE

The above discussion forms the bases for the EPC schedule shown on Figure 3-1.

Figure 3-1 Methanator Plant (40 MWT) - EPC Milestone schedule.

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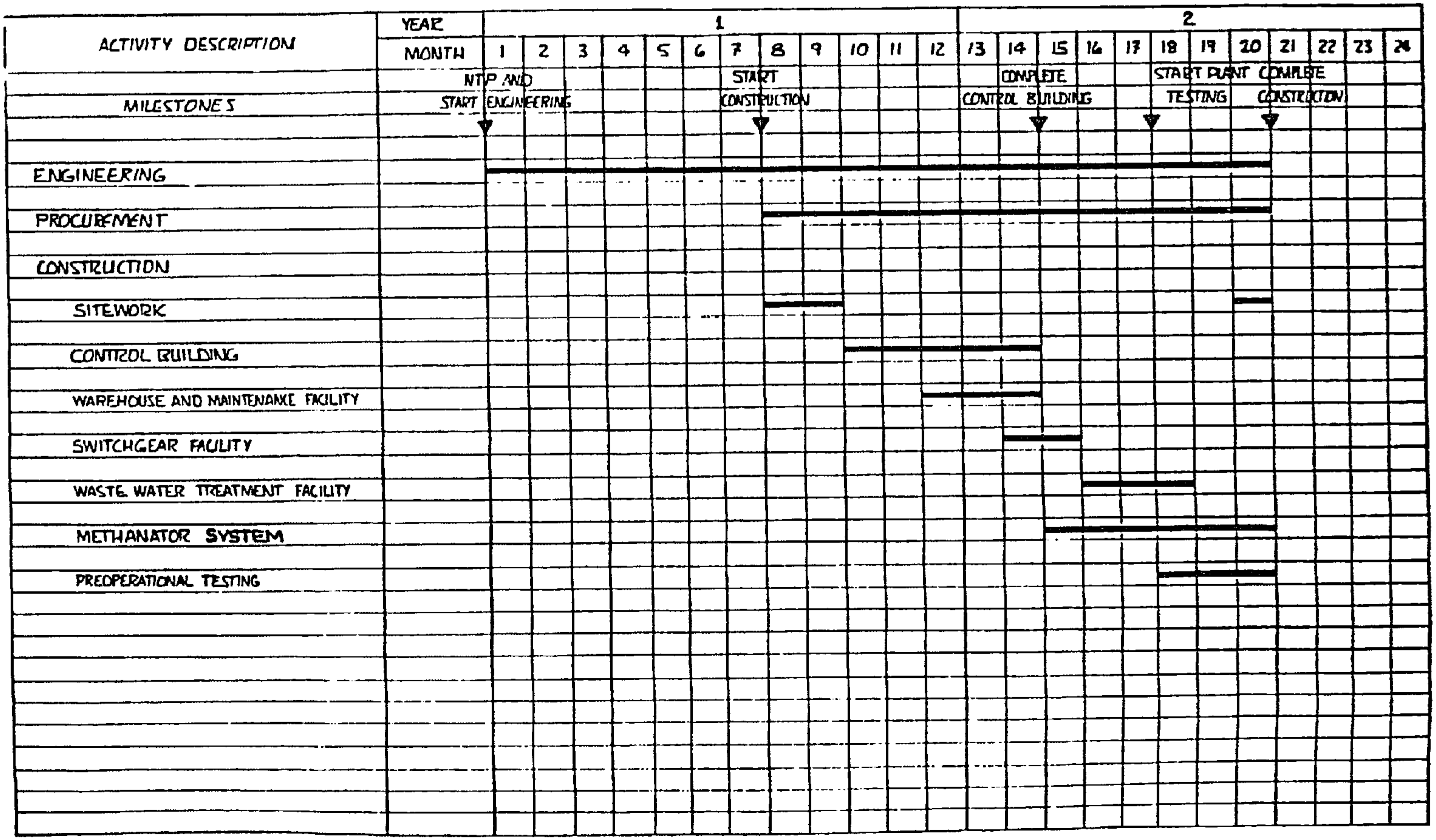

FIGURE 3-1. METHANATION PLANT — 40 MWTH
GENERAL ELECTRIC COMPANY
EPC MILESTONE SCHEDULE

40 MWt METHANATION PLANT

MAJOR EQUIPMENT

METHANATION SYSTEM	EQUIPMENT DESCRIPTION
Desulfurizer No. 1	Provided by GE/ARSD
Desulfurizer No. 2	Provided by GE/ARSD
Methanator/Evaporator No. 1	Provided by GE/ARSD
Methanator No. 2	Provided by GE/ARSD
Methanator No. 3	Provided by GE/ARSD
Syngas Preheater	Provided by GE/ARSD
Process Steam Super Heater	Provided by GE/ARSD
Feed Water Heater No. 1	Provided by GE/ARSD
Feed Water Heater No. 2	Provided by GE/ARSD
Feed Water Heater No. 3	Provided by GE/ARSD
Air Cooler	Provided by GE/ARSD
Product Gas Dryer	
Type (Sivalls Model DHT-30-250-210)	TEG-Scrubbing package unit with absorber and regenerator
Operating Temp. & Press	100°F/870 psia
Outlet Moisture	7 lbs $H_2O/10^6$ scf
Reboiler Duty	100,000 Btu/hr
Dimensions of Each Unit	Absorber: 2 1/2 ft(Dia.) X 15 ft(V.H.) Regenerator: 2 ft(Dia.) X 7 ft(V.H.)
No. of Units	1+1 (100% spare)
Steam Drum	Provided by GE/ARSD
K.O. Drum	
Temp/Press	100°F/870 psia
Dimension	27 in.(Dia.) X 132 in.(V.H.)
No. of units	1
Steam Drum Condensate Recycle Pump	
Capacity/HP	80 gpm/5 HP each
Press in/out	870/900 psia
Dimension of each unit	24 in.(L) X 18 in.(W)
No. of Units	1+1 (100% spare)

40 MWt METHANATION PLANT

MAJOR EQUIPMENT (Cont'd)

METHANATION SYSTEM	EQUIPMENT DESCRIPTION
Feed Water Pump	
Capacity/HP	210 gpm/160 HP each
Press in/out	20/925 psia
Dimension of each unit	Vertical turbine pump 2 ft.(Dia.) X 4 ft.(V.H.)
No. of Units	2+1 (50% spare)
Deaerator	
Capacity/Press	3000 gallons/250 psia
Dimension of each unit	4 1/2 ft.(Dia.) X 24 ft.(V.H.) and 1/4 in. wall thickness
No. of Units	1
Recycle Gas Compressor	
Type	Centrifugal, Elliott 29M or equivalent
Temp. and Press (in)	651°F/878 psia
Temp. and Press (out)	668°F/918 psia
Polytropic Head	4683 ft - lb/lb
No. Stage	1
Horse Power	210 each
Dimension of Each Unit	61 in.(H) X 57 in.(W) X 50 in.(L)
No. of Units	2+1 (50% spare)

SUPPORT SYSTEM

Start-up Gas Heater No. 1	
Heat Duty	5 X 10^6Btu/hr, gas fired
Dimension of Each Unit	8 ft.(Dia.) X 20 ft.(V.H.)
No. of Units	1
Start-up Gas Heater No. 2	
Heat Duty	15 X 10^6Btu/hr, gas fired
Dimension of Each Unit	12 ft.(Dia.) X 30 ft.(V.H.)
No. of Units	1
Feed Water Start-up Heater	
Heat Duty	10 X 10^6Btu/hr, gas fired
Dimension of Each Unit	9 ft.(Dia.) X 28 ft.(V.H.)
No. of Units	1

40 MWt METHANATION PLANT

MAJOR EQUIPMENT (Cont'd)

SUPPORT SYSTEM	EQUIPMENT DESCRIPTION
Start-up Nitrogen Storage	Package unit with necessary recycle pump and controls.
Holding Capacity	1 X 10^6SCF Permanent storage tank
Dimension of Unit	9 ft.(Dia.) X 40 ft.(V.H.)
No. of Units	1
Emergency Power Supply System	
Type (Cummines-West Co., Model #NT-855-GS-230 KW)	230 KW Diesel generator, Package unit, fuel use 150 gal/day
Dimension	40 in.(W) X 60 in.(H) X 120 in.(L)
K.O. Drum with Flare Stack	
Type (National Air Oil Burner Co. Mod.# NAO-NIP-IM-2A-E)	Package unit self-supporting with 2 pilot ignitors, pilot gas 160 scfh.
Capacity	350 cfm
Temp/Press	750°F/1000 psia
Dimension of each unit	Base K.O. Drum 4 ft.(Dia.) X 9 ft.(V.H.) Stack 14 in.(Dia.) X 32 ft. (V.H.)
No. of Units	1
Condensate Storage Tank	
Capacity	2000 gallons
Temp/Press	300°F/900 psia
Dimension of each unit	50 in.(Dia.) X 24 ft.(V.H.) & 1/2 in. wall thickness, Low-Ni alloy steel.
No. of Units	1