E. PRODUCT STUDY: METHANOL VS METHANOL AND AMMONIA

It was proposed by the co-sponsors of this TDP study that project technical, financial and commercial viability could well rest on the decision to either co-produce methanol and ammonia, or, to produce only methanol. This subject had been addressed previously, in 1984, in the course of a plantship feasibility study for the Department of Transportation. A definite synergism for co-production was found to exist under the following conditions:

a) When methanol is produced via conventional reforming.

b) When both the methanol and ammonia plants are land-based and in proximity one to the other.

c) When world prices for methanol and ammonia approximate those of 1984, i.e., when good margins on ammonia sales can compensate poor margins on methanol.

The foregoing technical and marketing issues were re-visited to assure that this TDP study was premised on the appropriate product/product mix.

E.1 - Technical Considerations

When conventional reforming is the route to methanol production a by-product of methanol production is an excess of hydrogen gas; this is recycled to the reformer and burned to assist production of syngas from natural gas and steam. Since hydrogen is a required feedstock in ammonia production, an alternative to combusting it in the reformer would be to supply it at near zero cost to an ammonia plant. In the DOT study, which addressed specifically the feasibility of a methanol plantship, it was found that plantship feasibility was improved by selection of a catalytic auto-thermal partial oxidation route to syngas production rather than the conventional reforming route. This has the consequence of reducing availability of low-cost by-product hydrogen for feedstock to an ammonia plant. A review of the basis for selection of CAT-POX over conventional reforming sustained this decision; this does not preclude co-production of methanol and ammonia on the plantship but it marginally increases the cost of ammonia production, perhaps $5 to $10 per ton. A further serious consideration is refrigerated ammonia storage. Discussions with the U.S. Coast Guard to obtain concurrence in the design of the catalytic auto-thermal system (which employs a liquid oxygen production plant and associated cryogenic storage/piping subsystems) provided guidance respecting consideration of on-board storage of liquid ammonia (-28°F); it is just not feasible. Ammonia storage would necessarily be provided aboard a second moored vessel; ammonia would be pumped from the plantship to the storage vessel via pipelines on the seafloor and riser connections therefrom at each the plantship and the storage vessel. At this point not only has the operating cost advantage of low-cost hydrogen availability been lost but significant added capital expense is represented by the ammonia transfer and storage facilities. A further operating cost increase is attendant refrigerated ammonia storage on the storage vessel; a refrigeration system is required at an operating expense of $5 to 10/ ton of produced ammonia (this is in addition to a like cost for storage at the point of ammonia delivery).

In summary, both capital and operating costs are increased when co-production is entertained. To meet technical considerations for safe operation, a capital cost increment rather than a decrement exists for storage when co-production of methanol occurs aboard ship rather than ashore, also, co-production via CAT-POX aboard ship incurs a $15 to $30 operating cost penalty for ammonia as compared to co-production via conventional reforming shore-side.

Finally, even should free natural gas be available at sea, it is difficult to postulate methanol/ammonia market price differentials that could justify the technical complexity of a plantship co-production facility.

E.2 - Ammonia Market Considerations

Ammonia production and sales over the next several years were reviewed with industry sources. Import and export data on ammonia and competing nitrogenous products were examined, also, the market prices they command (see attachment which follows).

Most industry leaders were found to think that ammonia production and marketing will not be attractive in the next few years, unless one has access to extremely low cost natural gas, i.e., less than $0.50/MMBTU at the burner tip. Currently, June 1987, ammonia sells at $100/110 per short ton FOB Gulf Coast terminals. With gas conversion costs at land-based plants conservatively estimated at $20/ton (as noted in paragraph F1, plantship ammonia would be higher), one needs a very low gas cost to have any return on the plant investment based on the current ammonia market price.

There has been a recent upturn in ammonia prices. This is in part due, however, to a spate of U.S. plant closings; ammonia profits have been so poor that eleven (11) domestic plants have shutdown since 1984. One reason for the poor profits on domestic production is the pressure on selling prices that has been created by the large volume of ammonia imports from Mexico, Russia and Trinidad, specifically:

1984 (1)	3,259,000 ST
1985	2,956,000 ST
1986	2,814,000 ST
1987	2,054,000 ST

(1) June through June excepting 1987 which is June 1986 through April 1987.

A second source of pressure on ammonia prices has been the substantial import of other nitrogenous products, particularly the ammonia derivative, urea. Urea import prices have been so low that the U.S. Trade Commission has recently ruled that Russia, Romania and East Germany are dumping in the United States.

Finally, the recent improvement in methanol market prices (final quarter 1987 through first half 1988) further argues against a co-production facility; unlike the period in 1984 when the selling price differential between ammonia and methanol made it prudent to consider co-production to support methanol sales, it is difficult to project a recurrence of such a situation.

Like the technical considerations, marketing considerations result in a strong recommendation that the plantship only produce one product, and that be methanol.

F. PLANTSHIP STUDY: CONVERSION VS PURPOSE-BUILT

F.1 VLCC Conversion vs. New Construction General Considerations

It was not a purpose of the TDP study to reopen a comparison of all the possible plantvessel configurations. The DOT study decision that a single-hulled vessel with product storage onboard represented the best solution was based on many factors, most of which are still valid or even more telling than before. The basis for preference of a purpose built hull to a VLCC conversion may however have changed; this decision is reviewed herein. Before any quantitative comparisions are made however,the following qualitative and perceptive considerations can be listed:

VLCC Conversion

Advantages

a. Savings in initial steel cost by using a large portion of the VLCC structure which is bought at nominally "scrap" price.

b. Possibility of using existing asset should an equity partner contribute from inventory a useable VLCC.

c. Possible reduction in hull construction time.

Disadvantages

a. Difficulty in assessing the basic price before actually buying a vessel due to the changeability of the market.

b. Difficulty in estimating the cost of conversion without a prior detailed and expensive survey of the condition of hull and other parts intended for retention.

c. Original design objectives for a large tanker are at odds with requirements for a plantvessel, specifically:

- Carriage of a maximum amount of liquid cargo rather that support of heavy loads; and
- Design for minimum initial cost and an intended life of only 20 years results in light steel work and no special coatings.

New Construction

Advantages

a. Design form will be configured for minimum size and best utilization of capacity.

b. Ability to use broader range of hull forms or proportions for motion reduction/de-tuning.

c. Freedom to design structure for least cost, best adaptation to local loads and for maximum life without frequent recourse to repair yards.

Disadvantages

a. Higher cost for basic steel shell

b. A possible longer period required for system construction/outfitting/ deployment.

F.2 VLCC Market Conditions

The market price for VLCCs has increased dramatically in the past 18 months. Vessels that would have been easily available for $5-8 million in 1985 are now being offered for $12-15 million, this though the vessel is now older and less desirable for conversion purposes.

The Figure G-1 to G-4 graphs have been prepared to show the number of VLCCs built during previous years and the number of vessels that will be in a given age group in succeeding years. (No allowance for Persian Gulf losses or for future vessel scrappings has been included.) Two conclusions emerge from this data:

a. The building of large numbers of VLCCs was a very short-lived phenomenon.

b. Within the next five years, VLCCs aged five to eight years, those considered ideal for conversion, will become very scarce; and presumably very expensive.

The scrap rate for VLCCs has also been rising. Recent quotes have been in the range of $115 to $140 per L. Ton, a significant increase from the $85 prices of 2 years ago. The scrap price indicates the "floor" for any conversion purchase. From all this the conclusion can be drawn that a conversion would have to be done with a somewhat older vessel than is preferred, and the initial price of the vessel will be higher.

F.3 VLCC Conversion vs. New Construction Estimated Cost Comparison

Table G-1 shows a comparison of the probable cost ranges for a VLCC conversion and for new construction. The cost numbers are based on the basic values developed in the DOT system study and on preliminary estimates at 7/87 for this site specific study, all as factored to suit the new VLCC pricing noted previously.

The following notes apply to the table:

1. The cost of new construction assumes that the plant vessel can be downsized slightly to a 39,000 L. ton lightship weight from the current 43,000 L. tons. Lowest cost assumes Korean hull construction and Gulf Coast outfitting; labor is costed at 125% of the DOT study rate.

2. VLCC characteristics are for the "Massachusetts" currently for sale by MarAd. Approximate dimensions are:

DWT	264,000
Length	1,060 Ft.
Beam	178 Ft.
Hull Depth	86 Ft.
Year Built	1975

3. For the conversion cases, the lowest cost case assumes that the work will be done in Portugal at a gross cost of $1.35 per lb. while the highest cost case assumes U.S. work at $2.50 per lb. Towing costs are included appropriately for each the new construction and conversion cases.

4. Basis outfitting costs are the same as for the DOT system with the noted 125% factor for labor rate; an override 15% additional labor is believed appropriate for the conversion cases and has been applied.

5. The basic USA cost is from the DOT estimate. An additional factor is used for the added cost of installation on a conversion; 10% for the low cost case and 25% for the high cost case.

F.4 CONCLUSION

A purpose built hull remains preferred to a VLCC conversion, on economic, technology and operating bases.

G. PROCESS PLANT DESIGN DETAIL

1. PROCESS DESIGN BASIS

2. TECHNICAL INFORMATION

   A.) Process Description
   B.) Raw Material & Utilities Imported
   C.) Product & Waste Streams Exported
   D.) Electrical Power Summary
   E.) Equipment List

3. DISCUSSION

   A.) Catalytic Partial Oxidation (CPO)
   B.) Tube Cooled Methanol Converter (TCC)

4. PROCESS FLOW DIAGRAM

5. DESIGN ENGINEERING

SECTION 1

PROCESS DESIGN BASIS

A. PLANT CAPACITY

The plant has been designed to produce 3000 short tons per day of methanol.

B. RAW MATERIALS AND UTILITIES SPECIFICATIONS

1. Natural Gas

Component	% Mole
Methane	99.83
Ethane	0.14
Propane	0.03
Sulfur	Trace
Avg. Molecular Weight	16.071
LHV, Btu/SCF	910.51
HHV, Btu/SCF	1011.21
Supply Pressure	500 psig
Supply Temperature	60°F

2. Oxygen (Internally Generated)

Component	% Mole
Oxygen	99.5
Argon	0.5
Others	Trace
Pressure/Temp.	430 psig/ambient

3. Nitrogen (Internally Generated)

Component	% Mole
Nitrogen	99.9
Oxygen	<0.1
Others	Trace
Pressure/Temp.	100-140 psig/ambient

4. Sea Water (Once Thru)

Surface Water Temp.	82°F
Draw Water Temp.	80°F
Return Water Temp. (Max.)	100°F
Supply Pressure	40 psig
Design Pressure	75 psig

5. Cooling Water (Desalinated Closed Circuit)

Supply Temperature	90°F
Return Temperature (max.)	120°F
Supply Pressure	60 psig
Design Pressure	75 psig

6. Power Generation (Internally Generated)

Via three Gas Fired Diesel Engines	15 Megawatts (4160 V)

7. Steam Generation (Internally Generated)

Pressure	775 psig
Temperature	775°F
Rate	500,000 lb/hr

8. Steam Levels (within Plant)

Level 1	750 psig/750°F
Level 2	60 psig/307°F

9. Ambient Air

Pressure	14.7 psia
Temperature	90°F
Relative Humidity	86%

10. Plant Air

Operating Pressure	60 psig
Design Pressure	100 psig
Temperature	Ambient

11. Instrument Air

Operating Pressure	60 psig
Design Pressure	100 psig
Temperature	Ambient
Dew Point	40°F

Nitrogen is provided as automatic backup for Instrument Air.

12. Electric Power:	Large Motors	Small Motors	Instrument Lighting
Voltage Levels	4160 V	480 V	120 V
Phase	3 phase	3 phase	1
Frequency, Hz	60	60	60

C. PRODUCT SPECIFICATION

Methanol Purity

% methanol by wt.	99.83
% water and heavies	0.17
Specific gravity	0.798 at 20°C/20°C

SECTION 2

TECHNICAL INFORMATION

A. PROCESS DESCRIPTION

Area 01 - CPO Reactor With Heat Recovery (Ref. Dwg. FS-1001)

Saturated feedstock is preheated in the synthesis gas train before entering the fixed bed catalytic reactor along with preheated oxygen. The natural gas and oxygen are mixed in a specially designed chamber in the top of the reactor to ensure thorough mixing before the reactants contact the catalyst. The proprietary catalyst initiates the combustion of methane, hydrogen and carbon monoxide with the oxygen. The CPO vessel is internally lined with refractory and it may be mounted horizontally or vertically.

Process Heat Recovery

The synthesis gas exiting the CPO reactor is cooled to 100°F by the CPO boiler, feed heater, column reboiler, deaerator water heater, and finally, the synthesis gas cooler. Condensate is removed after the final exchanger and recycled to the saturator. The cooled gas passes to the synthesis gas compressor.

Area 02 - Synthesis Gas Compression (Ref. Dwg. FS - 2001)

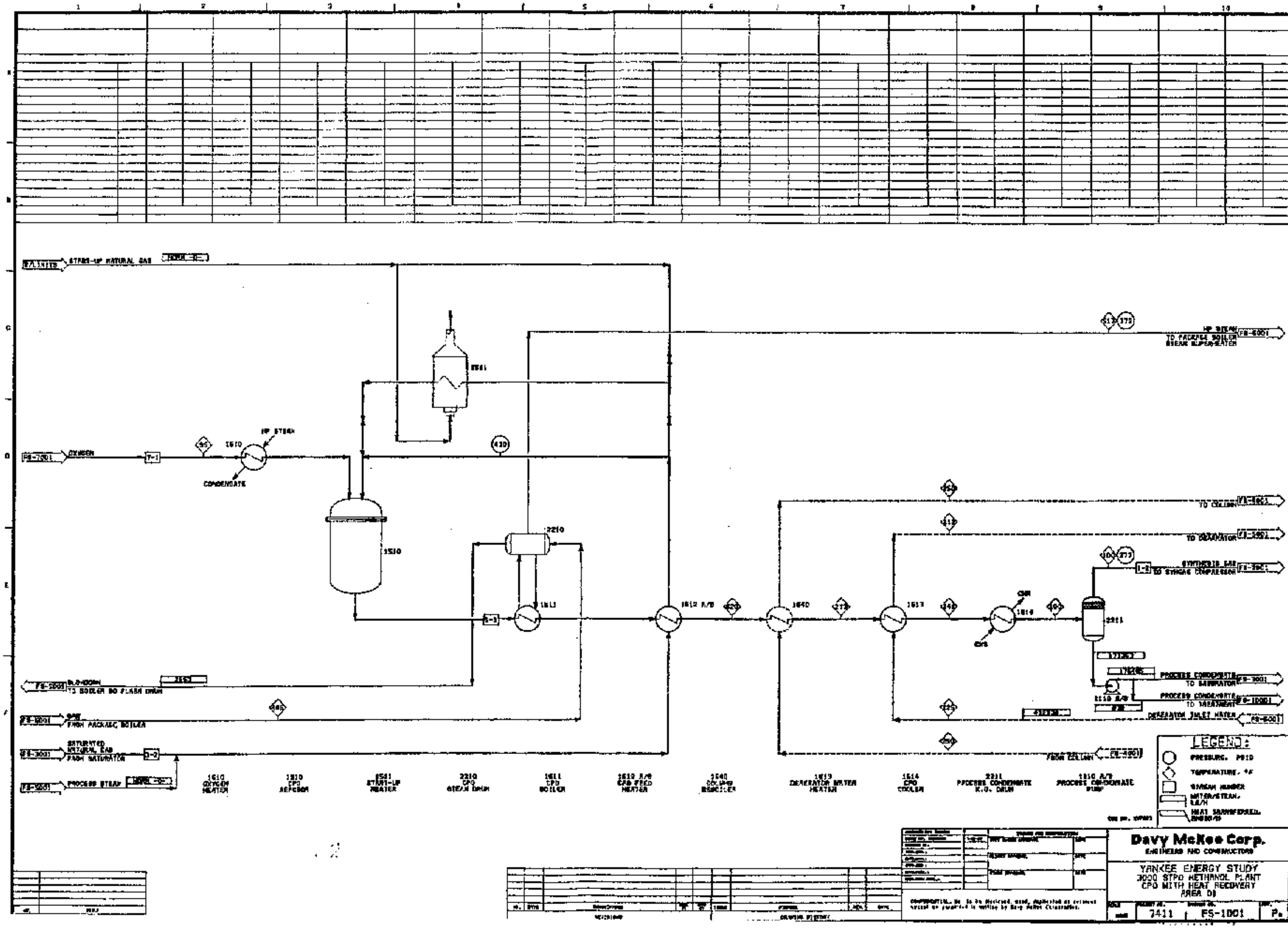
The make-up synthesis gas is compressed to the methanol synthesis loop pressure of approximately 80 atmospheres via a two stage centrifugal compressor. The compressor is equipped with an interstage cooler and separator drum for condensate removal, as well as a spillback cooler.

Compressed gas from the synthesis gas compressor is first passed through the guard bed prior to being mixed with the recycle synthesis loop gas at the discharge of the circulator. The combined gas then passes to the synthesis loop. The make-up synthesis gas compressor and the circulator share a common steam turbine drive, the turbine being a condensing type unit.

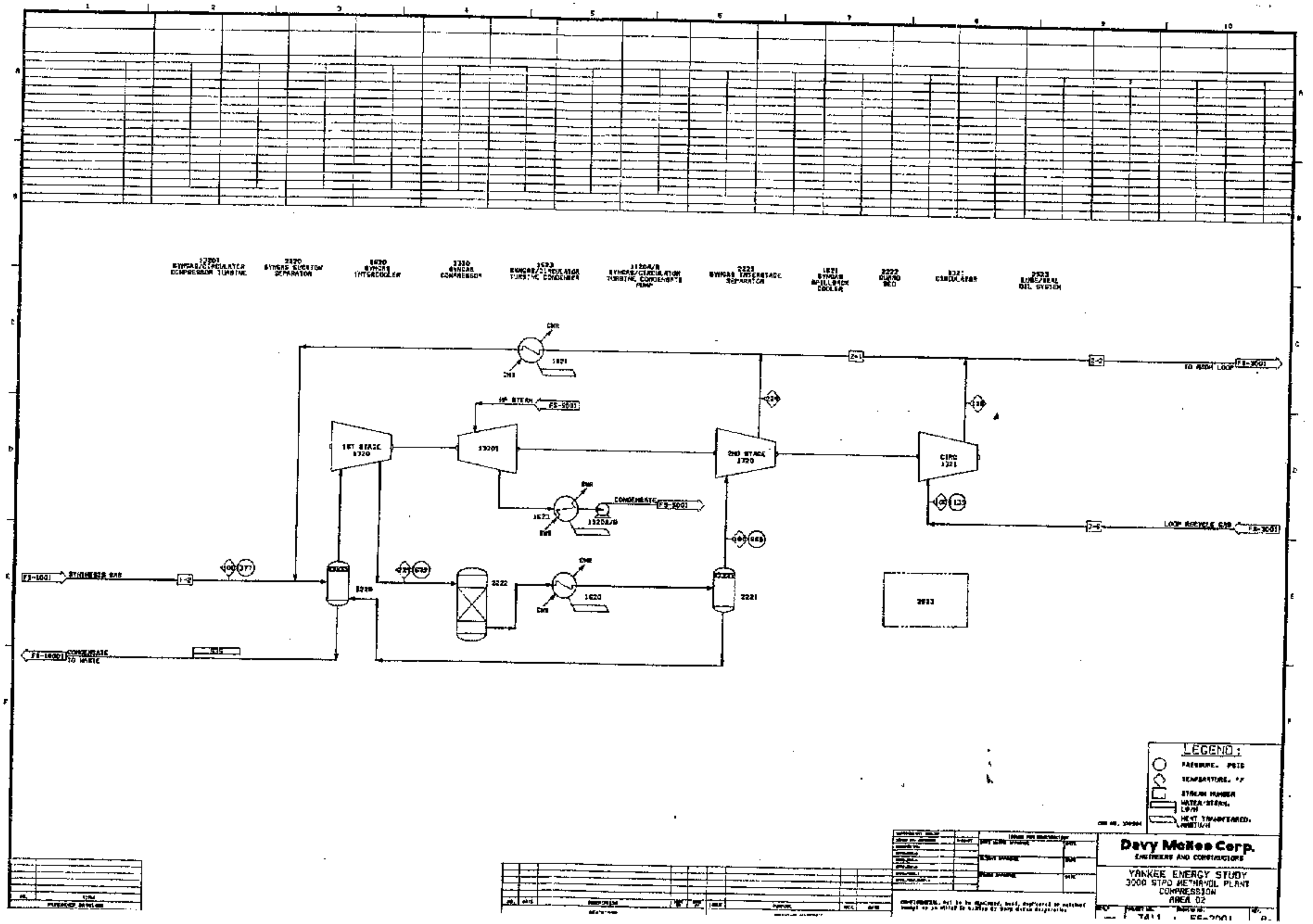
Area 03 - Methanol Synthesis Loop (Ref. Dwg. FS-3001)

The combined feed gas is heated in the loop interchanger and passed to the methanol converter. A tube cooled converter which requires no quench gas, has been utilized in order to maintain a simple design and operation for the methanol plant. In this converter, the catalyst fills the open spaces between vertical tubes which extend the length of the vessel. The feed gas enters the vessel through a manifold at the bottom and is distributed to empty tubes. As the gas passes through the tubes, it is heated by absorption of the heat of reaction occurring over the catalyst on the outside of the tubes. During the beginning-of-catalyst life conditions, a proportion of the converter feed gas will be bypassed around the tubes and fed directly to the top of the converter.

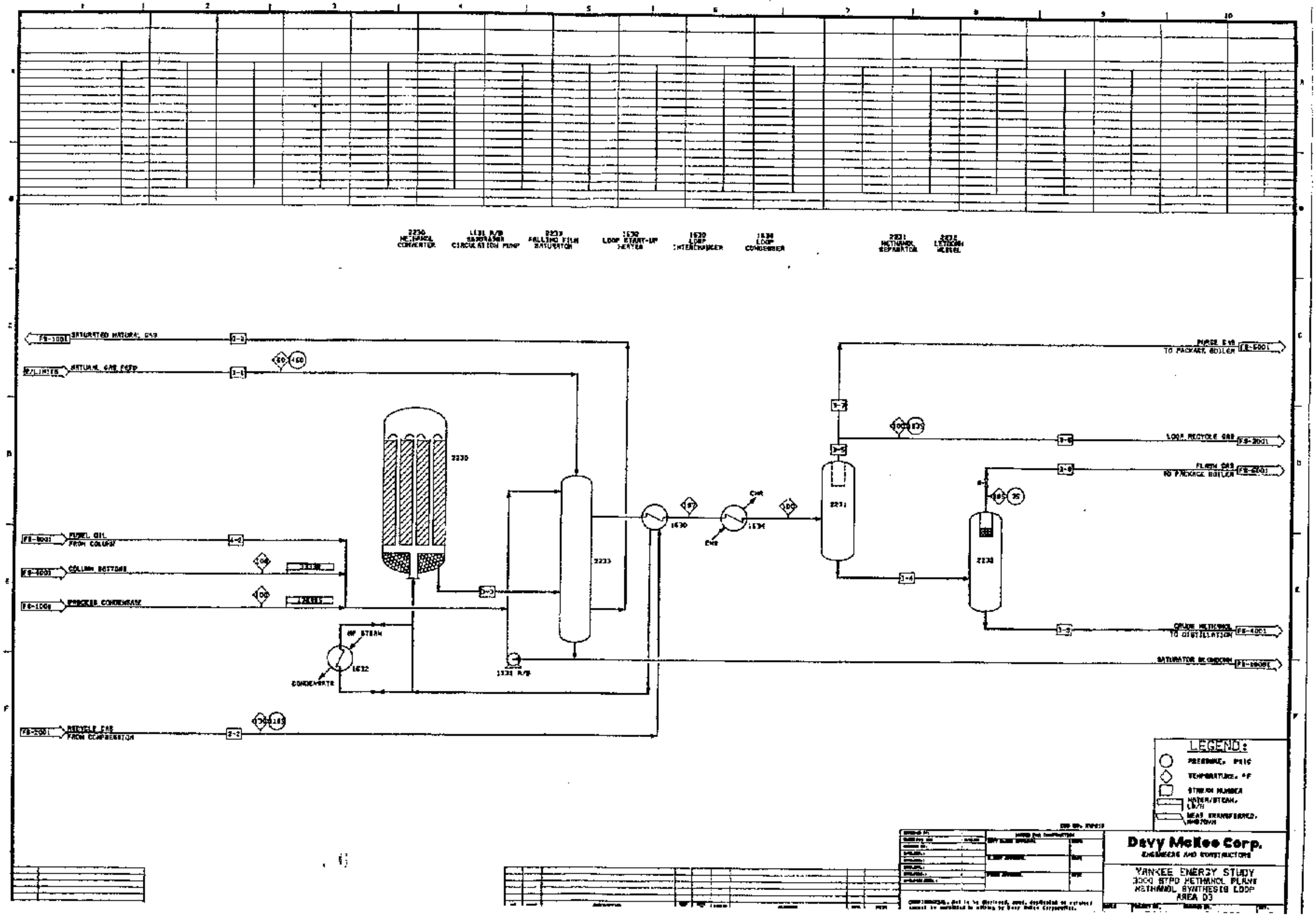
The converter effluent gas passes directed to the falling film saturator. The gas then passes to the methanol loop interchanger where it exchanges heat with the loop feed gas, the gas is then finally cooled in the loop condenser.



START-UP NATURAL GAS
OXYGEN
HP STEAM
CONDENSATE
1510
1541
2210
1511
1512 A/B
1540
1513
1514
2211
CW
HP STEAM
TO PACKAGE BOILER
STEAM SUPERHEATER
TO COLUMN
TO DEAERATOR
SYNTHESIS GAS
TO SYNGAS COMPRESSION
PROCESS CONDENSATE
TO SATURATOR
PROCESS CONDENSATE
TO TREATMENT
DEAERATOR INLET WATER
FROM COLUMN
BLOWDOWN
TO BOILER BD FLASH DRUM
BFW
FROM PACKAGE BOILER
SATURATED
NATURAL GAS
FROM SATURATOR
PROCESS STEAM
1610
OXYGEN
HEATER
1510
CPO
REFORMER
1541
START-UP
HEATER
2210
CPO
STEAM DRUM
1511
CPO
BOILER
1512 A/B
CPO FEED
HEATER
1540
COLUMN
REBOILER
1513
DEAERATOR WATER
HEATER
1514
CPO
COOLER
2211
PROCESS CONDENSATE
K.O. DRUM
1810 A/B
PROCESS CONDENSATE
PUMP
LEGEND:
PRESSURE, PSIG
TEMPERATURE, °F
STREAM NUMBER
HEAT TRANSFERRED,
Davy McKee Corp.
ENGINEERS AND CONSTRUCTORS
YANKEE ENERGY STUDY
3000 STPD METHANOL PLANT
CPO WITH HEAT RECOVERY
AREA DI
7411
FS-1DO1



1ST STAGE
1320
2ND STAGE
1320
CIRC
1321
13201
SYNTHESIS GAS
CONDENSATE
TO WASTE
HP STEAM
CONDENSATE
TO ASDM LOOP
LOOP RECYCLE GAS
2222
2221
2933
1620
1621
LEGEND:
PRESSURE, PSIG
TEMPERATURE, °F
STREAM NUMBER
Davy McKee Corp.
ENGINEERS AND CONSTRUCTORS
YANKEE ENERGY STUDY
3000 STPD METHANOL PLANT
COMPRESSION
AREA 02



2230 METHANOL CONVERTER
1630 LOOP INTERCHANGER
1634 LOOP CONDENSER
2231 METHANOL SEPARATOR
SATURATED NATURAL GAS
NATURAL GAS FEED
FUSEL OIL FROM COLUMN
COLUMN BOTTOMS
PROCESS CONDENSATE
RECYCLE GAS FROM COMPRESSION
PURGE GAS TO PACKAGE BOILER
LOOP RECYCLE GAS
FLASH GAS TO PACKAGE BOILER
CRUDE METHANOL TO DISTILLATION
SATURATOR BLOWDOWN
CONDENSATE
LEGEND:
PRESSURE, PSIG
TEMPERATURE, °F
STREAM NUMBER
Davy McKee Corp.
ENGINEERS AND CONSTRUCTORS
YANKEE ENERGY STUDY
3000 STPD METHANOL PLANT
METHANOL SYNTHESIS LOOP
AREA D3

The crude methanol product is recovered as a liquid from the loop gas in the methanol separator. The crude is let down in the letdown vessel and some entrained gas flashes off. The crude methanol then passes under its own pressure to either the distillation system or the crude methanol hold tank.

The inert components in the loop are controlled by adjusting the purge gas rate. The purge gas and letdown vessel flash gas are passed to the package boiler for use as fuel gas.

Area 04-Methanol Distillation (Ref. Dwg. FS-4001)

Since the methanol product from this plant is specified to be fuel grade, a single packed column distillation system is sufficient.

The crude methanol is sent to the single refining column and in order to prevent corrosion by small amounts of acidic impurities, a 1% caustic soda solution is injected into the crude methanol feed line. The rate is such as to maintain the refining column bottoms in a neutral or very slightly alkaline condition (ph 7.0 to 8.0).

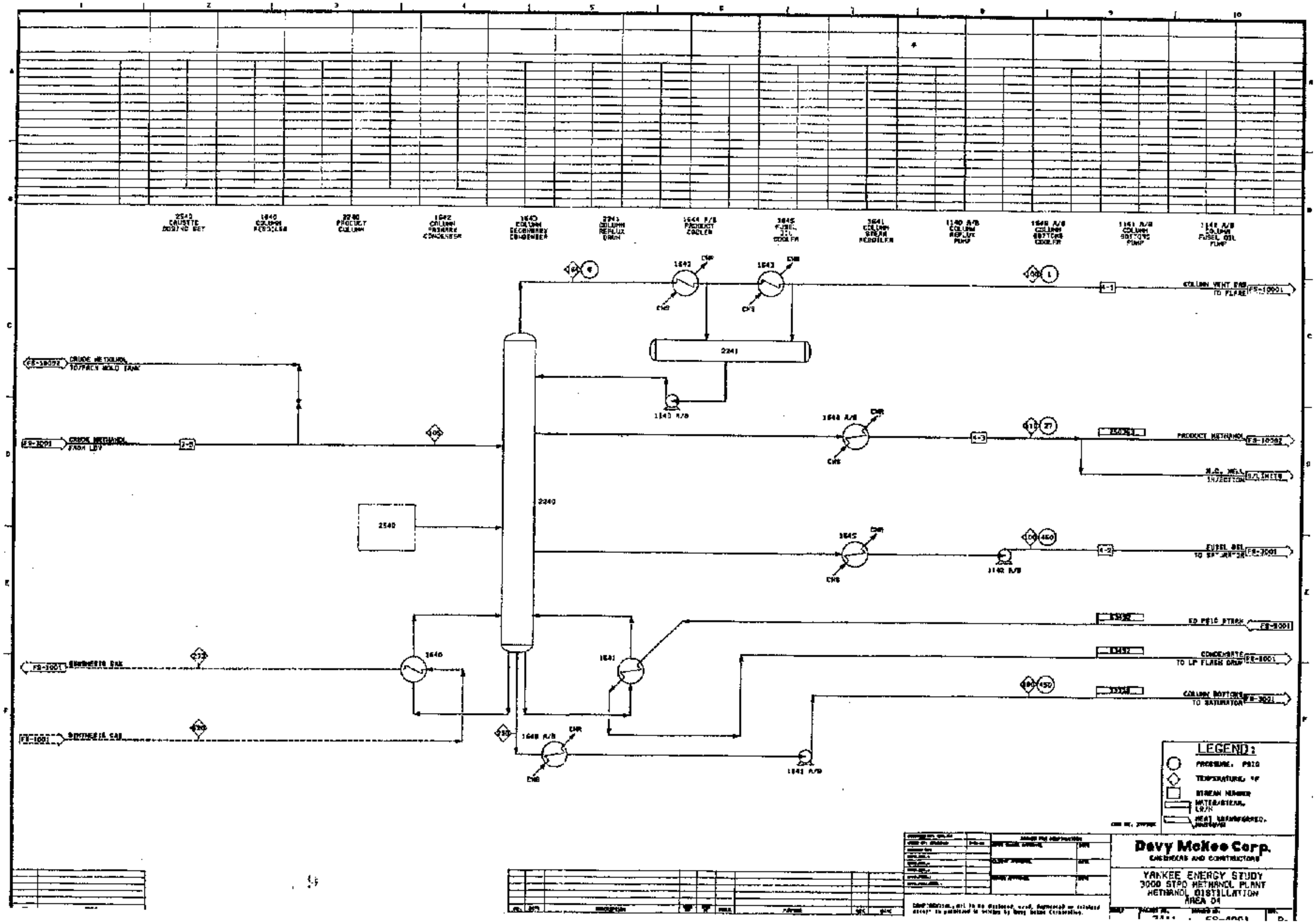
Overhead vapor from the column is passed through the column primary condenser where the bulk of the vapor is condensed and drains into the reflux drum. The remaining vapor passes through the secondary condenser where it is cooled and further condensation takes place. Condensate is returned to the column by the reflux pump. Uncondensed light ends, consisting of dissolved gases, dimethyl ether, ketones, methyl formate and methanol are vented to the flare.

Liquid product methanol is withdrawn from the refining column. It is cooled to approximately 110°F in the product cooler with the majority piped to the storage area described in the offsites section. Approximately .09% of the product methanol is removed for use as natural gas well injection.

A small quantity of heavy ends called fusel oil, consisting mostly of higher alcohols, concentrates in the stripping section of the column. A purge stream containing the heavies, methanol, and water is removed via a liquid side drawoff. Its water content prevents recombination with the product methanol.

This purge stream is cooled by the fusel oil cooler and pumped to the saturator where the methanol and heavy ends are vaporized into the saturated CPO reactor feed stream and reconverted to synthesis gas.

Below the fusel oil take-off, the column liquid downflow concentrates to almost 100% water. The bottoms stream withdrawn from the base of the column and cooled in the bottoms cooler is water containing approximately 300 ppm organics. This stream is pumped into the saturator system as part of the necessary water make-up.



2540 CAUSTIC DOSING SET
1640 COLUMN REBOILER
2240 PRODUCT COLUMN
1642 COLUMN PRIMARY CONDENSER
1643 COLUMN SECONDARY CONDENSER
2241 COLUMN REFLUX DRUM
1644 A/B PRODUCT COOLER
1645 FUSEL OIL COOLER
1641 COLUMN STEAM REBOILER
1140 A/B COLUMN REFLUX PUMP
1648 A/B COLUMN BOTTOMS COOLER
1141 A/B COLUMN BOTTOMS PUMP
1142 A/B COLUMN FUSEL OIL PUMP
COLUMN VENT GAS TO FLARE
PRODUCT METHANOL
FUSEL OIL TO SATURATOR
CONDENSATE TO LP FLASH DRUM
COLUMN BOTTOMS TO SATURATOR
SYNTHESIS GAS
2240
2241
2540
LEGEND:
PRESSURE, PSIG
TEMPERATURE, °F
STREAM NUMBER
Davy McKee Corp.
ENGINEERS AND CONSTRUCTORS
YANKEE ENERGY STUDY
3000 STPD METHANOL PLANT
METHANOL DISTILLATION
AREA 04

Area 05 - Steam System (Ref. Dwg. FS-5001)

The steam system is designed for two pressure levels: 750 psig, 750°F and 60 psig saturated. The high pressure level is used for steam turbine drives, steam ejectors, and start-up process heat. The low pressure level steam is used for process heat, stripping requirements and make-up of the offsite diesel generator.

Steam is generated in the synthesis gas train and the package boiler at 775 psig, 517°F and superheated to 775°F in the package boiler. The machines being driven by condensing turbines are the syngas, circulator and the air compressors. These turbines condense at 4" Hg absolute. Extraction turbines letting down to the 50 psig level, are used to drive the BFW pump and several miscellaneous small drivers.

The deaerator serves to degas the demineralized water feed which is a combination of condensate from the turbine condensers, distillation reboiler, desalination plant heaters, natural gas heaters and fresh desalinated make-up water. The turbine condensate and fresh make-up water are combined and preheated in the deaerator water heater before being added to the deaerator. Condensate from the reboiler and heaters first passes through a flash drum with the flash steam entering the deaerator for stripping purposes and the condensate being sent to the top of the packed section. Boiler feedwater from the deaerator is then preheated in the package boiler.

Area 06 - Package Boiler (Ref. Dwg. FS 6001)

In the package boiler HP steam is raised and superheated (including steam from the CPO boiler) and all boiler feedwater is preheated before being split to the two steam drums. The HP superheated steam is distributed to the various users.

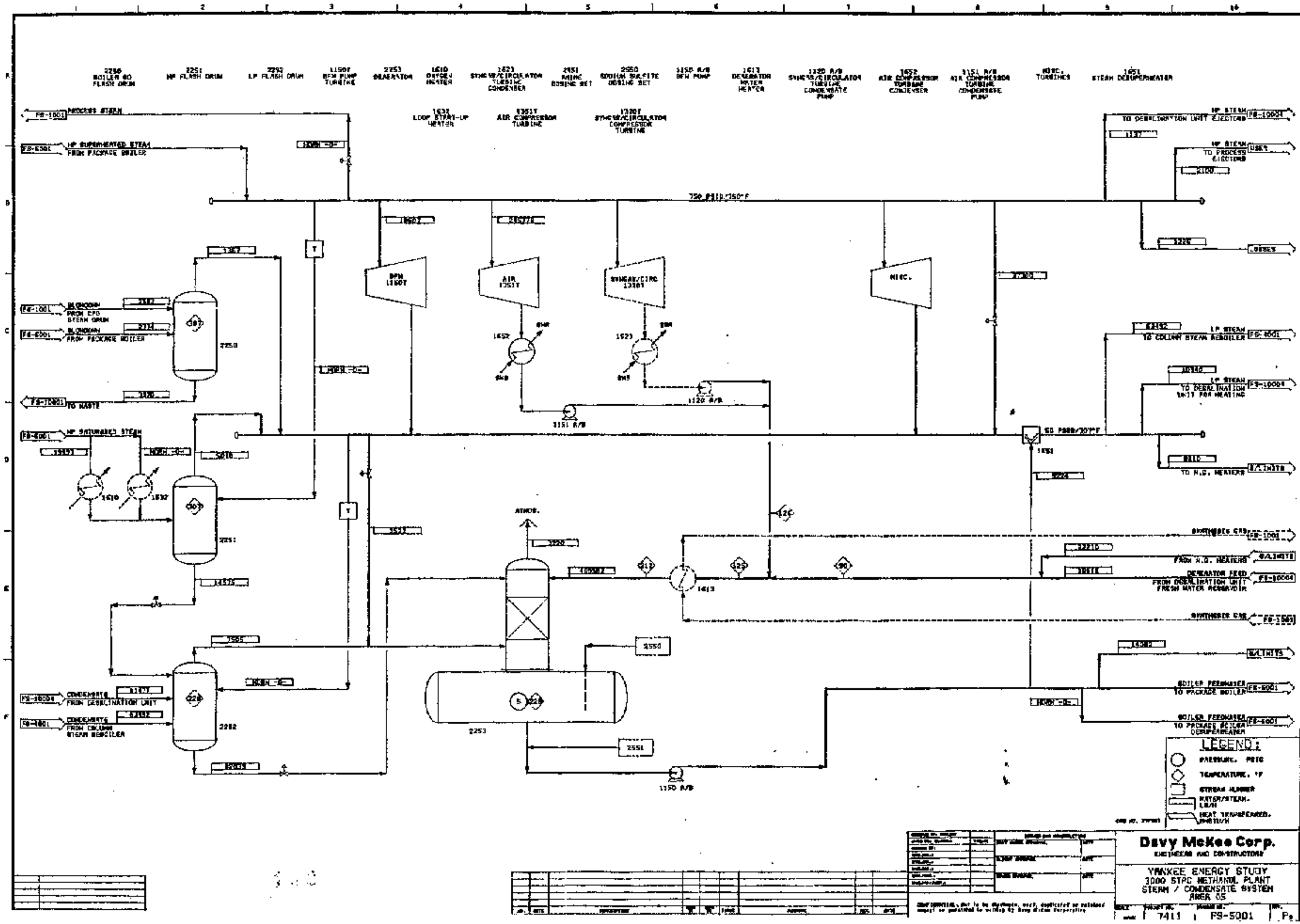
Area 07 - Gaseous Oxygen Plant (Ref. Dwg. FS-7001)

The oxygen plant is a package unit which produces gaseous 02 for the process and gaseous N2 for start-up requirements and storage tank blanketing. Atmospheric air is compressed, passed over molecular sieve adsorbers, cooled in a plate-and-frame multi-pass interchanger, and expanded into a cyrogenic column. In the two section column (HP bottom section. LP top section) air is separated into liquid oxygen and gaseous nitrogen. The liquid oxygen (of 99.5% purity) is pumped through the heat interchanger, vaporized and delivered to the CPO Reactor at 430 psig. The gaseous nitrogen (99.9% purity) is partially compressed for use on the plant with the excess being vented to atmosphere.

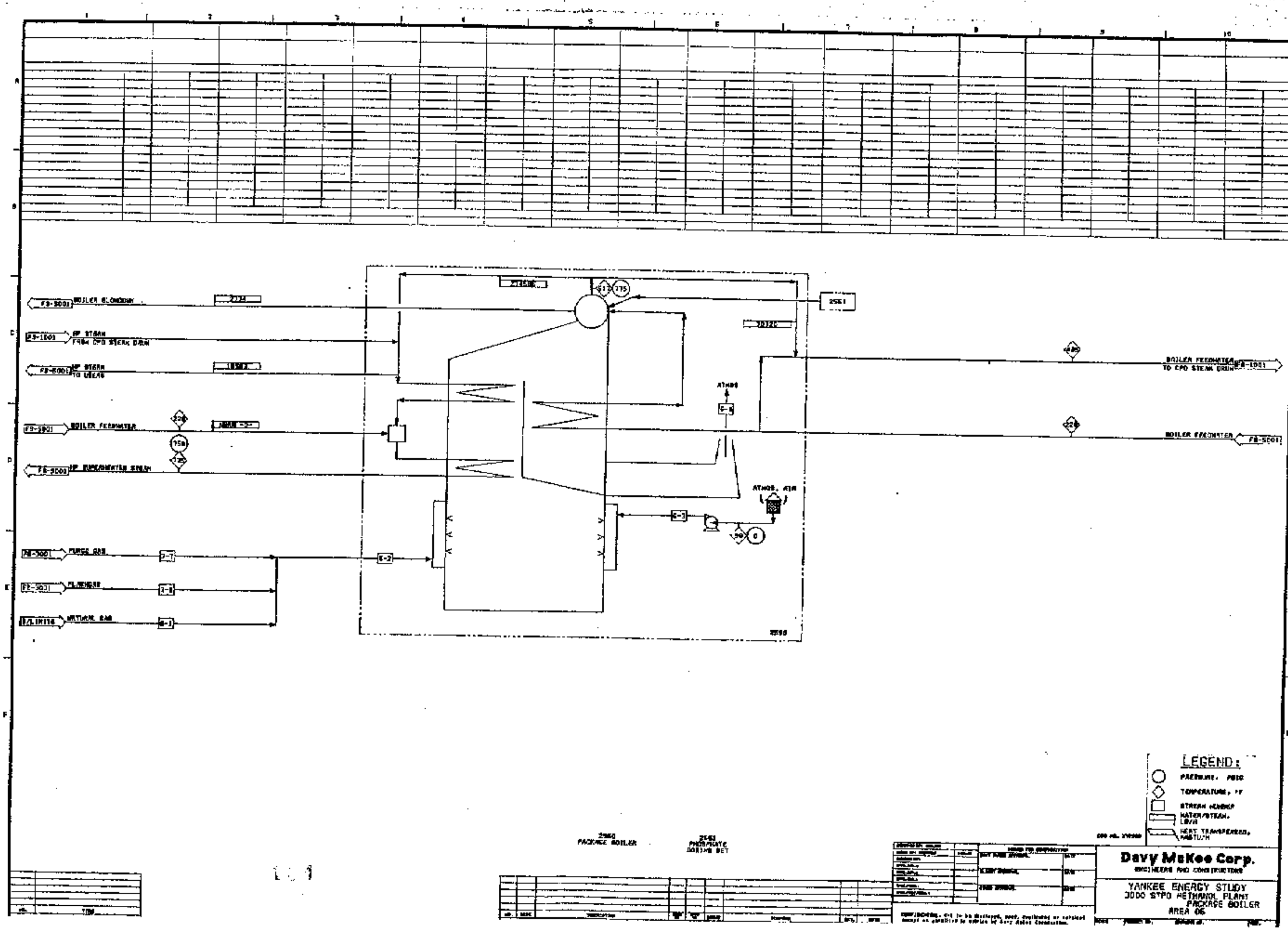
Area 10 - Offsites (Ref. Dwgs. FS-10001 thru FS-10004)

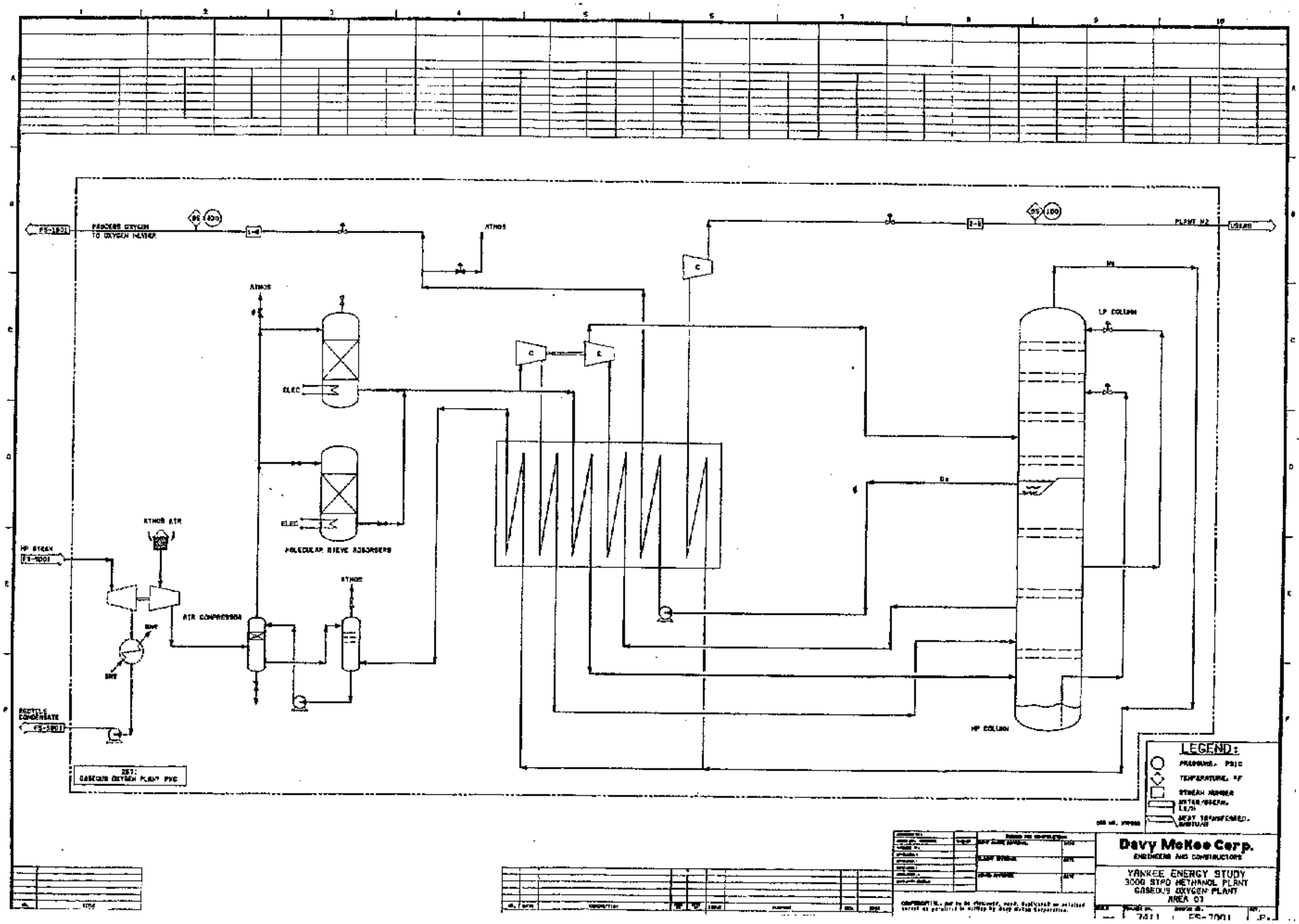
Flare/Waste System

A flare system is provided to dispose of vent gases from normal plant operations and larger vent gas volumes generated during equipment start-up

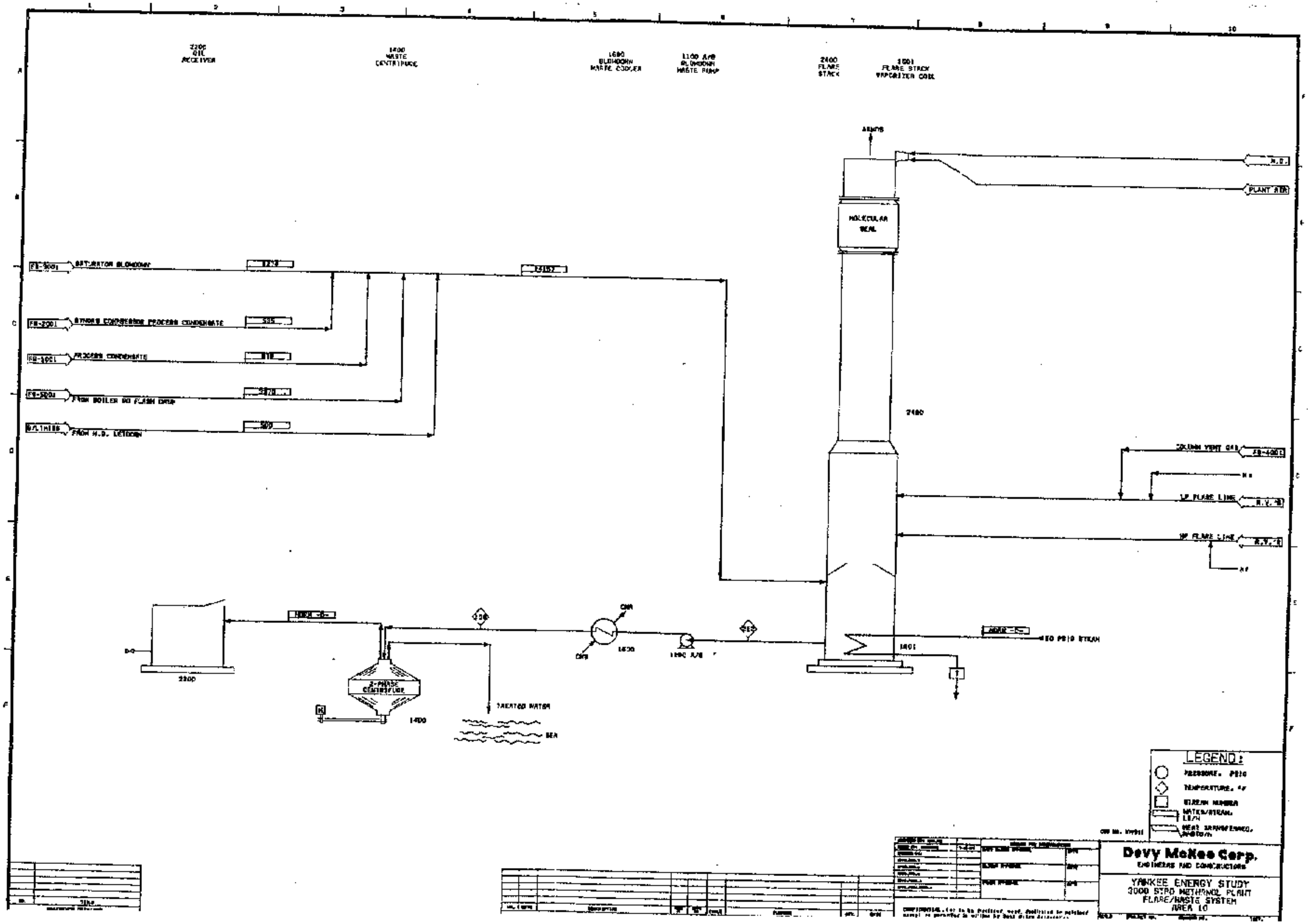


2250 BOILER BD FLASH DRUM
2251 HP FLASH DRUM
2252 LP FLASH DRUM
1150T BFW PUMP TURBINE
2253 DEAERATOR
1610 OXYGEN HEATER
1632 LOOP START-UP HEATER
1623 SYNGAS/CIRCULATOR TURBINE CONDENSER
1351T AIR COMPRESSOR TURBINE
2351 AMINE DOSING SET
2550 SODIUM SULFITE DOSING SET
1330T SYNGAS/CIRCULATOR COMPRESSOR TURBINE
1150 A/B BFW PUMP
1613 DEAERATOR WATER HEATER
1120 A/B SYNGAS/CIRCULATOR TURBINE CONDENSATE PUMP
1652 AIR COMPRESSOR TURBINE CONDENSER
1151 A/B AIR COMPRESSOR TURBINE CONDENSATE PUMP
MISC. TURBINES
1651 STEAM DESUPERHEATER
PROCESS STEAM
HP SUPERHEATED STEAM FROM PACKAGE BOILER
BLOWDOWN FROM CPG STEAM DRUM
BLOWDOWN FROM PACKAGE BOILER
TO WASTE
HP SATURATED STEAM
CONDENSATE FROM DESALINATION UNIT
CONDENSATE FROM COLUMN STEAM REBOILER
750 PSIG/750°F
50 PSIG/307°F
ATMOS.
HP STEAM TO DESALINATION UNIT EJECTORS
HP STEAM TO PROCESS EJECTORS
LOSSES
LP STEAM TO COLUMN STEAM REBOILER
LP STEAM TO DESALINATION UNIT FOR HEATING
TO N.G. HEATERS
SYNTHESIS GAS
FROM N.G. HEATERS
DEAERATOR FEED FROM DESALINATION UNIT FRESH WATER RESERVOIR
BOILER FEEDWATER TO PACKAGE BOILER
BOILER FEEDWATER TO PACKAGE BOILER DESUPERHEATER
LEGEND:
PRESSURE, PSIG
TEMPERATURE, °F
STREAM NUMBER
WATER/STEAM, LB/H
HEAT TRANSFERRED, MMBTU/H
Davy McKee Corp.
ENGINEERS AND CONSTRUCTORS
YANKEE ENERGY STUDY
3000 STPD METHANOL PLANT
STEAM / CONDENSATE SYSTEM
AREA 05
7411
FS-5001

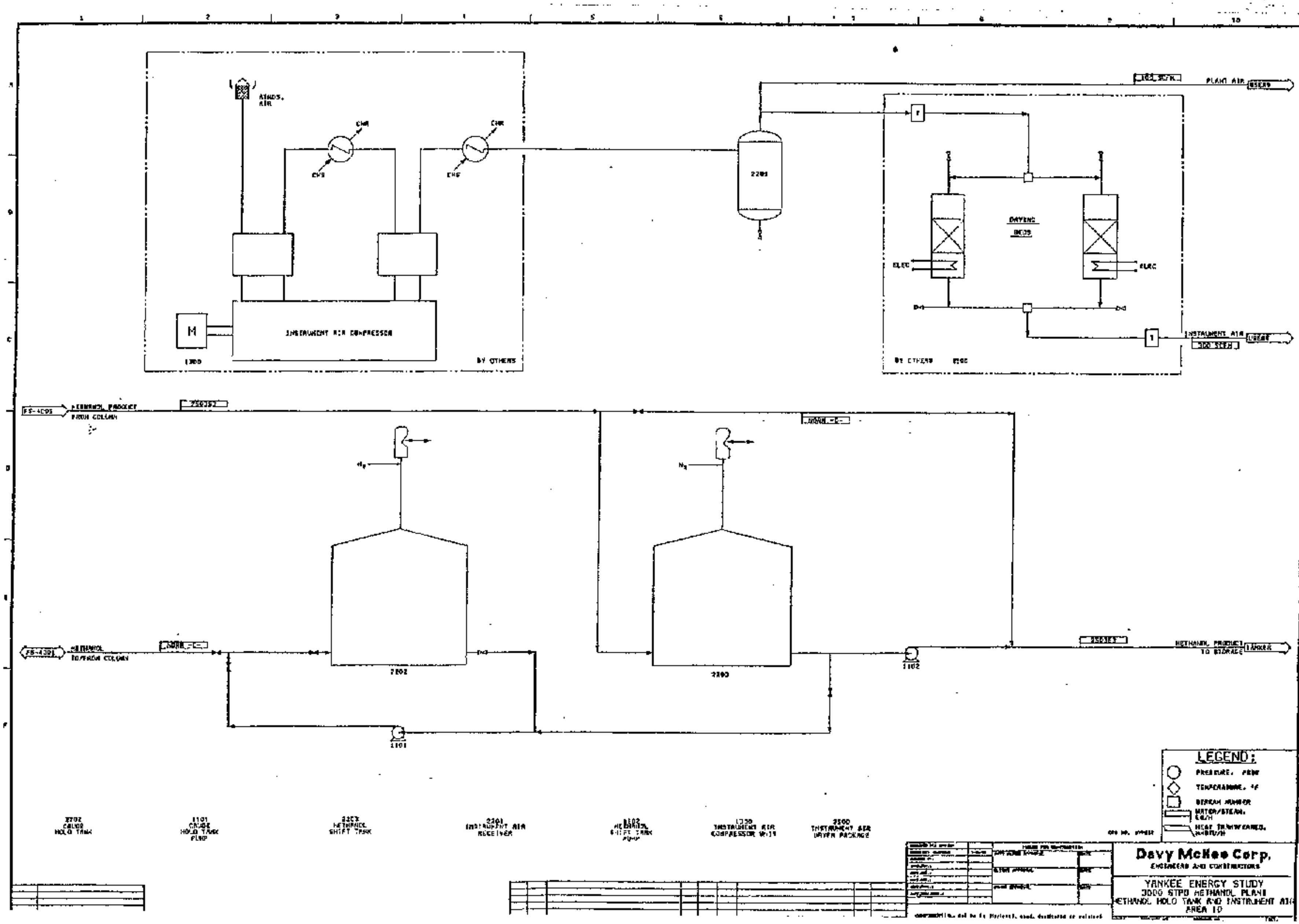

BOILER BLOWDOWN
HP STEAM FROM CPO STEAM DRUM
HP STEAM TO USERS
BOILER FEEDWATER
HP SUPERHEATED STEAM
PURGE GAS
FLASHGAS
NATURAL GAS
BOILER FEEDWATER TO CPO STEAM DRUM
BOILER FEEDWATER
ATMOS.
ATMOS. AIR
2560 PACKAGE BOILER
2561 PHOSPHATE DOSING SET
LEGEND:
PRESSURE, PSIG
TEMPERATURE, °F
STREAM NUMBER
WATER/STEAM, LB/H
HEAT TRANSFERRED, MMBTU/H
Davy McKee Corp.
ENGINEERS AND CONSTRUCTORS
YANKEE ENERGY STUDY
3000 STPD METHANOL PLANT
PACKAGE BOILER
AREA 06

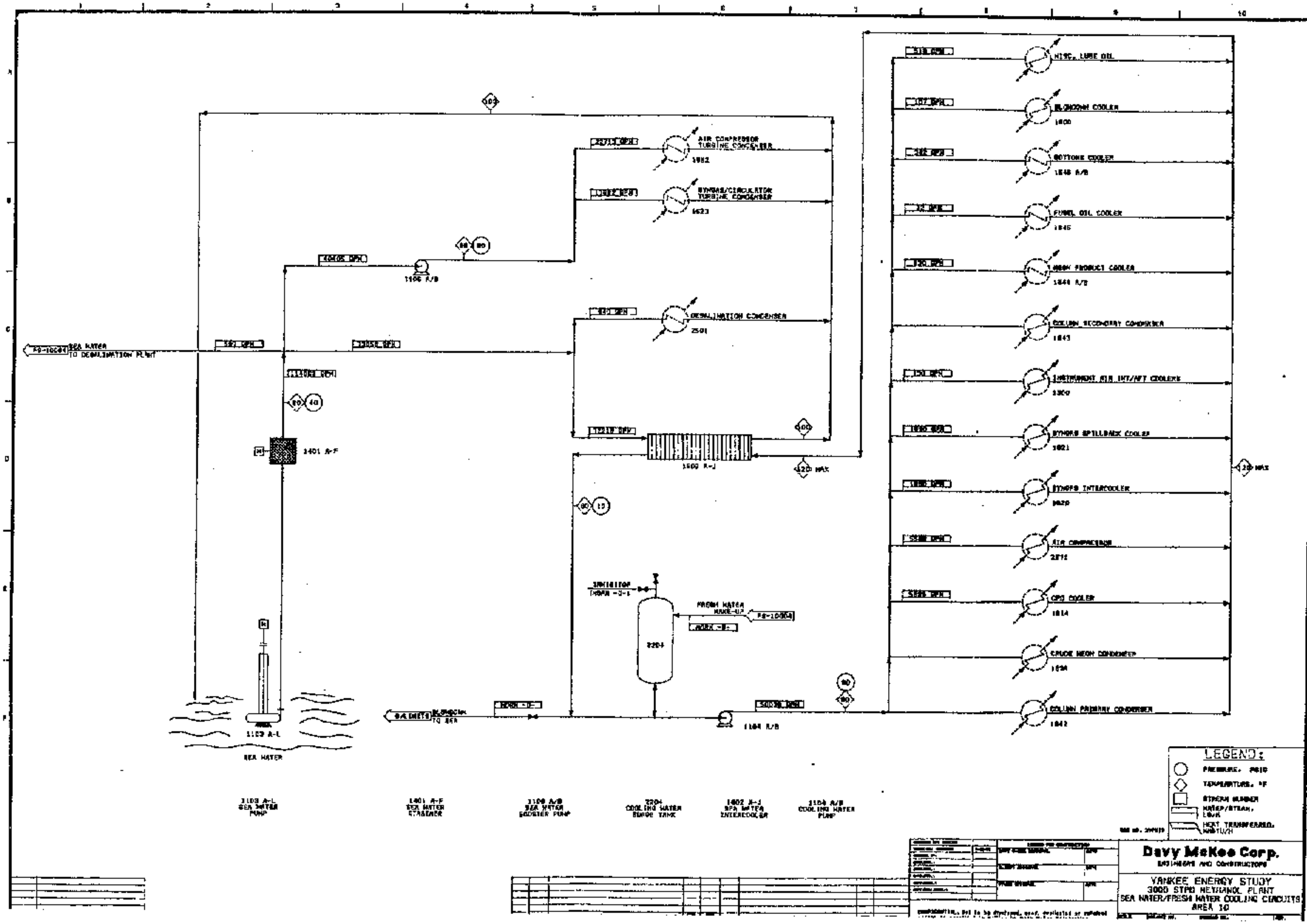


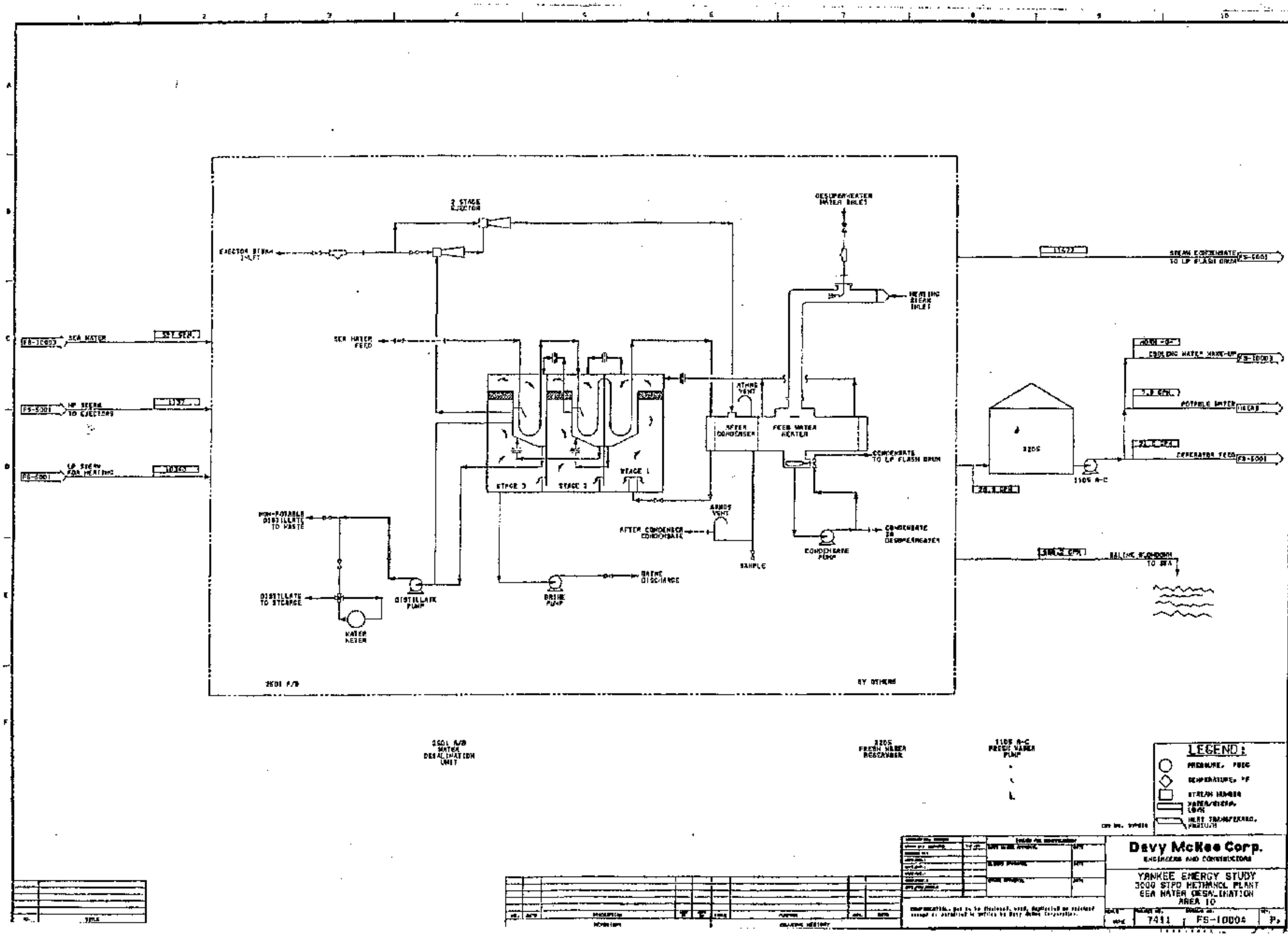
PROCESS OXYGEN TO OXYGEN HEATER
FS-1801
95
300
1-4
ATMOS
PLANT N2
USERS
95
100
2-2
C
N2
LP COLUMN
HP COLUMN
MOLECULAR SIEVE ADSORBERS
ELEC
ATMOS AIR
HP STEAM
FS-3001
AIR COMPRESSOR
RECYCLE CONDENSATE
FS-3801
257: GASEOUS OXYGEN PLANT PFC
LEGEND:
PRESSURE, PSIG
TEMPERATURE, °F
STREAM NUMBER
HEAT TRANSFERRED, MMBTU/H
Davy McKee Corp.
ENGINEERS AND CONSTRUCTORS
YANKEE ENERGY STUDY
3000 STPD METHANOL PLANT
GASEOUS OXYGEN PLANT
AREA 03
7411
FS-7001



2200 OIL RECEIVER
1400 WASTE CENTRIFUGE
1600 BLOWDOWN WASTE COOLER
1100 A/B BLOWDOWN WASTE PUMP
2400 FLARE STACK
1001 FLARE STACK VAPORIZER COIL
ATMOS
N.D.
PLANT AIR
MOLECULAR SEAL
SATURATOR BLOWDOWN
SYNGAS COMPRESSOR PROCESS CONDENSATE
PROCESS CONDENSATE
FROM BOILER BD FLASH DRUM
FROM H.D. LETDOWN
2400
COLUMN VENT GAS
LP FLARE LINE
HP FLARE LINE
TREATED WATER
SEA
2-PHASE CENTRIFUGE
1400
2200
1600
1401
CWR
CWS
LEGEND:
PRESSURE, PSIG
TEMPERATURE, °F
STREAM NUMBER
Davy McKee Corp.
ENGINEERS AND CONSTRUCTORS
YANKEE ENERGY STUDY
3000 STPD METHANOL PLANT
FLARE/WASTE SYSTEM
AREA 10


ATMOS. AIR
CWR
CWS
INSTRUMENT AIR COMPRESSOR
M
1300
BY OTHERS
2201
DRYING BEDS
ELEC
BY OTHERS
PLANT AIR
USERS
INSTRUMENT AIR
USERS
METHANOL PRODUCT FROM COLUMN
METHANOL TO/FROM COLUMN
N_2
2202
2203
1101
1102
METHANOL PRODUCT TO STORAGE
TANKER
2702 CRUDE HOLD TANK
1101 CRUDE HOLD TANK PUMP
2202 METHANOL SHIFT TANK
2201 INSTRUMENT AIR RECEIVER
1102 METHANOL SHIFT TANK PUMP
1300 INSTRUMENT AIR COMPRESSOR W/13
2200 INSTRUMENT AIR DRYER PACKAGE
LEGEND:
PRESSURE, PSIG
TEMPERATURE, °F
STREAM NUMBER
WATER/STEAM, LB/H
HEAT TRANSFERRED, MMBTU/H
Davy McKee Corp.
ENGINEERS AND CONSTRUCTORS
YANKEE ENERGY STUDY
3000 STPD METHANOL PLANT
METHANOL HOLD TANK AND INSTRUMENT AIR
AREA 10


LEGEND:
PRESSURE, PSIG
TEMPERATURE, °F
STREAM NUMBER
HEAT TRANSFERRED, MMBTU/H
Davy McKee Corp.
ENGINEERS AND CONSTRUCTORS
YANKEE ENERGY STUDY
3000 STPD METHANOL PLANT
SEA WATER/FRESH WATER COOLING CIRCUITS
AREA 10
MISC. LUBE OIL
BLOWDOWN COOLER
BOTTOMS COOLER
FUEL OIL COOLER
MEOH PRODUCT COOLER
COLUMN SECONDARY CONDENSER
INSTRUMENT AIR INT/AFT COOLERS
SYNGAS SPILLBACK COOLER
SYNGAS INTERCOOLER
AIR COMPRESSOR
CPO COOLER
CRUDE MEOH CONDENSER
COLUMN PRIMARY CONDENSER
AIR COMPRESSOR TURBINE CONDENSER
SYNGAS/CIRCULATOR TURBINE CONDENSER
DESALINATION CONDENSER
SEA WATER TO DESALINATION PLANT
SEA WATER
FRESH WATER MAKE-UP
1100 A-L SEA WATER PUMP
1401 A-F SEA WATER STRAINER
1106 A/B SEA WATER BOOSTER PUMP
2204 COOLING WATER SURGE TANK
1602 A-J SEA WATER INTERCOOLER
1104 A/B COOLING WATER PUMP



2 STAGE EJECTOR
EJECTOR STEAM INLET
SEA WATER FEED
DESUPERHEATER WATER INLET
HEATING STEAM INLET
SEA WATER
HP STEAM TO EJECTORS
LP STEAM FOR HEATING
STAGE 3
STAGE 2
STAGE 1
ATMOS VENT
AFTER CONDENSER
FEED WATER HEATER
CONDENSATE TO LP FLASH DRUM
CONDENSATE TO DESUPERHEATER
CONDENSATE PUMP
AFTER CONDENSER CONDENSATE
SAMPLE
NON-POTABLE DISTILLATE TO WASTE
DISTILLATE TO STORAGE
WATER METER
DISTILLATE PUMP
BRINE PUMP
BRINE DISCHARGE
2601 A/B
BY OTHERS
STEAM CONDENSATE TO LP FLASH DRUM
COOLING WATER MAKE-UP
POTABLE WATER
GENERATOR FEED
SALINE BLOWDOWN TO SEA
3205
1105 A-C
2601 A/B WATER DESALINATION UNIT
3205 FRESH WATER RECEIVER
1105 A-C FRESH WATER PUMP
LEGEND:
PRESSURE, PSIG
TEMPERATURE, °F
STREAM NUMBER
Davy McKee Corp.
ENGINEERS AND CONSTRUCTORS
YANKEE ENERGY STUDY
3000 STPD METHANOL PLANT
SEA WATER DESALINATION
AREA 10
7411
FS-10004