6.2.12 AMMONIA RECOVERY - UNIT 18

6.2.12.1 DESIGN BASIS

Purpose of Unit

Purpose of this unit is to recover the ammonia in the gas liquor from the Phenosolvan unit and to produce an anhydrous ammonia product. The stripped gas liquor is sent to the wastewater treatment unit. The acid gas containing mainly H_2S and CO_2 generated in the unit is sent to the Stretford unit for sulfur recovery.

Scope of Unit

The unit includes IFE solvent stripping followed by total acid gas stripping of the incoming gas liquor, selective solvent (phosphoric acid) absorption of ammonia from the acid gases generated in the stripper, solvent regeneration and purification of the ammonia to anhydrous ammonia product for sales. The NH_3 recovery process, called Phosam-W, is licensed by U.S. Steel.

General Design Criteria

The unit consists of two 50 percent parallel trains. The pumps, all motor driven, are sufficiently spared to provide reliable operation. The unit onstream factor is compatible with the overall plant stream factor of 332 days per year.

Process Performance Objectives

The process is expected to recover approximately 93 to 94 percent of the NH_3 in the feed liquor. Product is in the form of 99.5 percent pure

6.2.12.1 (Continued)

anhydrous ammonia liquid. The stripped liquor contains no more than 70 ppm weight of free ammonia and less than 25 ppm weight each of H_2S and CO_2.

Feedstock

Gas Liquor from Phenosolvan

Feed Rate	973,604 lb/hr	
NH_3 in Feed Liquor (Free & Fixed)	6,783 lb/hr	
Battery Limit Conditions:	Temperature	120°F
	Pressure	100 psia

Products

Anhydrous Liquid Ammonia Product

Production Rate	6,398 lb/hr or 76.8 ST/SD	
NH_3	99.5%	
H_2O	0.5%	
Battery Limit Conditions:	Temperature	100°F
	Pressure	220 psia

Stripped Gas Liquor Effluent

Liquor Flow Rate	943,951 lb/hr or 1,886 gpm	
Free NH_3	70 ppmw	
CO_2, H_2S	25 ppmw	
Battery Limit Conditions:	Temperature	100°F
	Pressure	145 psia

6.2.12.1 (Continued)

Acid Gases

Flow Rate		560.6 lb-mol/hr	
		or 23,255 lb/hr	
Composition (dry):	CO_2	99.15%	
	H_2S	0.85	
	NH_3	≤500 ppmw	
Battery Limit Conditions:	Temperature		120°F
	Pressure		20 psia

Utility Requirements

Steam consumed (600 psig sat'd)	80,000 lb/hr
Steam consumed (100 psig sat'd)	142,200 lb/hr
Electric Power	480 kW
Cooling Water (ΔT = 30°F)	8,165 gpm
Phosphoric acid (100 %) solvent makeup required	102 tons/year
Sodium Hydroxide (50 % solution) required	552 tons/year

6.2.12.2 PROCESS DESCRIPTION

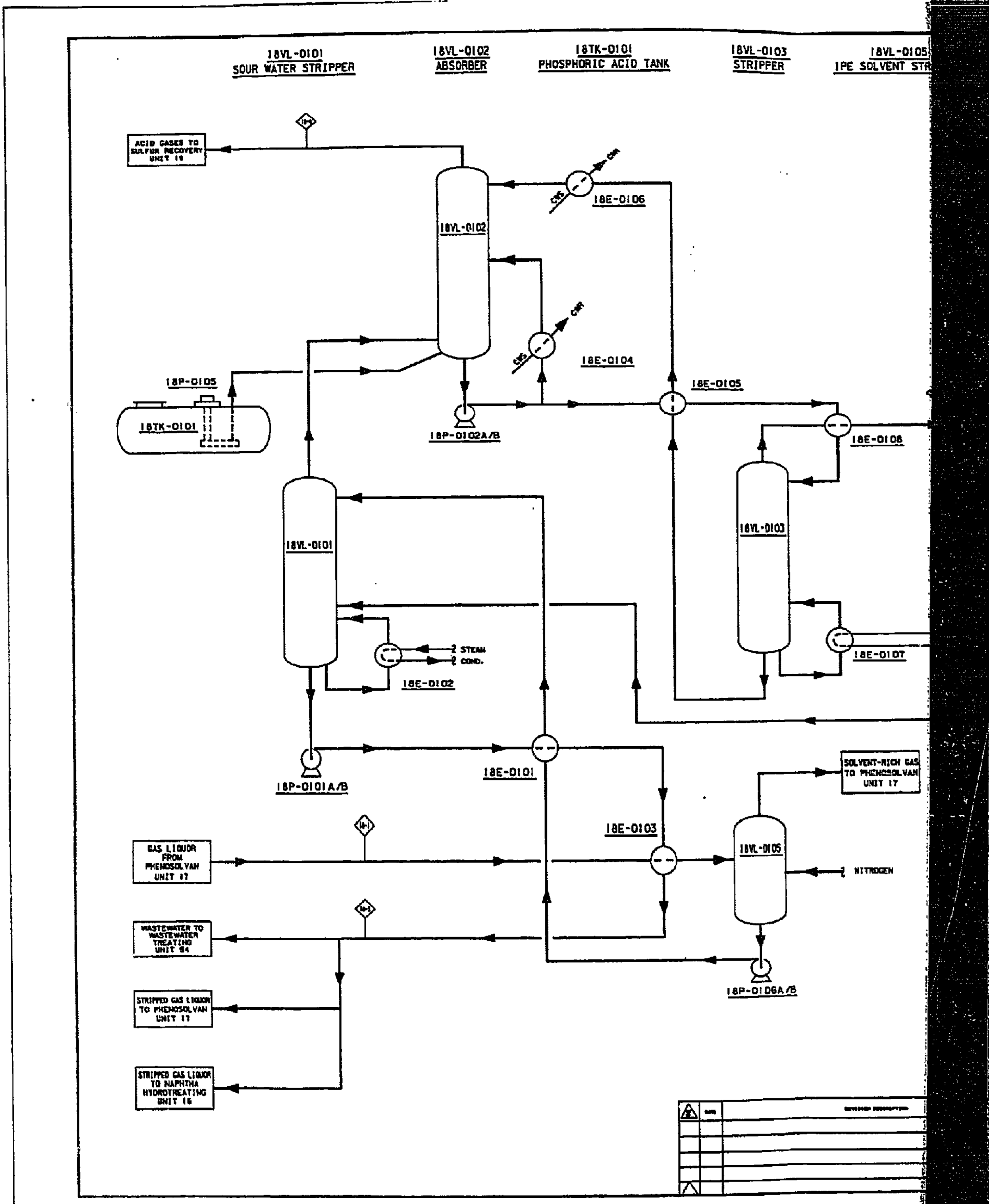
The process flow sketch of the unit is shown on Drawing No. 835704-18-4-101. Material balance and equipment list (Tables 6.2.12-1 and 6.2.12-2) for the unit follow.

The gas liquor feed to the unit from Phenosolvan extraction section contains recoverable IPE solvent. After preheating with the sour water stripper bottoms, the solvent is stripped from feed liquor with nitrogen-rich gas in the solvent stripper. The solvent-rich gas is recycled to the Phenosolvan unit. The gas liquor from the bottom of solvent stripper is preheated using stripped gas liquor effluent from the unit and sent to the

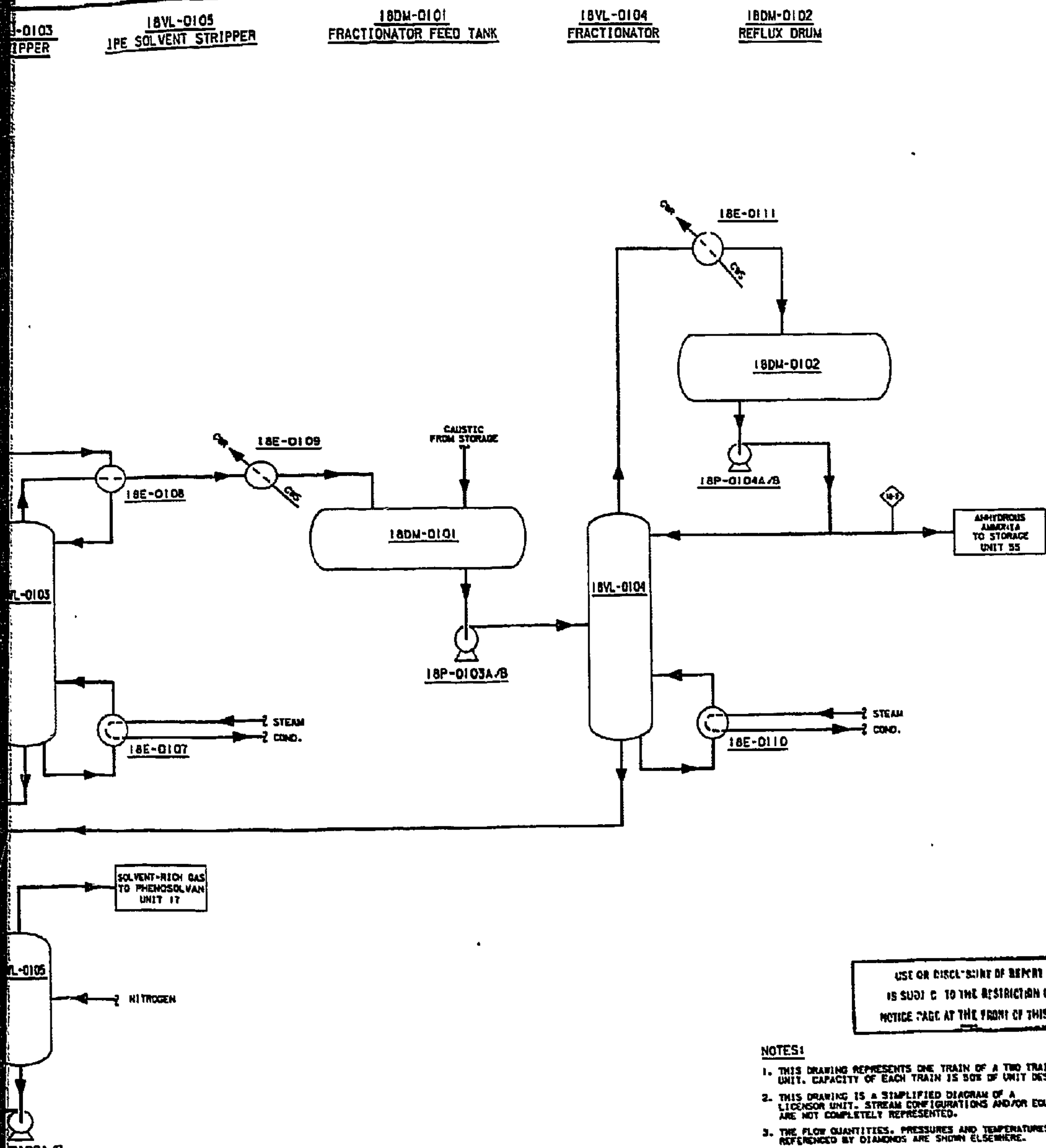
6.2.12.2 (Continued)

sour water stripper. In the steam-reboiled sour water stripper, volatile components, CO_2, H_2S, HCN and NH_3, are stripped from the gas liquor. The stripped gas liquor is cooled preheating the feed streams and pumped to the Waste Water Treating unit. Slipstreams of the stripped gas liquor are recycled to the Phenosolvan and Naphtha Hydrotreating units.

The offgas from the sour water stripper flow to the absorber, where ammonia is selectively absorbed into an ammonium phosphate solution. The unabsorbed acid gases (CO_2, H_2S and HCN) are directed to the Stretford section of the Sulfur Recovery unit. Ammonia-rich solution leaving the absorber bottom is regenerated in a steam-reboiled stripper. The phosphate solvent is recycled to the absorber with some fresh phosphoric acid makeup. The ammonia released from the stripper is condensed and fed to the fractionator. The steam-reboiled fractionator column purifies the ammonia to produce liquid anhydrous ammonia which is sent to storage. The water removed at the bottom of the fractionator is recycled to the sour water stripper.



18VL-0101
SOUR WATER STRIPPER
18VL-0102
ABSORBER
18TK-0101
PHOSPHORIC ACID TANK
18VL-0103
STRIPPER
18VL-0105
IPE SOLVENT STR
ACID GASES TO SULFUR RECOVERY UNIT 19
18E-0106
18VL-0102
18E-0104
18P-0105
18E-0105
18TK-0101
18P-0102A/B
18E-0108
18VL-0101
18VL-0103
STEAM
COND.
18E-0102
18E-0107
18P-0101A/B
18E-0101
SOLVENT-RICH GAS TO PHENOSOLVAN UNIT 17
18E-0103
GAS LIQUOR FROM PHENOSOLVAN UNIT 17
18VL-0105
NITROGEN
WASTEWATER TO WASTEWATER TREATING UNIT 84
18P-0106A/B
STRIPPED GAS LIQUOR TO PHENOSOLVAN UNIT 17
STRIPPED GAS LIQUOR TO NAPHTHA HYDROTREATING UNIT 16



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NOTES:

1. THIS DRAWING REPRESENTS ONE TRAIN OF A TWO TRAIN UNIT. CAPACITY OF EACH TRAIN IS 50% OF UNIT DESIGN.
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FLUOR

D.P.HALVERSON
C.C.ABATAY
R.O'BELMITO
R.McCARTHY
R.LANG

PROCESS FLOW DIAGRAM
AMMONIA RECOVERY
UNIT 18

CROW TRIBE OF INDIANS — SYNFUELS FEASIBILITY STUDY

Scale	Drawing Number	Revision
NONE	835704-18-4-101	1

003 3571810I

TABLE 6.2.12-1

MATERIAL BALANCE

AMMONIA RECOVERY - UNIT 18

Stream Number Stream Name	18-1 Gas Liquor Feed	18-2 Ammonia Product	18-3 Stripped Gas Liquor	18-4 Acid Gas to Stretford
CO_2, lb/hr	22,161		10	22,151
H_2S, lb/hr	148		-	148
H_2O, lb/hr	942,920	34	941,930	956
Phenols, lb/hr	301		301	
Fatty Acids, lb/hr	1,106		1,106	
Ammonia, lb/hr	6,783	6,364	419	
HCl, lb/hr	185		185	
TOTAL, lb/hr	973,604	6,398	943,951	23,255
Pressure, psia	100	220	145	20
Temperature, °F	120	100	100	120

NOTE: Flow quantities, pressures, and temperatures shown are for the total unit on a stream-day basis, are to be used solely for process design purposes, and are not necessarily the conditions which will be attained during actual operations.

TABLE 6.2.12-2

EQUIPMENT LIST

AMMONIA RECOVERY - UNIT 18

Item No.	Equipment Name	Number Required Oper.	Number Required Spare
18VL-0101	Sour Water Stripper	2	0
18VL-0102	Absorber	2	0
18VL-0103	Stripper	2	0
18VL-0104	Fractionator	2	0
18VL-0105	IPE Solvent Stripper	2	0
18DM-0101	Fractionator Feed Tank	2	0
18DM-0102	Reflux Drum	2	0
18TK-0101	Phosphoric Acid Tank	1	0
18E-0101	Stripped Liquor Cooler	2	0
18E-0102	Sour Water Stripper Reboiler	2	0
18E-0103	Liquor Feed Preheater	2	0
18E-0104	Absorber Solvent Pump-Around Cooler	2	0
18E-0105	Solution Exchanger	2	0
18E-0106	Lean Solution Cooler	2	0
18E-0107	Stripper Reboiler	2	0
18E-0108	Stripper Condenser I	2	0
18E-0109	Stripper Condenser II	2	0
18E-0110	Fractionator Reboiler	2	0
18E-0111	Fractionator Condenser	2	0

TABLE 6.2.12-2 (Continued)

EQUIPMENT LIST

AMMONIA RECOVERY - UNIT 18

Item No.	Equipment Name	Number Required Oper.	Spare
18P-0101 A/B	Stripped Liquor Pump	2	2
18P-0102 A/B	Absorber Bottoms Pump	2	2
18P-0103 A/B	Fractionator Feed Pump	2	2
18P-0104 A/B	Ammonia Pump	2	2
18P-0105	Phosphoric Acid Pump	1	0
18P-0106 A/B	Sour Water Stripper Feed Pump	2	2

NOTE: Train No. 2 equipment numbers which are not shown are the same as indicated above except the train designation is 02 instead of 01.

Example:	Train No. 1	Train No. 2
	18VL-0101	18VL-0201

SECTION 6.2.13
SULFUR RECOVERY - UNIT 19

6.2.13.1 DESIGN BASIS

Purpose of Unit

The purpose of the Sulfur Recovery unit is to recover sulfur from acid gas streams generated in the process area. The majority of the acid gas is recovered in the Lurgi selective Rectisol gas purification unit from raw gas produced in the Gasification and POX units. Other streams treated are sour gases from Gas Liquor Separation, Ammonia Recovery, and Naphtha Hydrotreating. Molten sulfur is recovered as a saleable product.

Scope of Unit

The unit includes the following sections:

ADIP - for hydrocarbon removal and H_2S enrichment
Claus - for recovery of sulfur from high H_2S concentration streams
SCOT - for conversion of sulfur compounds in Claus tail gas to H_2S
Stretford - for removal of H_2S from low H_2S concentration streams

The configuration of the sections is given in Figure 6.2.13-1. Each section is described in more detail below.

General Design Criteria

Operation of the Sulfur Recovery unit is compatible with the overall plant onstream factor of 332 days per year. Spare equipment is provided in critical services based on licensor current practice. All rotating equipment are motor driven. Of the sulfur entering the unit, 98.7 percent is recovered as saleable grade yellow sulfur.

6.2.13.1 (Continued)

Feedstocks

Table 6.2.13-1 gives the compositions, flow rates, and battery limit conditions of the primary feedstocks.

Possible contaminants include CH_3OH and NH_3 in the Rectisol and Ammonia Recovery unit gases respectively.

Products

Sulfur

Sulfur is produced at the rate of 87.2 tons per day and leaves the unit at 285°F and 50 psia. The product specification is as follows:

Purity	99.8 wt % (min)
Carbon Content	0.06 wt % (max)
Ash Content	0.06 wt % (max)
Free Acidity	0.003 wt % (max)
Free of As, Se, Te	
Color	Bright Yellow
State	Molten

6.2.13.1 (Continued)

Flue Gas

Flue gas to be incinerated leaves the unit at the rate of 11,838.1 lb-mol/hr at 100°F and 18 psia. The gas has the following composition:

Component	Mol %
CO_2	87.9
H_2S	10 ppmv
COS	240 ppmv
CO	0.2
H_2	0.6
CH_4	0.3
C_2+	0.6
N_2	4.5
H_2O	5.9

Utilities

Steam		
100 psig	21,900	lb/hr produced
50 psig	22,900	lb/hr consumed (net)
BFW	29,300	lb/hr consumed
Condensate	1,500	lb/hr consumed
Cooling Water	2,900	gpm (Δ T = 30°F)
Fuel Gas	11.6	MM Btu/hr consumed
Reducing Gas (99.5% H_2)		12.6 lb-mol/hr
Electric Power	3,000	kW

6.2.13.1 (Continued)

ADIP Section

The ADIP section treats the H_2S rich gas from the Rectisol unit. A portion of the gas bypasses the ADIP absorption column to maintain temperature control in the Claus furnace.

The maximum H_2S content in the ADIP reject gas stream is 1000 ppm. The reject stream is further treated in the Stretford section.

The H_2S enriched gas stream which is fed to the Claus section contains 40 mol percent H_2S.

The ADIP section consists of two 50 percent capacity trains. Each train may be scheduled for a shutdown, of a seven days maximum duration, once every eighteen (18) months for routine inspection and maintenance.

Claus Section

The Claus section treats the H_2S enriched gas from the ADIP section.

Ninety-three percent of the H_2S entering the Claus section is recovered as saleable grade yellow sulfur.

The section consists of two 50 percent capacity trains.

SCOT Section

The SCOT section converts the SO_2, COS, and molecular sulfur present in the Claus tailgas to H_2S.

Hydrogen from the Hydrogen Production unit is used as the reducing gas.

6.2.13.1 (Continued)

The treated gas is routed to the Stretford section for H_2S removal.

Water condensed in the quench column is routed to Phenosolvan for treatment.

The SCOT section consists of two 50 percent capacity trains. A train turndown to 40 percent of design can be achieved.

Stretford Section

Feed to the Stretford section consists of:

- H_2S lean gas from Rectisol unit
- Sour gas from Naphtha Hydrotreating unit
- Off gas from Ammonia Recovery unit
- Expansion gas from Gas Liquor Separation unit
- Reject gas from the ADIP Section
- Tailgas from the SCOT Section

H_2S is removed to 10 ppmv in the vent gas. H_2S conversion to sulfur is 99.9 percent.

The section consists of two 50 percent capacity trains.

FIGURE 6.2.13-1

SULFUR RECOVERY SCHEME

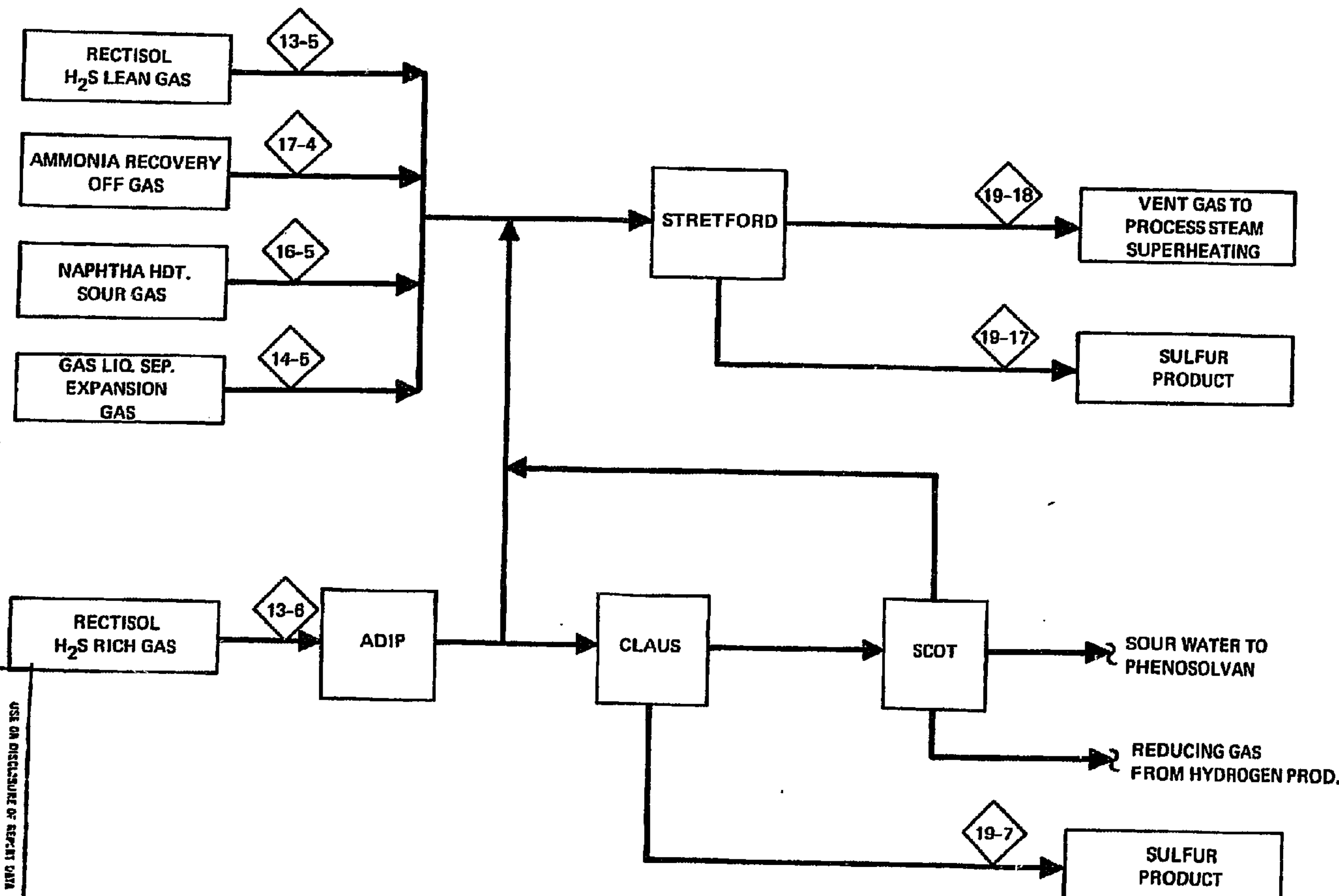


TABLE 6.2.13-1

FEEDSTOCKS

Stream	13-6	13-5	14-5	17-4	16-5
Source	Rectisol	Rectisol	Gas Liquor Separation	Ammonia Recovery	Naphtha Hydrotreating
Destination	ADIP	Stretford	Stretford	Stretford	Stretford
Component	lb-mol/hr	lb-mol/hr	lb-mol/hr	lb-mol/hr	lb-mol/hr
CO_2	717.7	9,046.8	116.0	503.3	0.1
H_2S	150.0	67.6	3.0	4.3	1.3
COS	1.7	1.5			
CO		18.7	8.6		0.1
H_2		35.4	17.7		6.5
CH_4		22.5	7.4		2.2
C_2H_6	0.4	27.8			1.0
C_3H_8	10.5	21.2			0.6
C_4H_{10}	7.2	1.8			0.4
N_2					1.6
Total Dry Gas	887.5	9,243.3	152.7	507.6	13.8
H_2O	13.1	-	9.5	53.0	-
Total Wet Gas	900.6	9,243.3	162.2	560.6	13.8
Dry Gas, lb/hr	37,696	403,382	5,603	22,299	223
H_2O, lb/hr	236	-	172	956	-
Naphtha, lb/hr	211	-	-	-	-
TOTAL, lb/hr	38,143	403,382	5,775	23,255	223
Pressure, psia	25	23	20	20	20
Temp., °F	80	75	135	120	100

6.2.13.2 PROCESS DESCRIPTION

The Sulfur Recovery unit consists of four sections using the following processes: ADIP, licensed by Shell; Claus, designed by Fluor; SCOT, licensed by Shell; and Stretford, licensed by Peabody Holmes. The ADIP section treats the H_2S-rich gas from Rectisol to enrich the H_2S content. Sulfur is recovered from the enriched gas in the Claus Section. The SCOT Section converts sulfur compounds in the Claus tailgas to H_2S. The Stretford section removes H_2S from the following streams: SCOT tailgas, ADIP offgas, Rectisol H_2S-lean gas, Gas Liquor Separation expansion gas, Ammonia Recovery offgas, and Naphtha Hydrotreating sour gas.

Process flow sketches are presented on Drawing Numbers 835704-19-4-101, 835704-19-R-102, 835704-19-4-103, and 835704-19-4-104. The unit material balance (Table 6.2.13-2) and equipment list follow the drawings. The plot plan is shown on Drawing No. 835704-19-4-050.

ADIP Section

The major portion of the H_2S-rich gas from Rectisol enters the ADIP absorber. The remainder by-passes the ADIP process and flows directly to the Claus section.

In the absorber H_2S and CO_2 are absorbed countercurrently by an amine solution. The H_2S level in the gas is reduced to 1,000 ppmv, and it is further treated in the Stretford section. The H_2S- and CO_2-rich amine solution flows from the bottom of the absorber, is heated, and flows to the stripper. The rich amine solution is stripped and becomes hot lean amine solution at the bottom of the stripper. A LP steam reboiler provides heat for the stripping process. The hot lean amine solution is cooled and returned to the absorber as the absorbing solution. The stripper overhead is partially condensed and separated. The condensate is pumped back to the stripper as the quench water, and the acid gas flows to the Claus section for further treatment.

6.2.13.2 (Continued)

Claus Section

Most of the acid gas from the ADIP section flows to the combustion furnace where the H_2S is partially combusted to SO_2. The gas stream which bypasses the ADIP process is used to control the temperature in the furnace. Combustion air is supplied by a blower. The ratio of acid gas to air is controlled to maintain a volumetric H_2S/SO_2 ratio of 2:1. A slipstream of the acid gas is sent to each of three reheat furnaces, providing the temperature required for the converter operation.

Hot gas from the combustion furnace passes through the waste heat boiler, producing medium pressure steam, and through the first sulfur condenser. The gas is then reheated by combusting a slip stream of the acid gas in the first reheat furnace. After leaving the first reheat furnace, the SO_2 and H_2S in the gas are reacted over catalyst in the first converter to produce sulfur. The sulfur is condensed in the second condenser against BFW, producing low pressure (LP) steam.

The gas is reheated by combusting a slip stream of the acid gas in the second reheat furnace. In the second converter SO_2 and H_2S react to form sulfur which is condensed against BFW in the third condenser. Low pressure steam is produced in the condenser.

The gas is again heated by the combustion of the remaining acid gas in the third reheat furnace. Sulfur is formed by the reaction of SO_2 and H_2S in the third converter and is condensed in the fourth condenser. The tail-gas which contains COS and CS_2 in addition to unreacted H_2S and SO_2 flows to the SCOT section for further processing.

Liquid sulfur flows by gravity from the condensers to the sulfur pit. Sulfur is transferred by pump to the tank farm for dispatch. The sulfur

6.2.13.2 (Continued)

lines and seal legs are steam jacketed, and the sulfur pit is heated by steam coils.

SCOT Section

Tailgas from the Claus Section enters the SCOT furnace where it is heated by the combustion of fuel gas. The hot gas is mixed with hydrogen gas before entering the SCOT reactor. The sulfur compounds in the gas are reduced to H_2S over the catalyst bed. The gas is cooled in a waste heat boiler, producing LP steam, and further cooled by direct contact with quench water in the quench column. The gas then flows to the Stretford section where the H_2S is removed. The condensate formed during cooling is pumped to the Phenosolvan unit for treatment.

Stretford Section

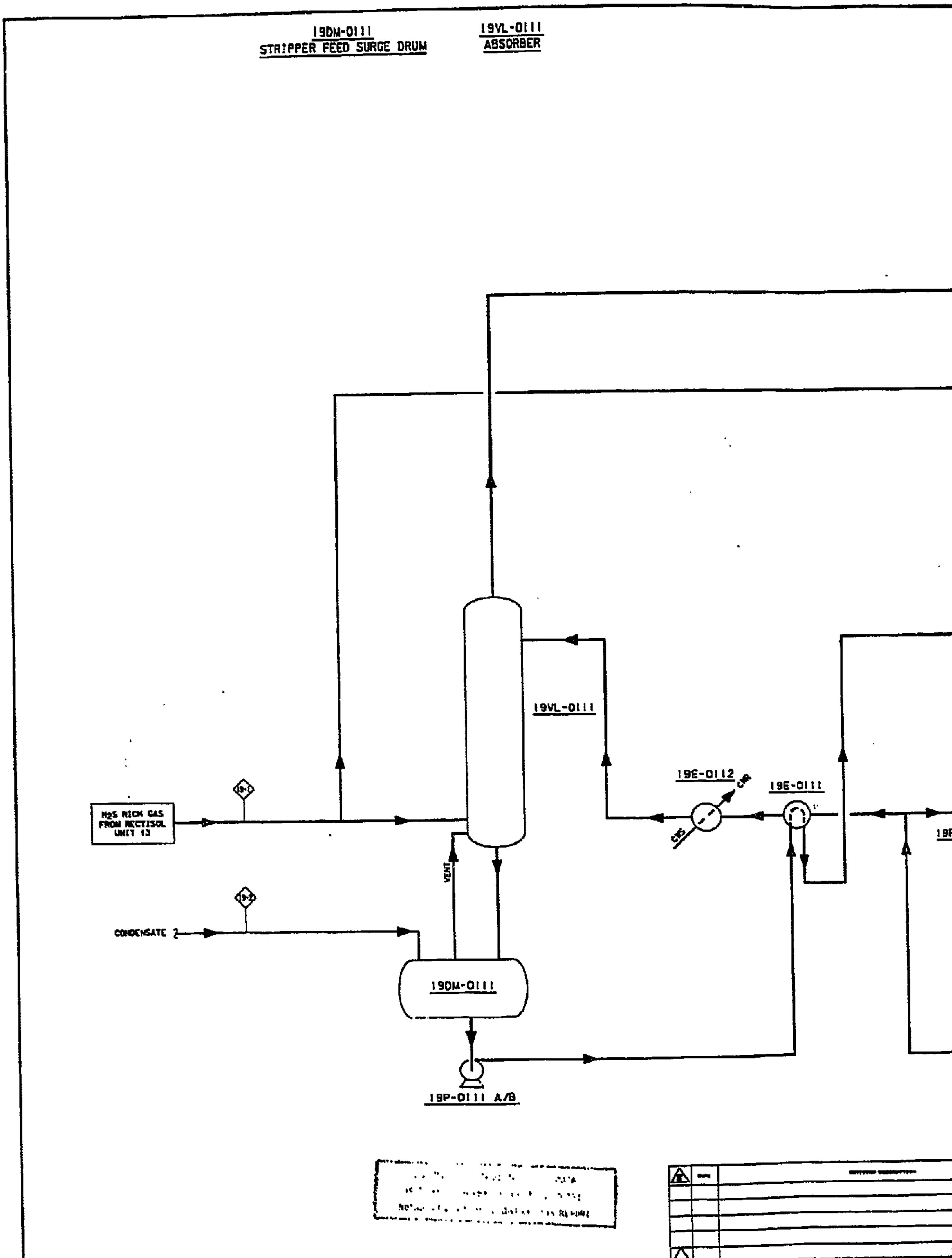
Gases containing H_2S from Rectisol, Gas Liquor Separation, Naphtha Hydrotreating, Ammonia Recovery, and the ADIP and SCOT sections of Sulfur Recovery are combined before entering the Stretford process.

The gas is washed with regenerated Stretford solution in the venturi absorber/reaction tank. The H_2S is selectively absorbed into the alkaline solution. The hydrosulfide formed reacts with vanadium in the Stretford solution and is oxidized to elemental sulfur. The liquor is regenerated by air blowing in the oxidizer tanks; the reduced vanadium is regenerated through a mechanism transferring oxygen via anthraquinone disulfonic acid (ADA). The sulfur is removed from the solution by air flotation, producing a froth containing the sulfur. The froth overflows the oxidizing tank into a skim tank. Next, it is heated to produce the molten sulfur product. Vent gas from the Stretford unit, containing volatile organic carbon (VOC), H_2S, and COS is routed to the Process Steam Superheater for combustion.

6.2.13.2 (Continued)

VOC is converted to CO_2, and the remaining H_2S and COS are converted to SO_2.

A portion of the absorbed sulfur forms sodium sulfate and sodium thiosulfate salts which are broken down to recover useful constituents. Reaction of sulfate and thiosulfate salts is accomplished in the fixed salt recovery area (FSR) by heating a slipstream of the solution. The FSR converts the sulfur compounds to H_2S for recovery in the main absorber. Valuable vanadium salt is recovered. The waste water purge stream associated with the Stretford process is eliminated.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STRIPPER FEED SURGE DRUM
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ABSORBER
19VL-0111
19E-0112
CWR
CWS
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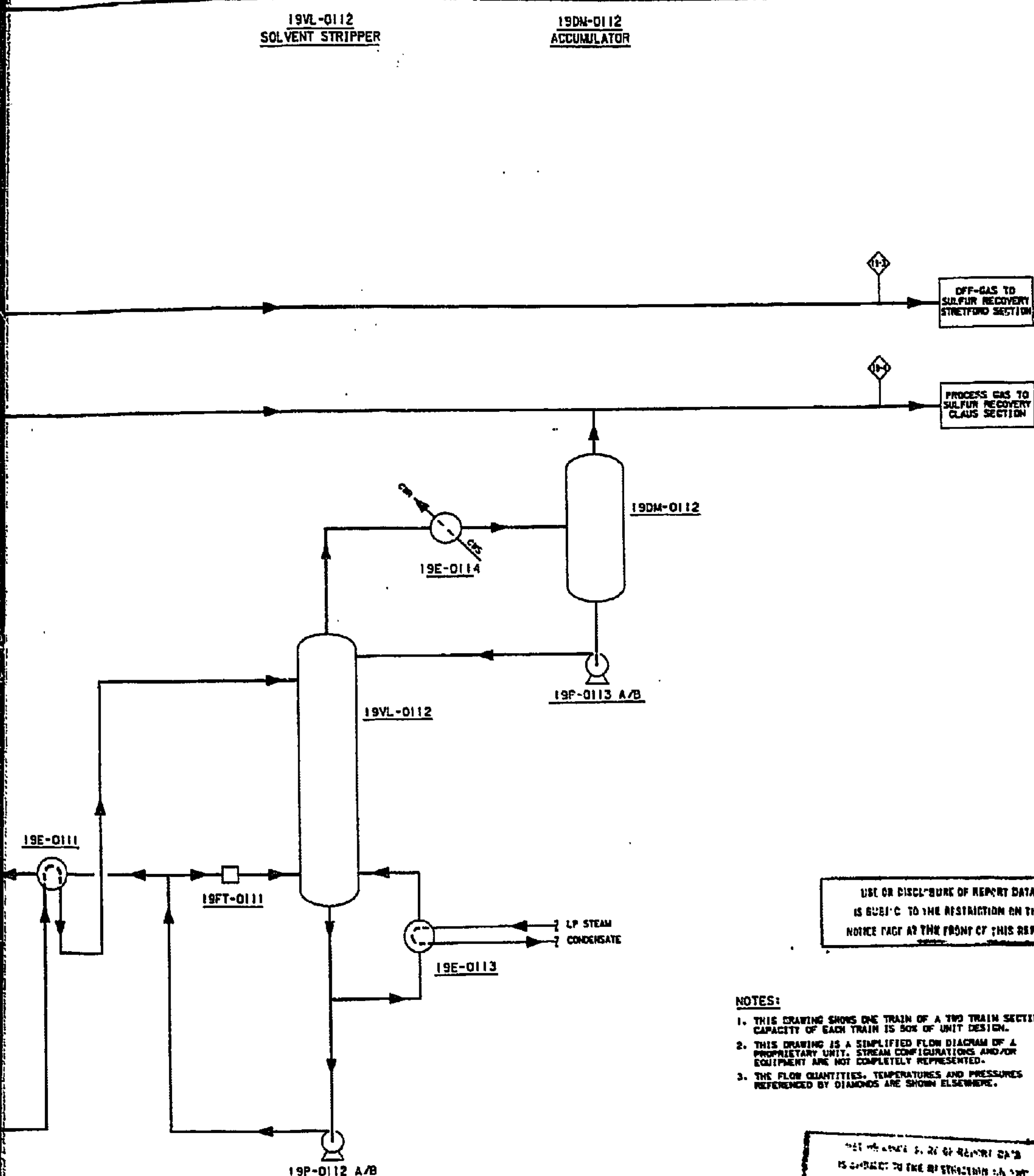
H2S RICH GAS
FROM RECTISOL
UNIT 13
CONDENSATE
VENT
19DM-0111
19P-0111 A/B



NOTES:

1. THIS DRAWING SHOWS ONE TRAIN OF A TWO TRAIN SECTION. CAPACITY OF EACH TRAIN IS 50% OF UNIT DESIGN.
2. THIS DRAWING IS A SIMPLIFIED FLOW DIAGRAM OF A PROPRIETARY UNIT. STREAM CONFIGURATIONS AND/OR EQUIPMENT ARE NOT COMPLETELY REPRESENTED.
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PROCESS FLOW DIAGRAM
SULFUR RECOVERY-ADIP SECTION
UNIT 19

CROW TRIBE OF INDIANS — SYNFUELS FEASIBILITY STUDY

NONE | 835704-19-4-101 | 1

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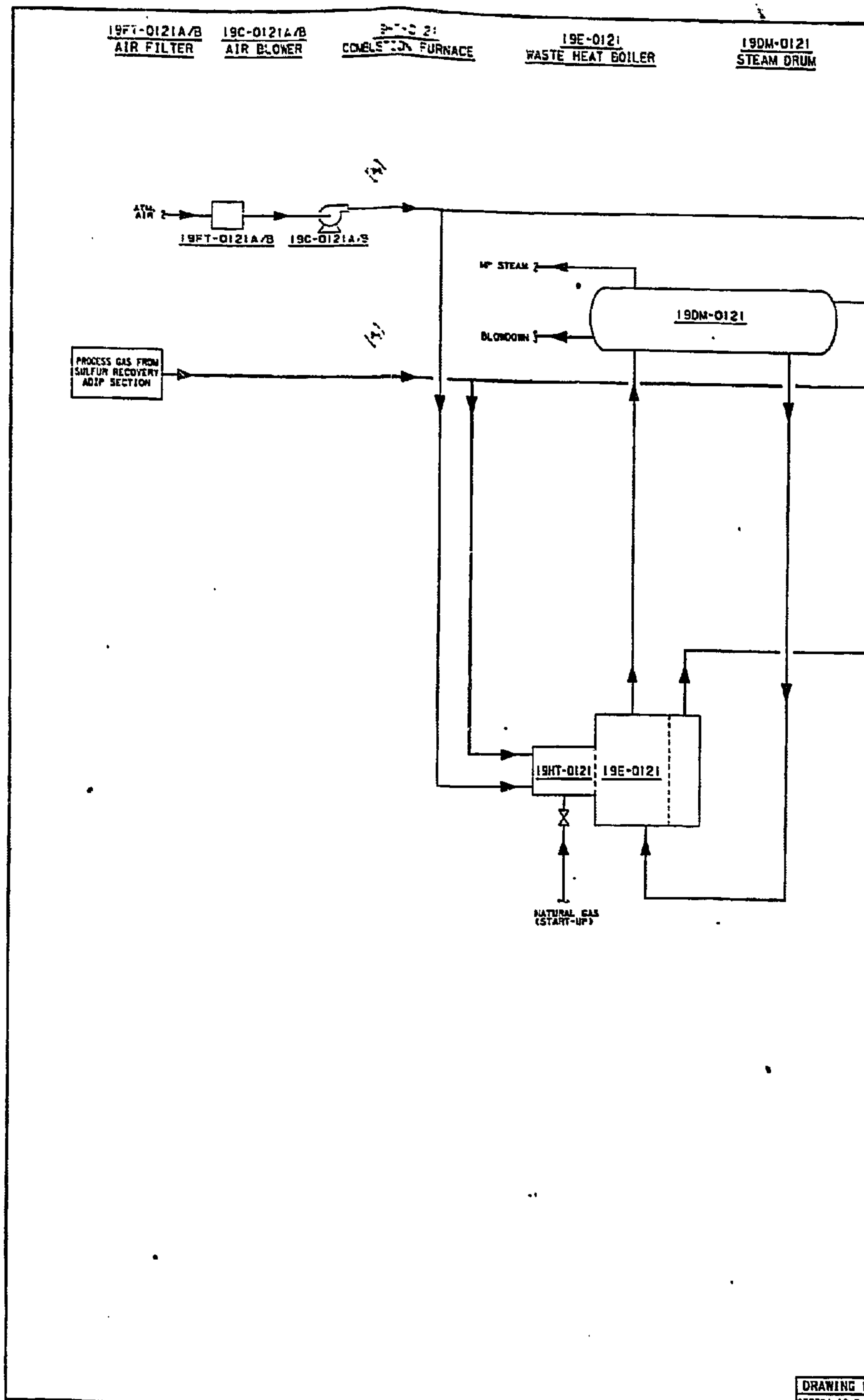
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AIR BLOWER
FURNACE
19E-0121
WASTE HEAT BOILER
19DM-0121
STEAM DRUM
ATM. AIR
19FT-0121A/B
19C-0121A/B
MP STEAM
BLOWDOWN
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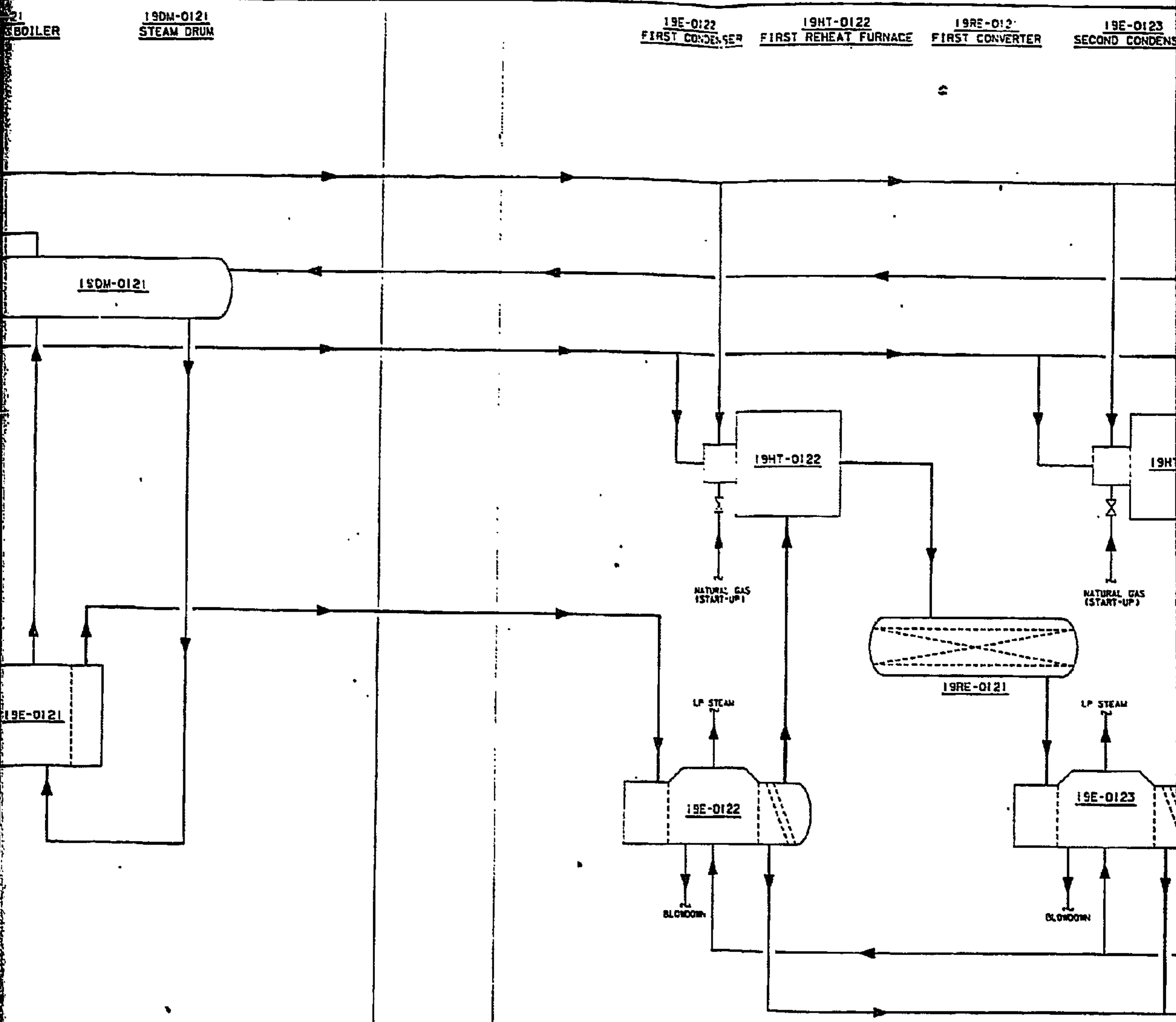
PROCESS GAS FROM SULFUR RECOVERY ADIP SECTION
19HT-0121
19E-0121
NATURAL GAS (START-UP)
DRAWING



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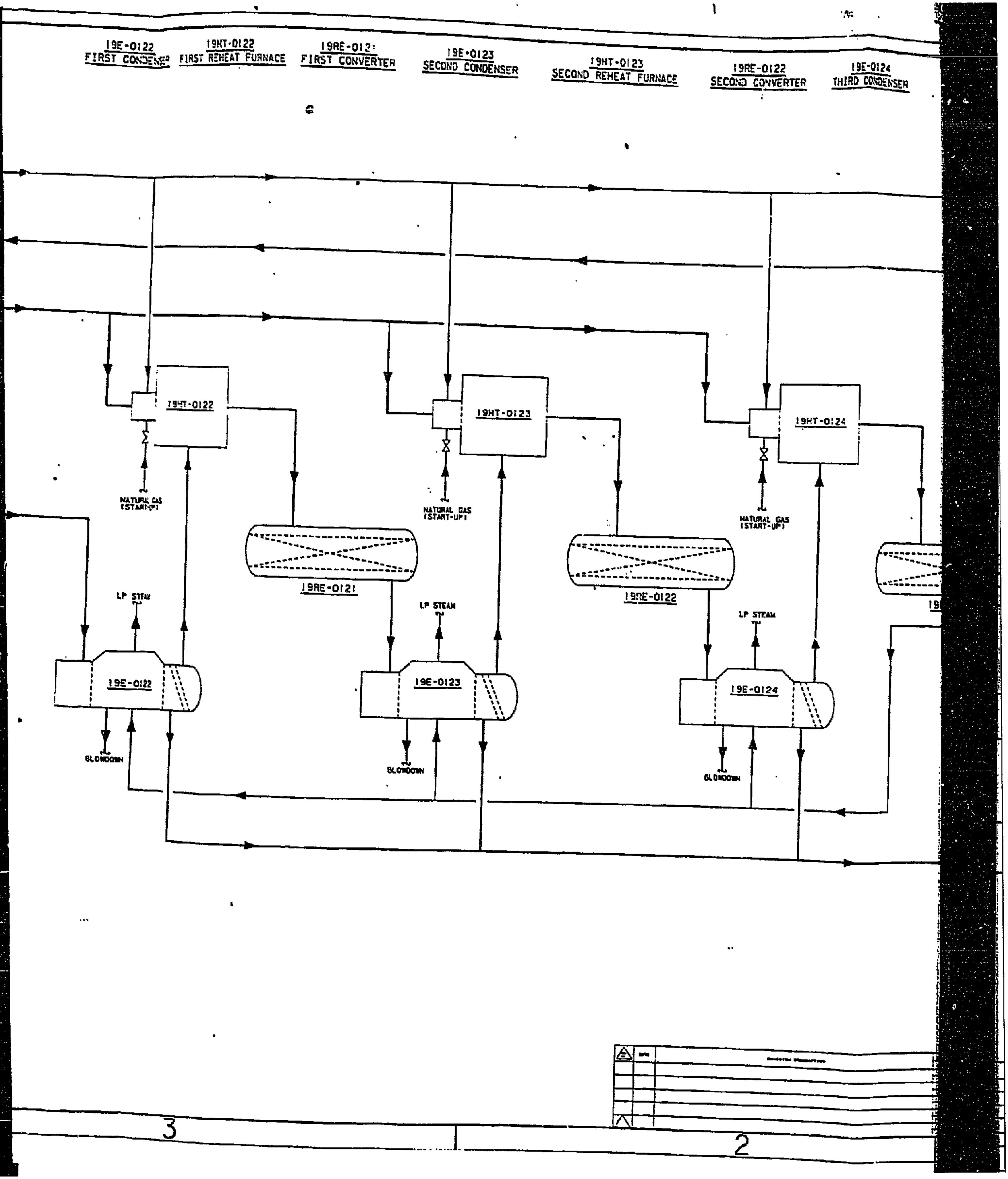
835704-19-R-102	1	2 OF 2

4

3



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FIRST CONDENSER
19HT-0122
FIRST REHEAT FURNACE
19RE-012
FIRST CONVERTER
19E-0123
SECOND CONDENSER
19HT-0123
SECOND REHEAT FURNACE
19RE-0122
SECOND CONVERTER
19E-0124
THIRD CONDENSER
19HT-0122
19HT-0123
19HT-0124
NATURAL GAS (START-UP)
NATURAL GAS (START-UP)
NATURAL GAS (START-UP)
19RE-0121
19RE-0122
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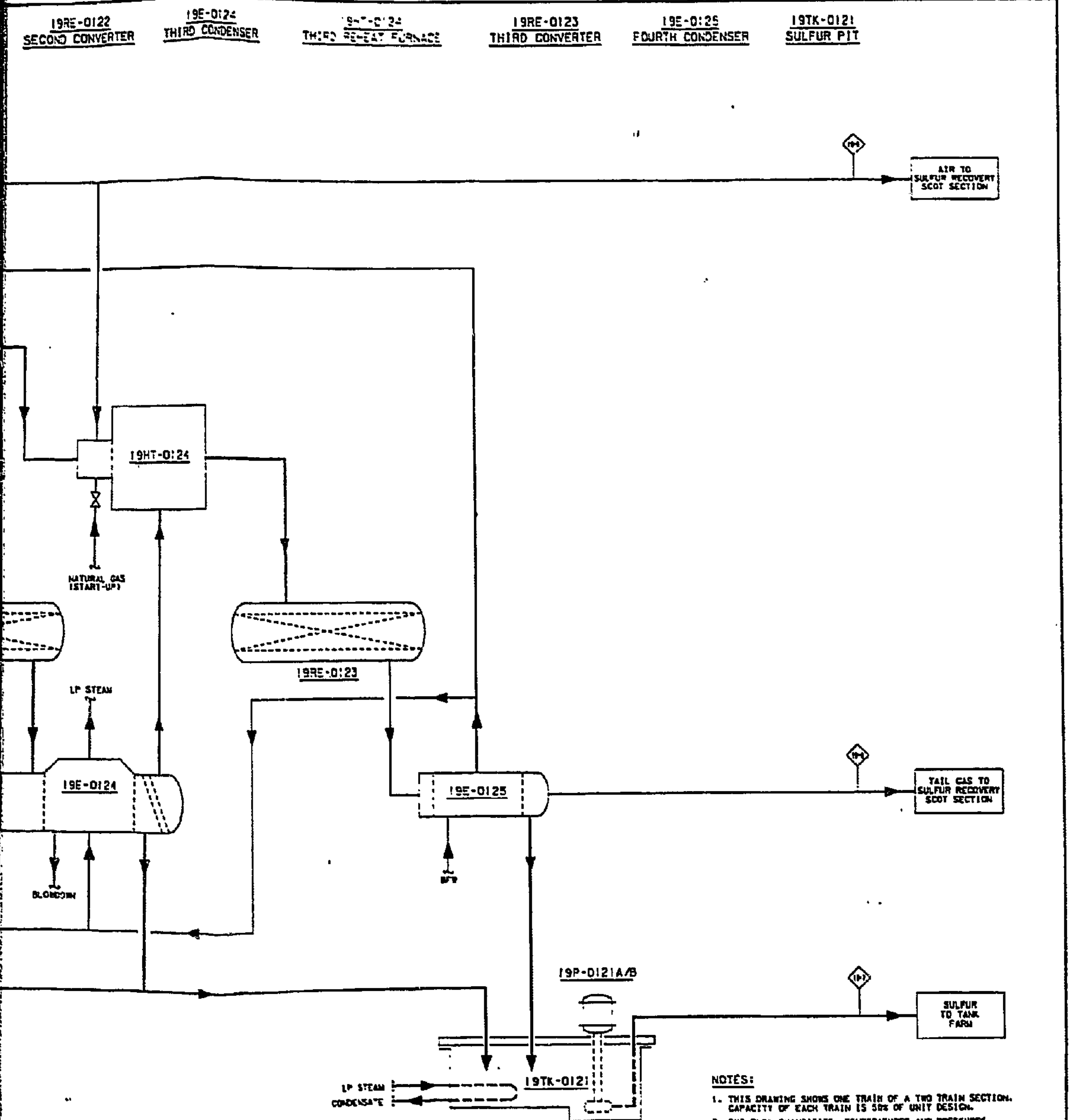
BLOWDOWN
BLOWDOWN
BLOWDOWN
3
2



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PROCESS FLOW DIAGRAM
SULFUR RECOVERY - CLAUS SECTION
UNIT 19

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NONE

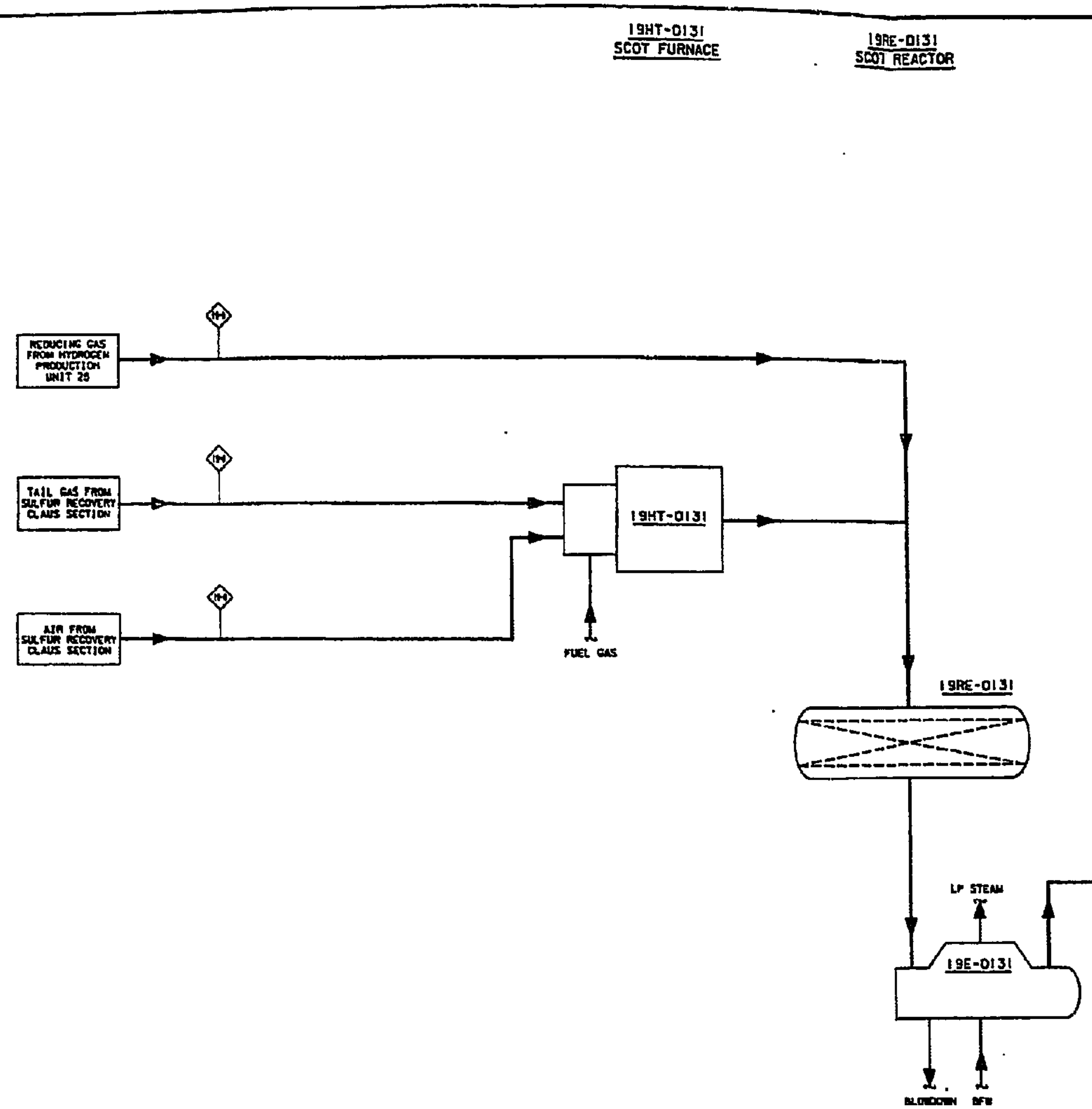
835704-19-R-102

MIRCOFILM FRAME 1 OF 2

1

03 35719102

2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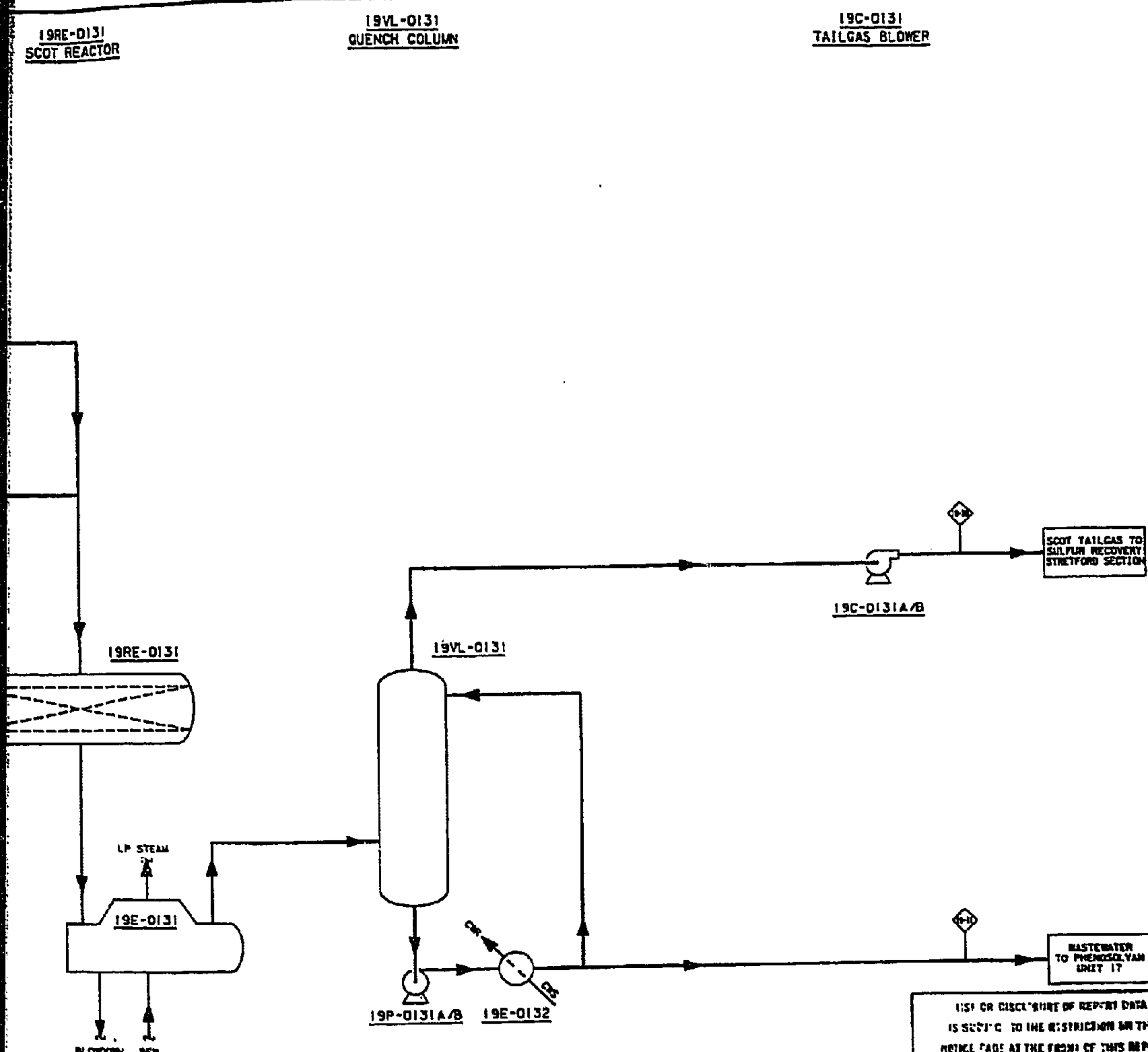
SCOT FURNACE
19RE-0131
SCOT REACTOR
REDUCING GAS FROM HYDROGEN PRODUCTION UNIT 25
TAIL GAS FROM SULFUR RECOVERY CLAUS SECTION
AIR FROM SULFUR RECOVERY CLAUS SECTION
19HT-0131
FUEL GAS
19RE-0131
LP STEAM
19E-0131
BLOWDOWN
BFW



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NOTES:

1. THIS DRAWING SHOWS ONE TRAIN OF A TWO TRAIN SECTION. CAPACITY OF EACH TRAIN IS 50% OF UNIT DESIGN.
2. THIS DRAWING IS A SIMPLIFIED FLOW DIAGRAM OF A PROPRIETARY UNIT. STREAM CONFIGURATIONS AND/OR EQUIPMENT ARE NOT COMPLETELY REPRESENTED.
3. THE FLOW QUANTITIES, TEMPERATURES AND PRESSURES REFERENCED BY DIAMONDS ARE SHOWN ELSEWHERE.

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FLUOR

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R. LANG

PROCESS FLOW DIAGRAM
SULFUR RECOVERY-SCOT SECTION
UNIT 19

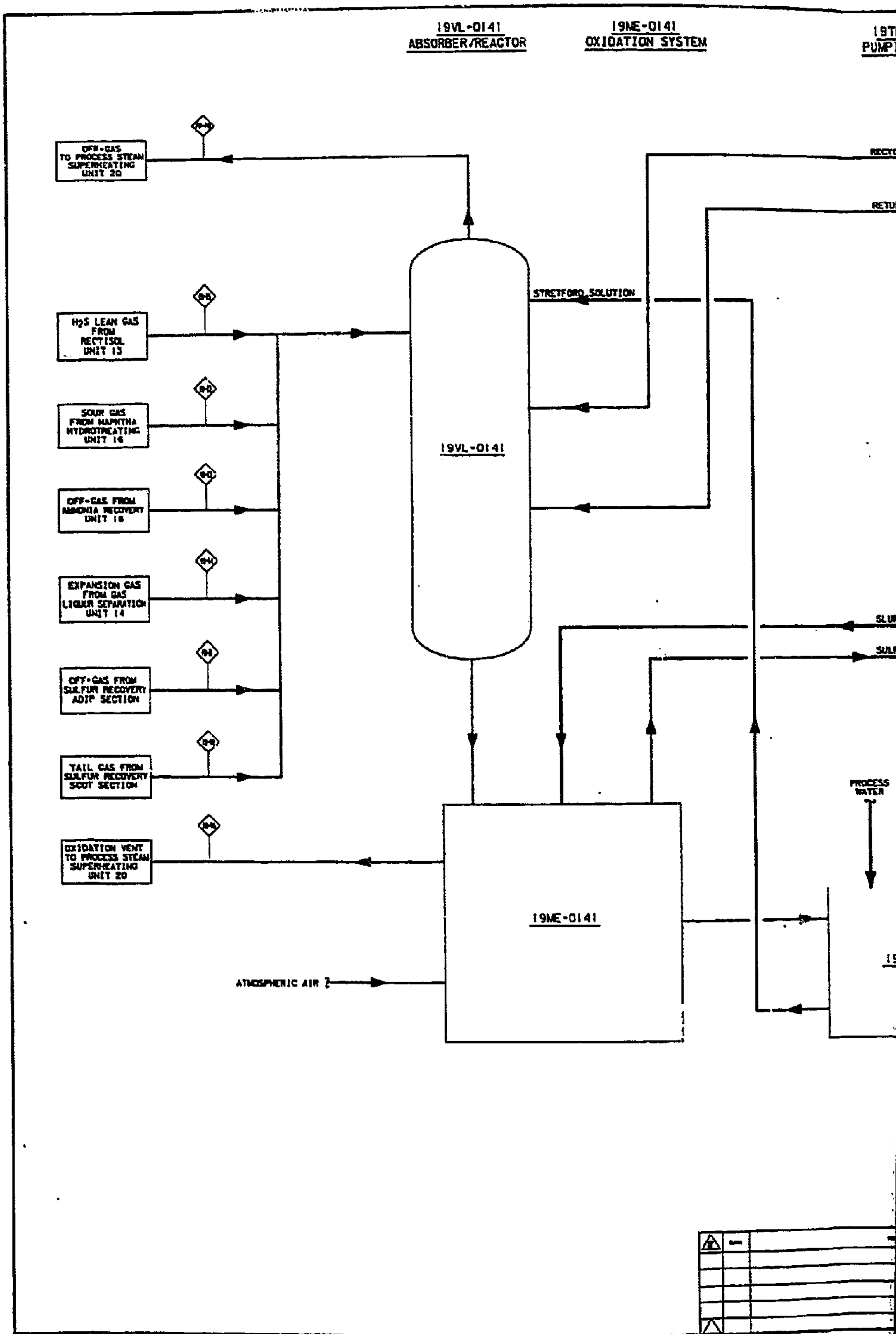
CROW TRIBE OF INDIANS — SYNFUELS FEASIBILITY STUDY

SCALE: NONE

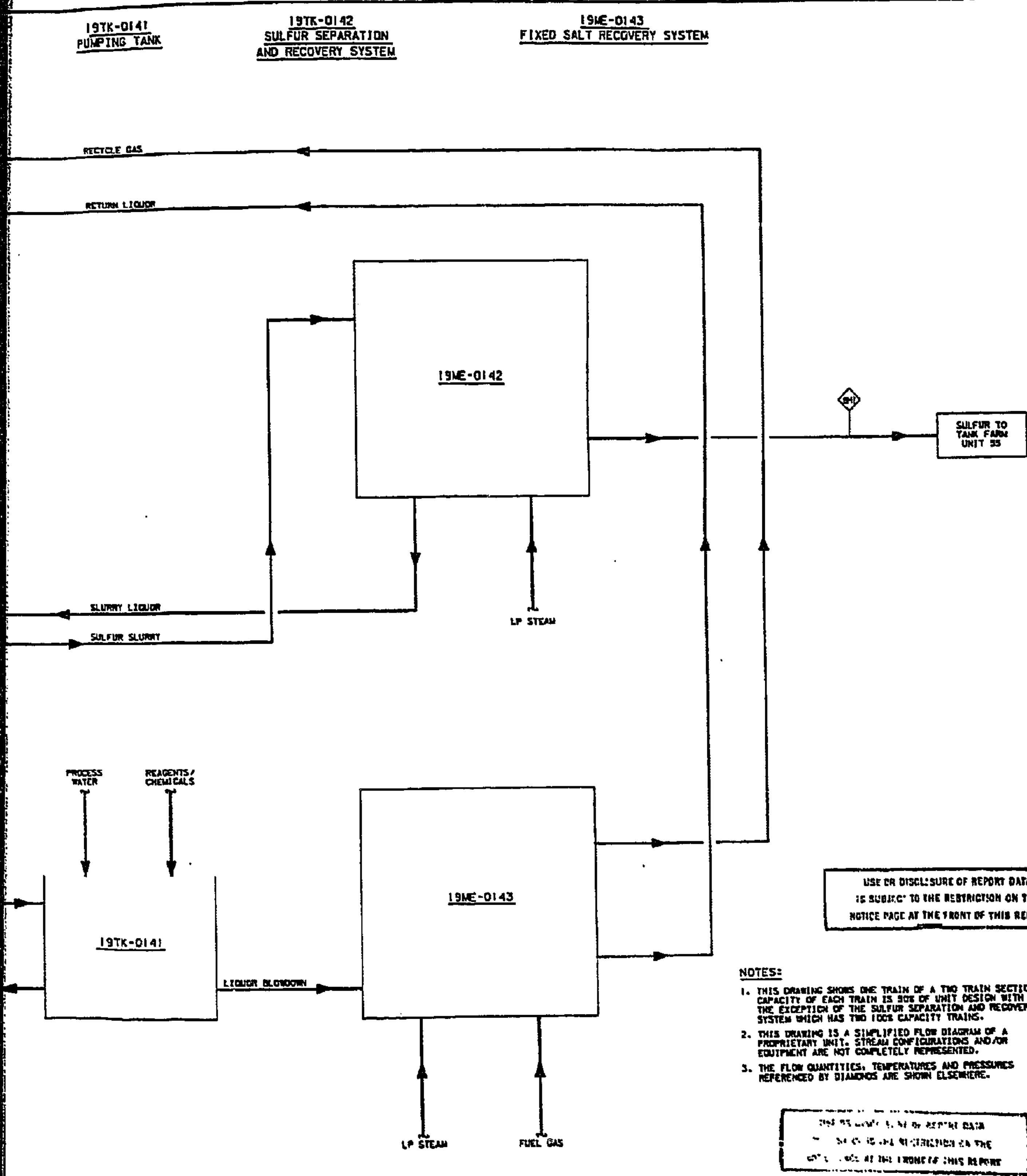
835704-19-4-103

1

001 3571910



19VL-0141
ABSORBER/REACTOR
19ME-0141
OXIDATION SYSTEM
OFF-GAS TO PROCESS STEAM SUPERHEATING UNIT 20
H_2S LEAN GAS FROM RECTISOL UNIT 13
SOUR GAS FROM NAPHTHA HYDROTREATING UNIT 16
OFF-GAS FROM AMMONIA RECOVERY UNIT 18
EXPANSION GAS FROM GAS LIQUOR SEPARATION UNIT 14
OFF-GAS FROM SULFUR RECOVERY ADIP SECTION
TAIL GAS FROM SULFUR RECOVERY SCOT SECTION
OXIDATION VENT TO PROCESS STEAM SUPERHEATING UNIT 20
STRETFORD SOLUTION
19VL-0141
19ME-0141
ATMOSPHERIC AIR
PROCESS WATER



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NOTES:

1. THIS DRAWING SHOWS ONE TRAIN OF A TWO TRAIN SECTION. CAPACITY OF EACH TRAIN IS 50% OF UNIT DESIGN WITH THE EXCEPTION OF THE SULFUR SEPARATION AND RECOVERY SYSTEM WHICH HAS TWO 100% CAPACITY TRAINS.
2. THIS DRAWING IS A SIMPLIFIED FLOW DIAGRAM OF A PROPRIETARY UNIT. STREAM CONFIGURATIONS AND/OR EQUIPMENT ARE NOT COMPLETELY REPRESENTED.
3. THE FLOW QUANTITIES, TEMPERATURES AND PRESSURES REFERENCED BY DIAMONDS ARE SHOWN ELSEWHERE.

FLUOR

D.P. HALVERSON
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PROCESS FLOW DIAGRAM
SULFUP RECOVERY-STRETFORD SECTION
UNIT 19

CROW TRIBE OF INDIANS — SYNFUELS FEASIBILITY STUDY

NONE | 835704-19-4-104 | 1

003 357191O4

TABLE 6.2.13-2

MATERIAL BALANCE

SULFUR RECOVERY - UNIT

Stream Number	19-1		19-2		19-3	
Stream Name	H_2S-Rich Gas From Rectisol		Condensate		Offgas to Stretford	
Component	lb-mol/hr	Mol%	lb-mol/hr	Mol%	lb-mol/hr	Mol%
H_2						
N_2						
O_2						
CO						
CH_4						
C_2H_6	0.4	0.05			0.3	0.06
C_3H_8	10.5	1.18			8.1	1.57
C_4H_{10}	7.2	0.81			5.4	1.04
CO_2	717.7	80.86			501.9	96.98
H_2S	150.0	16.91			0.5	0.10
COS	1.7	0.19			1.3	0.25
SO_2						
S_6-S_8						
Total Dry Gas	887.5	100.00			517.5	100.00
H_2O	13.1				29.5	
Total Wet Gas	900.6				547.0	
Dry Gas, lb/hr	37,696				22,864	
H_2O, lb/hr	236		470		531	
Naphtha, lb/hr	211					
Sulfur, lb/hr						
TOTAL, lb/hr	38,143		470		23,395	
Pressure, psia	25		60		22	
Temperature, °F	80		290		100	

NOTE: Flow quantities, pressures and temperatures shown for the total unit on a purposes and are not necessarily the conditions which will be attained during

TABLE 6.2.13-2

MATERIAL BALANCE

SULFUR RECOVERY - UNIT 19

19-2		19-3		19-4		19-5		19-6	
Condensate		Offgas to Stretford		Process Gas to Claus		Atmospheric Air		Air to SCOT	
b-mol/hr	Mol%	lb-mol/hr	Mol%	lb-mol/hr	Mol%	lb-mol/hr	Mol%	lb-mol/hr	Mol%
						530.1	79.00	80.6	79.00
						140.9	21.00	21.4	21.00
		0.3	0.06	0.1	0.03				
		8.1	1.57	2.4	0.65				
		5.4	1.04	1.8	0.48				
		501.9	96.98	215.8	58.31				
		0.5	0.10	149.5	40.42				
		1.3	0.25	0.4	0.11				
		517.5	100.00	370.0	100.00	671.0	100.00	102.0	100.00
		29.5		9.7		13.4		2.0	
	-	547.0		379.7		684.4		104.0	
		22,864		14,832		19,360		2,943	
470		531		175		241		36	
						211			
470		23,395		15,218		19,601		2,979	
60		22		28		24		24	
290		100		100		180		180	

peratures shown for the total unit on a streamday basis, are to be used solely for process design e conditions which will be attained during actual operations.

TABLE 6.2.13-2 (Cont

MATERIAL BALANC

SULFUR RECOVERY - U

Stream Number	19-7		19-8		19-9	
Stream Name	Sulfur to Tank Farm		Tailgas to SCOT		Reducing Gas from Hydrogen Production	
Component	lb-mol/hr	Mol%	lb-mol/hr	Mol%	lb-mol/hr	Mol%
H_2			0.1	0.01	12.5	99.2
N_2			449.5	63.73	0.1	0.79
O_2						
CO			0.03			
CH_4						
C_2H_6						
C_3H_8						
C_4H_{10}						
CO_2			246.0	34.88		
H_2S			5.5	0.78		
COS			0.8	0.11		
SO_2			3.2	0.45		
S_6-S_8			0.2	0.03		
Total Dry Gas			705.3	100.00	12.6	100.0
H_2O			192.9			
Total Wet Gas			898.3			
Dry Gas, lb/hr			23,905		29	
H_2O, lb/hr			3,475			
Naphtha, lb/hr						
Sulfur, lb/hr	4,460					
TOTAL, lb/hr	4,460		27,380		29	
Pressure, psia	50		18		900	
Temperature, °F	285		280		380	

TABLE 6.2.13-2 (Continued)

MATERIAL BALANCE

SULFUR RECOVERY - UNIT 19

	19-8		19-9		19-10		19-11		19-12	
	Tailgas to SCOT		Reducing Gas from Hydrogen Production		SCOT Tail-gas to Stretford		Wastewater to Phenosolvan		Offgas from Naphtha Hydrotreating	
	-mol/hr	Mol%	lb-mol/hr	Mol%	lb-mol/hr	Mol%	lb-mol/hr	Mol%	lb-mol/hr	Mol%
	0.1	0.01	12.5	99.21	7.8	0.97			6.5	47.10
	49.5	63.73	0.1	0.79	530.1	65.87			1.6	11.60
	0.03								0.1	0.72
							2.2	15.94		
							1.0	7.25		
							0.6	4.35		
							0.4	2.90		
	246.0	34.88			256.0	31.81			0.1	0.72
	5.5	0.78			10.9	1.35			1.3	9.42
	0.8	0.11								
	3.2	0.45								
	0.2	0.03								
	705.3	100.00	12.6	100.00	804.8	100.00			13.8	100.00
	192.9				61.8				0	
	898.3				866.8				13.8	
	23,905		29		26,505				223	
	3,475				1,113		3,133			
	27,380		29		27,618		3,133		223	
	18		900		20		65		20	
	280		380		100		100		100	

TABLE 6.2.13-2 (Continue

MATERIAL BALANCE

SULFUR RECOVERY - UNIT

Stream Number	19-13		19-14		19-15	
Stream Name	Offgas from Ammonia Recovery		Expansion Gas from Gas Liquor Separation		H_2S Lean Gas from Rectisol	
Component	lb-mol/hr	Mol%	lb-mol/hr	Mol%	lb-mol/hr	Mol%
H_2			17.7	11.59	35.4	0.38
N_2						
O_2			127.0	16.08		
CO			8.6	5.63	18.7	0.20
CH_4			7.4	4.85		
C_2H_6						
C_3H_8					21.2	0.23
C_4H_{10}					1.8	0.02
CO_2	503.3	99.15	116.0	75 97	9,046.8	97.88
H_2S	4.3	0.85	3.0	1.96	67.6	0.73
COS					1.5	0.02
SO_2						
S_6-S_8						
Total Dry Gas	507.6	100.00	152.7	100.00	9,243.3	100.00
H_2O	53.0		9.5			
Total Wet Gas	560.6		162.2			
Dry Gas, lb/hr	22,299		5,603		403,382	
H_2O, lb/hr	956		172			
Naphtha, lb/hr						
Sulfur, lb/hr						
TOTAL, lb/hr	23,255		5,775		403,382	
Pressure, psia	20		20		23	
Temperature, °F	120		135		75	

TABLE 6.2.13-2 (Continued)

MATERIAL BALANCE

SULFUR RECOVERY - UNIT 19

19-14		19-15		19-16		19-17		19-18	
nsion Gas Gas Liquor paration		H_2S Lean Gas from Rectisol		Oxidation Vent		Sulfur to Tank Farm		Vent Gas to Process Steam Superheater	
ol/hr	Mol%	lb-mol/hr	Mol%	lb-mol/hr	Mol%	lb-mol/hr	Mol%	lb-mol/hr	Mol%
.7	11.59	35.4	0.38					67.4	0.60
				649.4	82.20			531.7	4.77
.0	16.08								
.6	5.63	18.7	0.20					27.4	0.25
.4	4.85							32.1	0.29
								29.1	0.26
		21.2	0.23					29.9	0.27
		1.8	0.02	13.6	1.72			7.6	0.07
.0	75.97	9,046.8	97.88					10,409.8	93.46
.0	1.96	67.6	0.73					0.1	
		1.5	0.02					2.8	0.03
.7	100.00	9,243.3	100.00	790.0	100.00			11,137.9	100.00
.5				11.5				700.2	
.2				801.5				11,838.1	
5,603		403,382		22,856				477,256	
172				207				12,615	
						2,805			
5,775		403,382		23,063		2,805		489,871	
20		23		13.5		50		18	
135		75		104		285		100	

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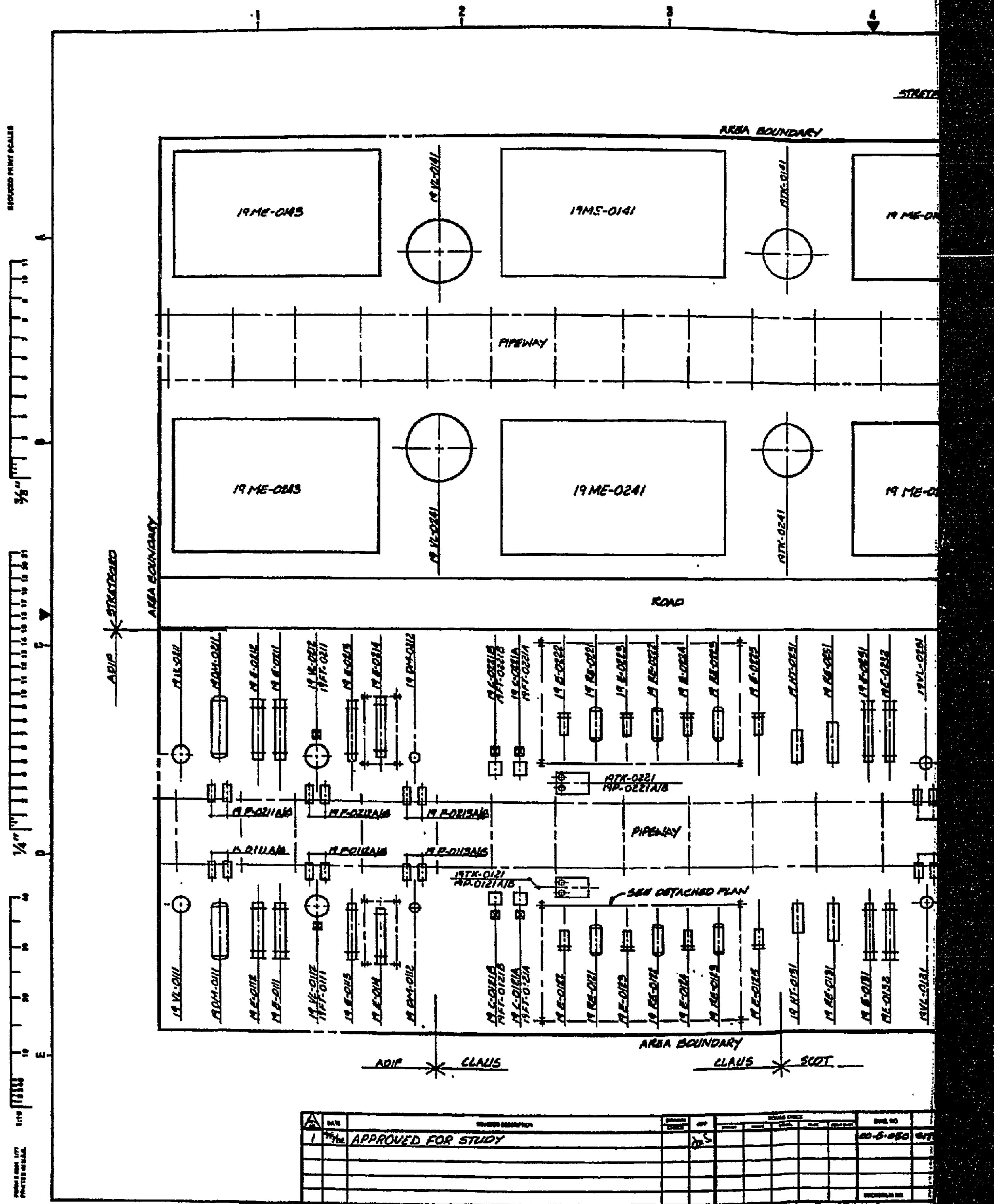
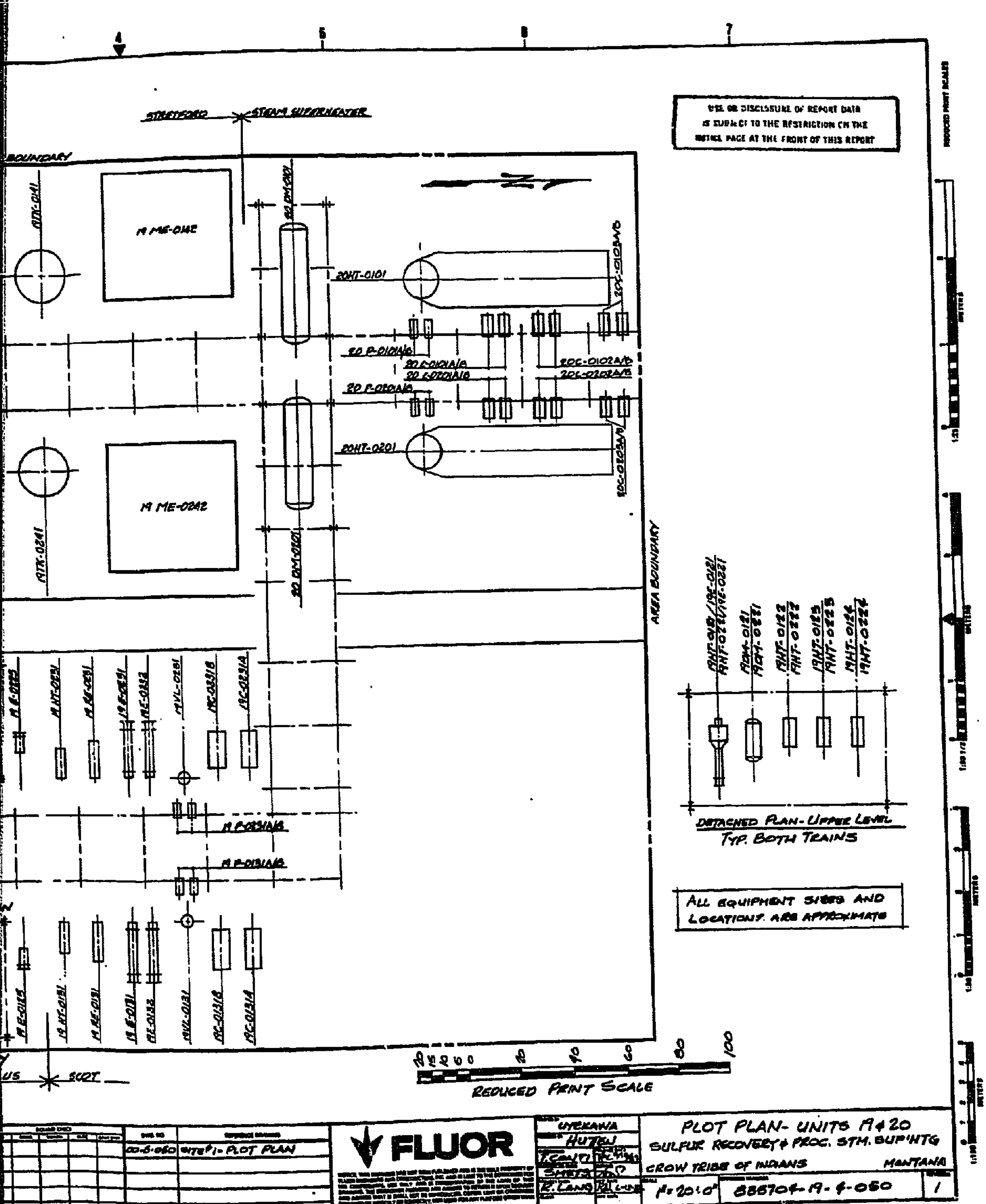


AREA BOUNDARY
19ME-0143
19MS-0141
19 ME-0243
19 ME-0241
PIPEWAY
ROAD
ADIP
AREA BOUNDARY
19TK-0221
19P-0221A/B
PIPEWAY
19TK-0121
19P-0121A/B
SEE DETACHED PLAN
AREA BOUNDARY
ADIP
CLAUS
CLAUS
SCOT
APPROVED FOR STUDY



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STRETFORD
STEAM SUPERHEATER
BOUNDARY
19 ME-0142
19 ME-0242
20HT-0101
20HT-0201
20 P-0101A/B
20 P-0201A/B
20C-0102A/B
20C-0202A/B
AREA BOUNDARY
19 P-0231A/B
19 P-0131A/B
DETACHED PLAN-UPPER LEVEL
TYP. BOTH TRAINS
ALL EQUIPMENT SIZES AND LOCATIONS ARE APPROXIMATE
REDUCED PRINT SCALE
SCOT
FLUOR
00-6-050
SITE #1- PLOT PLAN
UYEKAWA
HUTEN
T.CONTI
R.LANG
PLOT PLAN- UNITS 19 & 20
SULFUR RECOVERY & PROC. STM. SUP'HTG
CROW TRIBE OF INDIANS
MONTANA
1" = 20'-0"
886704-19-4-050
1

TABLE 6.2.13-3

EQUIPMENT LIST

SULFUR RECOVERY - UNIT 19

Item No.	Equipment Name	Number Required Oper.	Number Required Spare
ADIP SECTION			
19DM-0111	Stripper Feed Surge Drum	2	0
19DM-0112	Accumulator	2	0
19VL-0111	Absorber	2	0
19VL-0112	Solvent Stripper	2	0
19E-0111	Lean/Rich Solvent Exchanger	2	0
19E-0112	Lean Solvent Cooler	2	0
19E-0113	Stripper Reboiler	2	0
19E-0114	Stripper Reflux Cooler	2	0
19FT-0111	Filter	2	0
19P-0111 A/B	Stripper Feed Pump	2	2
19P-0112 A/B	Lean Solvent Pump	2	2
19P-0113 A/B	Reflux Pump	2	2
CLAUS SECTION			
19DM-0121	Steam Drum	2	0
19TK-0121	Sulfur Pit	2	0
19RE-0121	First Converter	2	0
19RE-0122	Second Converter	2	0
19RE-0123	Third Converter	2	0
19HT-0121	Combustion Furnace	2	0
19HT-0122	First Reheat Furnace	2	0

TABLE 6.2.13-3 (Continued)

EQUIPMENT LIST

SULFUR RECOVERY - UNIT 19

Item No.	Equipment Name	Number Required Oper.	Number Required Spare
CLAUS SECTION (Cont'd.)			
19HT-0123	Second Reheat Furnace	2	0
19HT-0124	Third Reheat Furnace	2	0
19E-0121	Waste Heat Boiler	2	0
19E-0122	First Condenser	2	0
19E-0123	Second Condenser	2	0
19E-0124	Third Condenser	2	0
19E-0125	Fourth Condenser	2	0
19FT-0121 A/B	Air Filter	2	2
19P-0121 A/B	Sulfur Pump	2	2
19C-0121 A/B	Air Blower	2	2
SCOT SECTION			
19VL-0131	Quench Column	2	2
19RE-0131	SCOT Reactor	2	0
19HT-0131	SCOT Furnace	2	0
19E-0131	Waste Heat Boiler	2	0
19E-0132	Quench Cooler	2	0
19P-0131 A/B	Quench Pump	2	2
19C-0131 A/B	Tailgas Blower	2	2

TABLE 6.2.13-3 (Continued)

EQUIPMENT LIST

SULFUR RECOVERY - UNIT 19

Item No.	Equipment Name	Number Required Oper.	Number Required Spare
STRETFORD SECTION			
19VL-0141	Absorber/Reactor	2	0
19TK-0141	Pumping Tank	2	0
19ME-0141	Oxidation System	2	0
19ME-0142	Sulfur Separation and Recovery System	2	0
19ME-0143	Fixed Salt Recovery System	2	0

NOTE: Train No. 2 equipment numbers which are not shown are the same as indicated above except the train designation is 02 instead of 01.

Example:	Train No. 1	Train No. 2
	19DM-0111	19DM-0211

6.2.14 PROCESS STEAM SUPERHEATING - UNIT 20

6.2.14.1 DESIGN BASIS

Purpose of Unit

The Process Steam Superheating unit generates high pressure steam from boiler feed water and superheats it along with the excess high pressure steam generated in the Methanation and the Partial Oxidation units. The unit also combusts the organic material and sulfur compounds in the Sulfur Recovery unit vent gas and the Rectisol unit CO_2-rich gas.

Scope of Unit

The unit consists of a gas-fired combustion chamber with attached waste heat recovery steam generators. Preheat coils for air and vent gases are also provided.

General Design Criteria

The unit consists of two trains, each with combustion chamber and waste heat recovery. The unit is compatible with the overall plant onstream factor of 332 days per year. Spare air blowers are provided.

Process Performance Objective

Complete conversion of sulfur-containing compounds to SO_2 and hydrocarbons to CO_2 is expected.

6.2.14.1 (Continued)

Feedstock

Temperature 100°F
Pressure 2 psig

Component	lb-mol/hr
CO_2	20,363.6
H_2S	0.1
COS	3.2
CO	27.4
H_2	67.4
CH_4	52.3
C_2H_6	82.3
C_3H_8	34.4
C_4H_{10}	7.6
N_2	531.7
H_2O	700.2
TOTAL	21,870.2

Product

Temperature 400°F

Component	lb-mol/hr
CO_2	21,602.1
H_2O	3,326.7
SO_2	3.3 (78 ppm vol.)
O_2	642.0
N_2	11,168.9
NO_2	4.9
CO	0.4
TOTAL	36,748.3

6.2.14.1 (Continued)

Utilities

Electric Power	600 kW
BFW Consumed	190,000 lb/hr
Steam	
650 psig, sat'd consumed	722,800 lb/hr
600 psig, 760°F produced	910,800 lb/hr

Fuel Gas 340.1 MM Btu/hr consumed

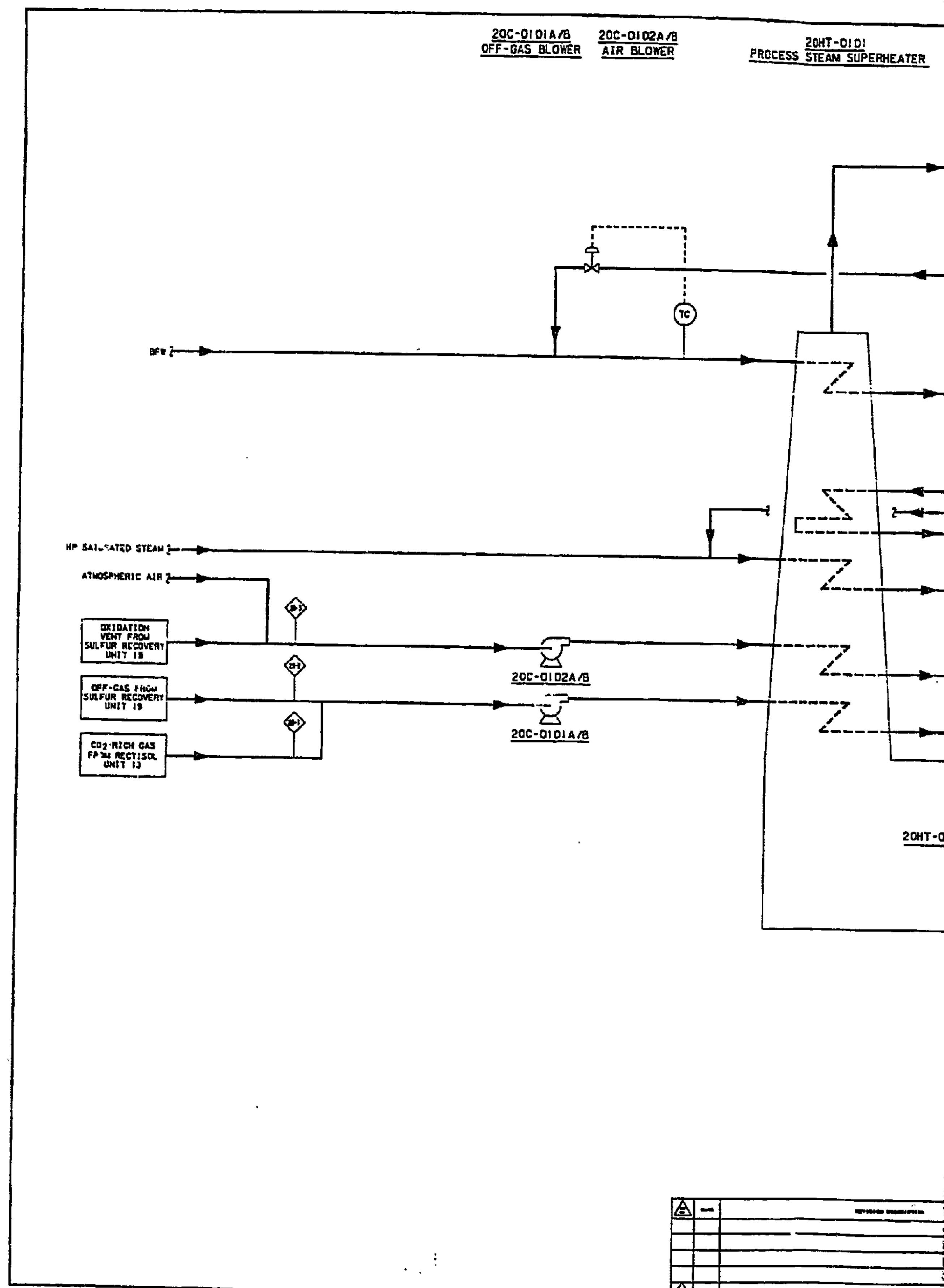
6.2.14.2 PROCESS DESCRIPTION

The process flow sketch for the Process Steam Superheating unit is presented on Drawing No. 835704-20-4-101. The unit material balance and equipment list follow (Tables 6.2.14-1 and 6.2.14-2).

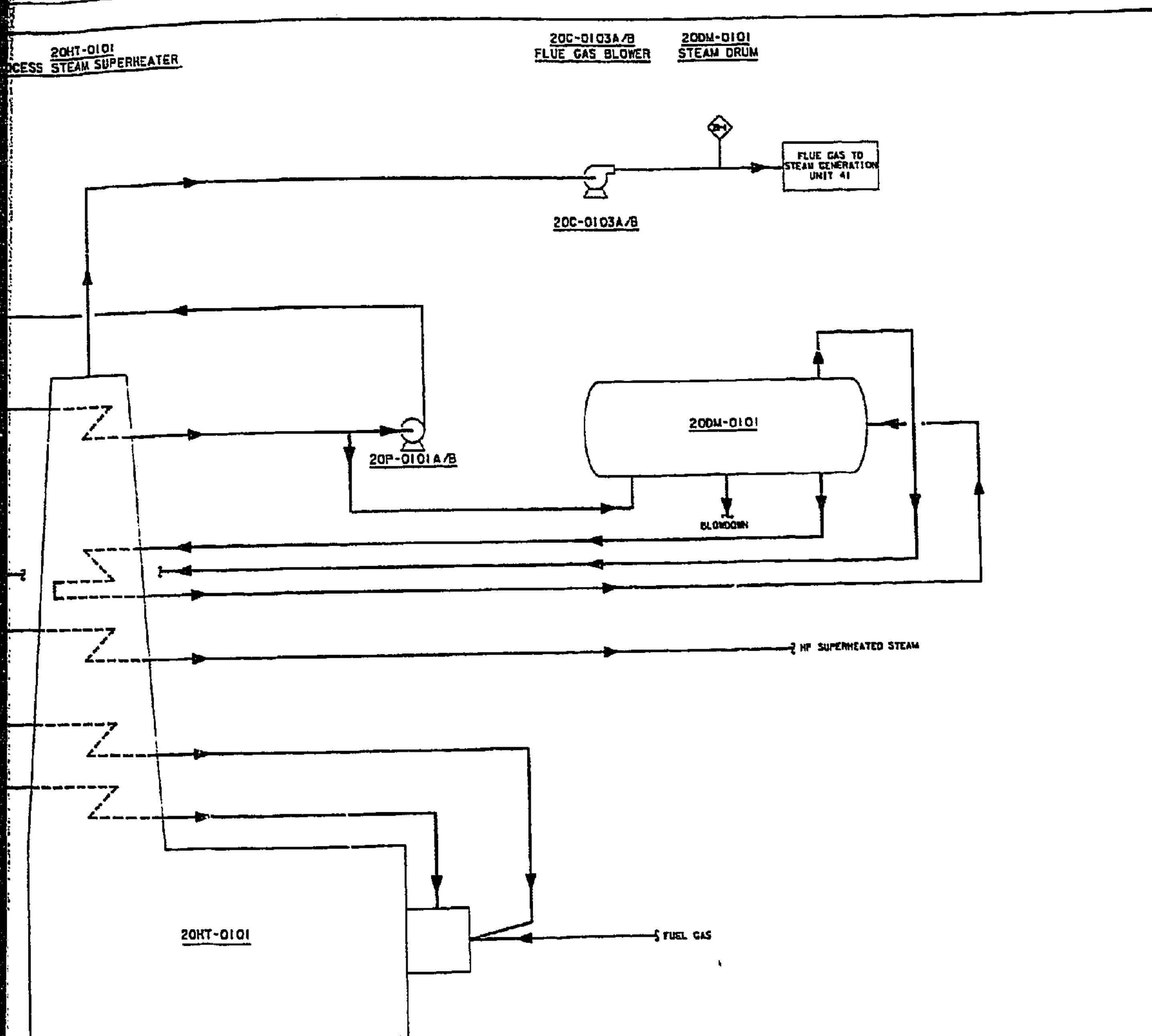
High pressure steam is generated from boiler feed water and is superheated along with the excess high pressure steam generated in the Methanation unit and the Partial Oxidation unit.

Heat is provided by the combustion of treated gas from the Sulfur Recovery unit and CO_2-rich gas from the Rectisol unit. These gas streams contain small amounts of hydrocarbon and sulfur compounds which are converted to carbon dioxide and sulfur dioxide during combustion. Since the gases do not contain adequate combustible material, assist fuel gas is used to bring the mixture up to the proper temperature.

The treated gases and combustion air are preheated by the hot flue gas to minimize the amount of assist fuel gas required. The gases are then mixed with the air and passed through the assist gas burner flame. The flue gas is cooled in the waste heat recovery section and is routed to the stack in the Steam Generation unit (41).



20C-0101A/B
OFF-GAS BLOWER
20C-0102A/B
AIR BLOWER
20HT-0101
PROCESS STEAM SUPERHEATER
TC
BFW
HP SATURATED STEAM
ATMOSPHERIC AIR
OXIDATION VENT FROM SULFUR RECOVERY UNIT 19
OFF-GAS FROM SULFUR RECOVERY UNIT 19
CO2-RICH GAS FROM RECTISOL UNIT 13
20C-0102A/B
20C-0101A/B



NOTES:

1. THIS DRAWING SHOWS ONE TRAIN OF A TWO TRAIN UNIT. CAPACITY OF EACH TRAIN IS 50% OF UNIT DESIGN.
2. THE FLOW QUANTITIES, TEMPERATURES AND PRESSURES REFERENCED BY DIAMONDS ARE SHOWN ELSEWHERE.

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FLUOR

Drawn by	R. WHITE
Checked by	C.C. ABATAY
	W. D'BELMITO
	R. McCARTHY
	R. LANG

PROCESS FLOW DIAGRAM
PROCESS STEAM SUPERHEATING
UNIT 20

CROW TRIBE OF INDIANS — SYNFUELS FEASIBILITY STUDY

Scale	Drawing Number	Rev
NONE	835704-20-4-101	1

003 3572010I

TABLE 6.2.14-1

MATERIAL BALANCE

PROCESS STEAM SUPERHEATING

Stream Number	20-1		20-2	
Stream Name	CO_2-Rich Gas From Rectisol		Vent Gas from Sulfur Recovery	
Component	lb-mol/hr	Mol%	lb-mol/hr	Mol%
H_2			67.4	0.60
N_2			531.7	4.77
CO			27.4	0.25
CH_4	20.2	0.20	32.1	0.29
C_2H_6	53.2	0.53	29.1	0.26
C_3H_8	4.5	0.04	29.9	0.27
C_4H_{10}			7.6	0.07
CO_2	9,953.8	99.22	10,409.8	93.46
H_2S	0.02		0.1	
COS	0.4	0.01	2.8	0.03
SO_2				
O_2				
NO_2				
Total Dry Gas	10,032.1	100.00	11,137.9	100.00
H_2O			700.2	
Total Wet Gas	10,032.1		11,838.1	
Dry Gas, lb/hr	440,224		477,256	
H_2O, lb/hr			12,615	
TOTAL, Lb/Hr	440,224		489,871	
Pressure, psia	25		18	
Temperature, °F	100		100	

NOTE: Flow quantities, pressures, and temperatures shown are for the total un design purposes, and are not necessarily the conditions which will be atta

TABLE 6.2.14-1

MATERIAL BALANCE

PROCESS STEAM SUPERHEATING - UNIT 20

20-2		20-3		20-4	
Vent Gas from Sulfur Recovery		Oxidant		Flue Gas to Steam Generation	
lb-mol/hr	Mol%	lb-mol/hr	Mol%	lb-mol/hr	Mol%
67.4	0.60				
531.7	4.77	10,633.7	79.19	11,168.9	33.42
27.4	0.25			0.4	0.001
32.1	0.29				
29.1	0.26				
29.9	0.27				
7.6	0.07				
10,409.8	93.46	13.6	0.10	21,602.1	64.63
0.1					
2.8	0.03				
				3.3	0.01
		2,781.1	20.71	642.0	1.92
				4.9	0.01
11,137.9	100.00	13,428.4	100.00	33,421.6	100.00
700.2		181.5		3,326.7	
11,838.1		13,609.9		36,748.3	
477,256		387,476		1,284,596	
12,615		3,270		59,934	
489,871		390,746		1,344,530	
18		13		13.5	
100		60		400	

ratures shown are for the total unit on a stream-day basis, are to be used solely for process
ly the conditions which will be attained during actual operations.

TABLE 6.2.14-2

EQUIPMENT LIST

PROCESS STEAM SUPERHEATING - UNIT 20

Item No.	Equipment Name	Number Required Oper.	Number Required Spare
20DM-0101	Steam Drum	2	0
20HT-0101	Process Steam Superheater	2	0
20P-0101 A/B	BFW Circulating Pump	2	2
20C-0101 A/B	Offgas Blower	2	2
20C-0102 A/B	Air Blower	2	2
20C-0103 A/B	Flue Gas Blower	2	2

NOTE: Train No. 2 equipment numbers which are not shown are the same as indicated above except the train designated is 02 instead of 01.

Example;	Train No. 1	Train No. 2
	20DM-0101	20DM-0201

6.2.15 METHANOL SYNTHESIS - UNIT 21

6.2.15.1 DESIGN BASIS

Purpose of Unit

Methanol Synthesis unit takes a slipstream of pure syngas from Rectisol and produces sufficient methanol to make up for the loss of methanol solvent in the Rectisol unit.

Scope of Unit

The unit includes feed compression, catalytic methanol synthesis, product cooling and flash. The unit is licensed by Lurgi Mineraloltechnik of Germany.

General Design Criteria

The unit is designed as a small single train, package unit of 30 ST/D crude methanol capacity. Since the methanol makeup rate to Rectisol is only about 6.7 ST/D, it is expected that the unit will be operated at capacity only part of the time. Consequently, minimal equipment sparing is provided in the unit.

Feedstock

Pure Syngas (net inflow, i.e. feed minus recycle to Rectisol)

Total Feed	88.9 lb-mol/hr (1,034 lb/hr)

6.2.15.1 (Continued)

Battery Limit Conditions: Temperature 68°F
Pressure 355 psia

Composition:	Dry Mol %
H_2	59.0
N_2/Ar (as N_2)	0.3
CO	16.2
CH_4	17.6
C_2H_6	0.2
CO_2	6.6

Products

Crude Methanol

Production Rate 557 lb/hr
6.7 ST/D

	MeOH	543 lb/hr
	H_2O	10 lb/hr
Dissolved CH_4 and CO_2		4 lb/hr

Battery Limit Conditions: Temperature 100°F
Pressure 75 psia

Purge and Flash Gases to Fuel Gas

Flow Rate 35.6 lb-mol/hr
(477 lbs/hr)

Battery Limit Conditions: Temperature 100°F
Pressure 75 psia

Composition:	Dry Mol %
H_2	35.6
N_2/Ar (as N_2)	1.1
CO	6.7
CH_4	50.9
C_2H_6	0.8
CO_2	4.5
MeOH	0.4
HHV	670.3 Btu/SCF

6.2.15.1 (Continued)

Utility Requirements

Electric Power	65 kW
Cooling Water (ΔT = 30°F)	80 gpm
Fuel Gas Produced	9.05 MM Btu/hr

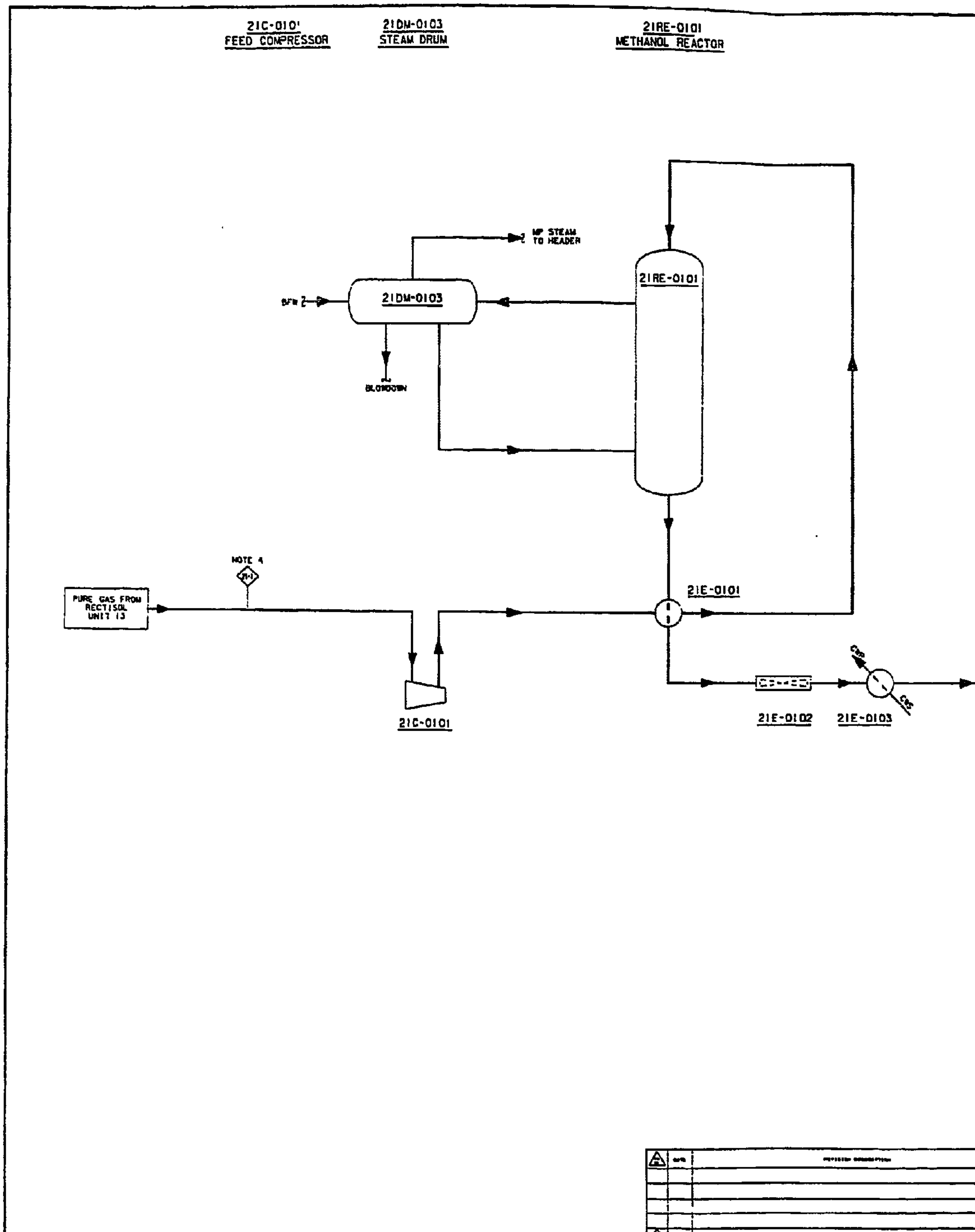
6.2.15.2 PROCESS DESCRIPTION

Drawing No. 835704-21-4-101 is the process flow sketch for the unit. The unit material balance and equipment list follow (Tables 6.2.15-1 and 6.2.15-2). The plot plan is shown on Drawing No. 835704-21-4-050.

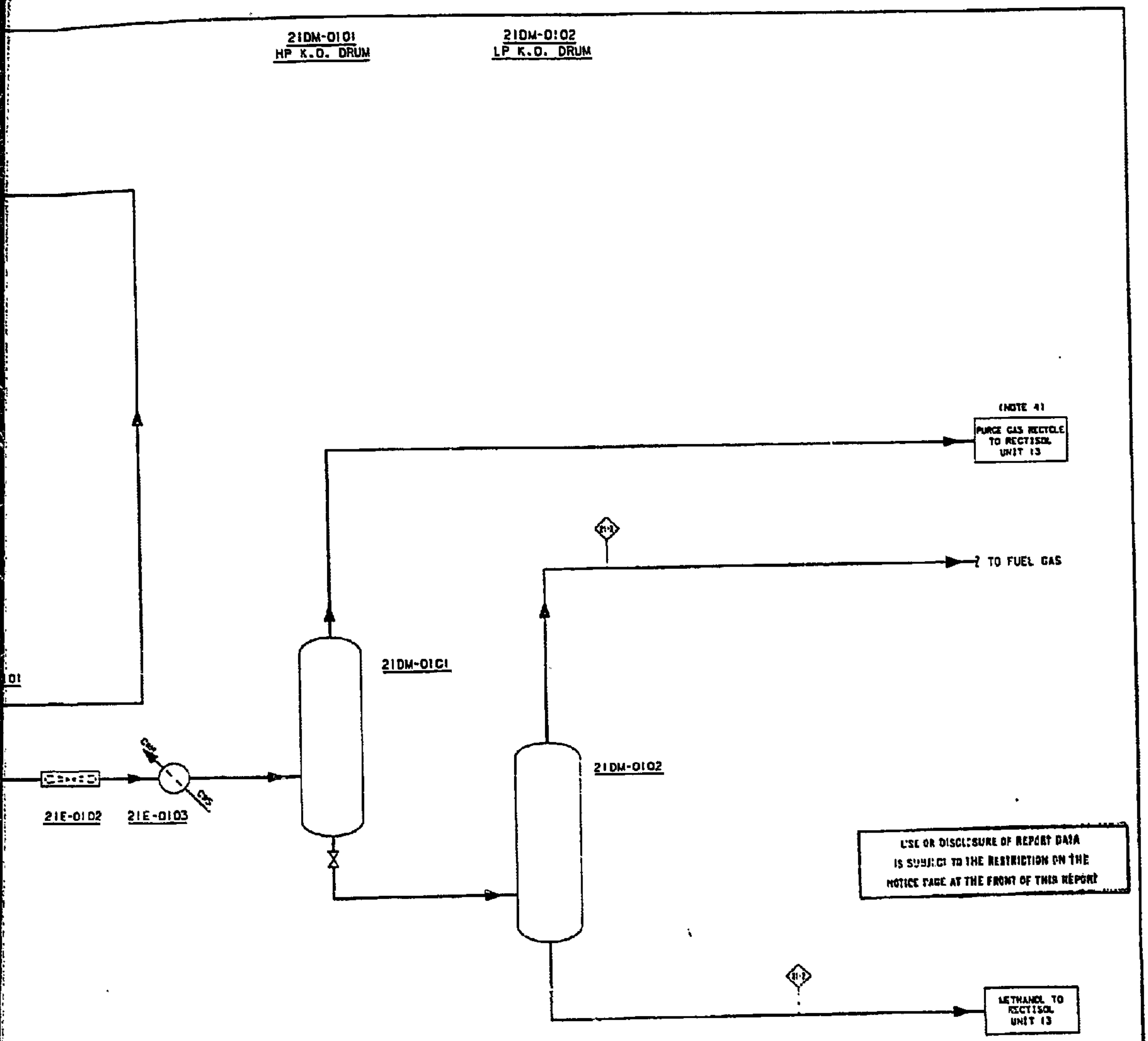
A slipstream of pure synthesis gas from the Rectisol unit is processed to produce methanol in sufficient quantity to supply makeup requirements for the Rectisol unit. The feed gas is compressed to about 740 psia, preheated in a feed-effluent exchanger and fed to the methanol synthesis reactor. The following synthesis reactions occur over the methanol synthesis catalyst:

$$CO + 2H_2 \longrightarrow CH_3OH$$
$$CO_2 + 3H_2 \longrightarrow CH_3OH + H_2O$$

Both reactions produce heat which is removed by producing steam on the shellside of the synthesis reactor. The reactor-effluent is cooled preheating the feed, air cooled, water cooled, and the liquid phase flashed. The liquid stream is further flashed at a lower pressure to remove the residual unreacted gases. The conversion to methanol per pass through the reactor is low and the vapor phase is recycled back to the Rectisol unit and combined with pure synthesis gas. The offgas from the low pressure flash is sent to the plant fuel gas system. The methanol-water mixture is sent to the Methanol Regeneration section of the Rectisol unit. The makeup methanol combines with the regenerated methanol and is used to absorb CO_2 from SNG and acid gases from raw gas.



21C-0101
FEED COMPRESSOR
21DM-0103
STEAM DRUM
21RE-0101
METHANOL REACTOR
MP STEAM
TO HEADER
BFW
21DM-0103
BLOWDOWN
21RE-0101
NOTE 4
PURE GAS FROM
RECTISOL
UNIT 13
21E-0101
21C-0101
21E-0102
21E-0103
CWR
CWS



NOTES:

1. THIS DRAWING REPRESENTS A SINGLE TRAIN METHANOL UNIT.
2. THIS DRAWING IS A SIMPLIFIED FLOW DIAGRAM OF A PROPRIETARY UNIT. STREAM CONFIGURATIONS AND/OR EQUIPMENT ARE NOT COMPLETELY REPRESENTED.
3. THE FLOW QUANTITIES, PRESSURES, AND TEMPERATURES REFERENCED BY DIAMONDS ARE SHOWN ELSEWHERE.
4. STREAM 21-1 IS THE NET FLOW FROM RECTISOL UNIT 13. (TOTAL UNIT FEED LESS PURGE GAS RECYCLED TO RECTISOL).

FLUOR

R. WHITE
C. C. ABATAY
R. O'BENITO
R. McCARTHY
R. LANG

CROW TRIBE OF INDIANS

PROCESS FLOW DIAGRAM
METHANOL SYNTHESIS
UNIT 21

SYNFUELS FEASIBILITY STUDY

NONE

835704-21-4-101

1

003 35721101

TABLE 6.2.15-1

MATERIAL BALANCE

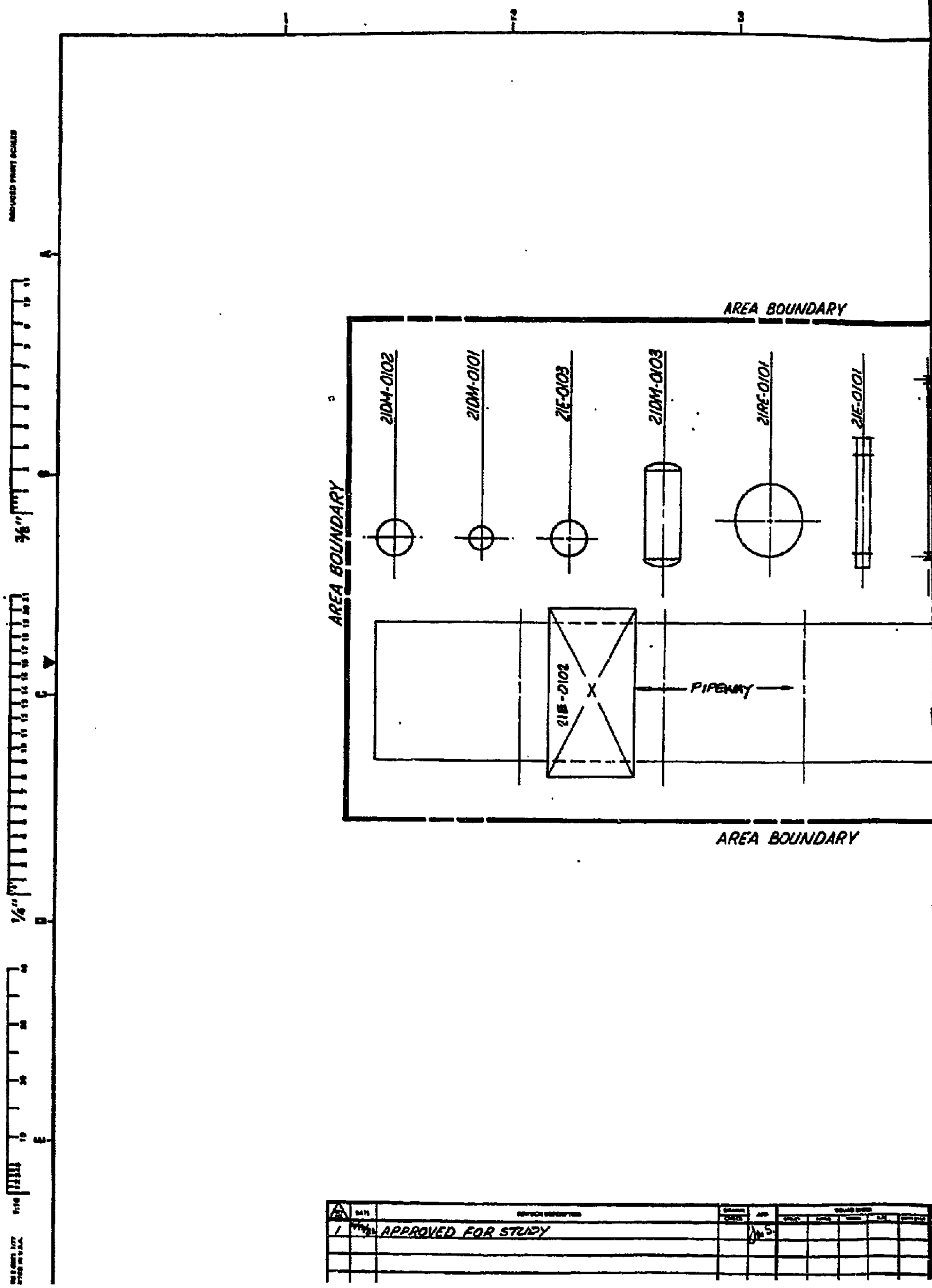
METHANOL SYNTHESIS - UNIT 21

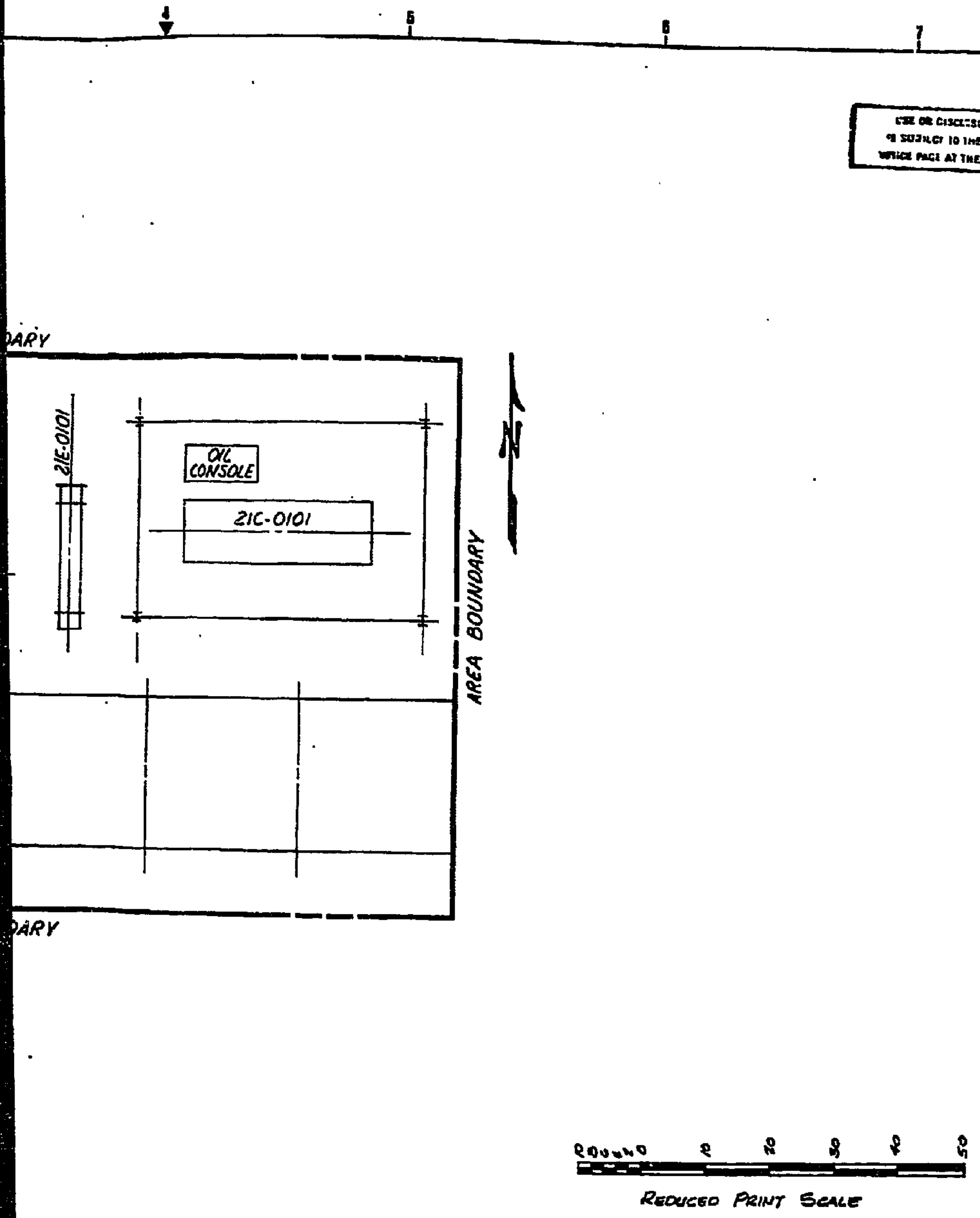
Stream Number	21-1		21-2		21-3	
Stream Name	Syngas Feed[2]		Methanol to Rectisol		Purge Gas to Fuel Gas	
COMPONENT						
H_2	52.5	59.03			12.67	35.58
N_2	0.3	0.31			0.40	1.12
CO	14.3	16.18			2.40	6.74
CH_4	15.7	17.63	0.05		18.13	50.94
C_2H_6	0.2	0.22			0.27	0.75
CO_2	5.9	6.63	0.07		1.60	4.49
CH_3OH					0.13	0.37
		100.00		100.00		100.00
Dry Gas, lb-mol/hr	88.9	100.00	0.12	100.00	35.6	100.00
H_2O vapor, lb-mol/hr	-	-0-	-	-0-	-0-	-0-
Wet Gas, lb-mol/hr	88.9	100.00	0.12[3]	100.00	35.6	100.00
Dry Gas, lb/hr	1,034		4[3]		477	
H_2O vapor, lb/hr					-	
H_2O liquid, lb/hr			10		-	
CH_3OH, liquid, lb/hr			543		-	
TOTAL, lb/hr	1,034		557		477	
Pressure, psia	355		75		75	
Temperature, °F	68		100		100	

NOTES: (1) Flow quantities, pressures, and temperatures shown are for the total unit on a stream-day basis, are to be used solely for process design purposes, and are not necessarily the conditions which will be attained during actual operations.

(2) Represents net syngas feed to the unit (total feed less purge gas recycle to Rectisol).

(3) Represents dissolved gases in the methanol liquid stream.


AREA BOUNDARY
AREA BOUNDARY
AREA BOUNDARY
21DM-0102
21DM-0101
21E-0103
21DM-0103
21RE-0101
21E-0101
21B-0102
PIPEWAY
1 APPROVED FOR STUDY



ALL EQUIPMENT SIZES AND LOCATIONS ARE APPROXIMATE.

DWG NO	REFERENCE DRAWINGS
00-5-050	SITE #1 PLOT PLAN

FLUOR

J. PARODI
T. CONTI
J. SMETS
R. LANG

PLOT PLAN - UNIT 21
METHANOL SYNTHESIS
CROW TRIBE OF INDIANS MONTANA

Scale: 1" = 10'-0"
835704-21-4-050
1

TABLE 6.2.15-2

EQUIPMENT LIST

METHANOL SYNTHESIS - UNIT 21

Item No.	Equipment Name	Number Required Oper.	Number Required Spare
21RE-0101	Methanol Reactor	1	0
21DM-0101	HP K.O. Drum	1	0
21DM-0102	LP K.O. Drum	1	0
21DM-0103	Steam Drum	1	0
21E-0101	Reactor Feed-Effluent Exchanger	1	0
21E-0102	Product Cooler	1	0
21E-0103	Product Trim Cooler	1	0
21C-0101	Feed Compressor	1	0

6.2.16 METHANATION - UNIT 22

6.2.16.1 DESIGN BASIS

Purpose of Unit

The methanation unit receives pure syngas from the Rectisol unit and catalytically reacts carbon monoxide and carbon dioxide with hydrogen to produce methane. The substitute natural gas (SNG) reaction product flows to the SNG Purification and Compression unit.

Scope of Unit

The Methanation unit encompasses the multistage methanation reaction, recycle gas compression, recovery of the heat of reaction (by generating high pressure steam), further cooling of the SNG product and condensate separation. SNG purification and compression are included in Unit 23. The separated condensate (~100% H_2O) flows to the BFW system. The methanation process is licensed by Lurgi Mineraloltechnik of Germany.

General Design Criteria

The unit consists of two 50 percent capacity parallel trains. Each train consists of four wet loop recycle methanation reactors in series followed by one damp methanation reactor. In consideration of critical service of the methanation recycle compressor, a spare recycle compressor, common to both trains, is provided. All rotating equipment in the unit are motor driven. The onstream factor for the unit is compatible with the overall plant stream factor of 332 days per year.

6.2.16.1 (Continued)

It is expected that the trains can be overloaded to a certain extent for a short time depending upon the performance of the recycle compressor and methanated gas compressor downstream of the unit and the maximum catalyst temperature allowable. The turndown of each of the trains is approximately 30 percent of the design capacity. Design catalyst life is one year; expected catalyst life is two years.

Process Performance Objectives

The plant design is based on a H_2/CO molar ratio of 3.65 and a CO_2 content of about 6.6 percent volume in the fresh synthesis gas feed. The methanation unit can handle a range of synthesis gas conditions deviating from the design H_2/CO ratio and/or CO_2 content, but it cannot compensate for deviations from scheduled stoichiometric number (SN) which is defined as:

$$SN = \frac{H_2 - 2C_2H_4 - C_2H_6}{3\,CO + 4\,CO_2}$$

The design value of SN is 0.7835. The hydrogen content of product SNG will increase or decrease as the SN value of the syngas increases or decreases. Stoichiometric numbers lower than 0.65 should be avoided as the catalyst may be damaged.

Feedstock

Synthesis Gas Feed

Dry Gas	43,647.3 lb-mol/hr		
Moisture	-0-		
Total Feed Rate	507,295 lb/hr		
Battery Limit Conditions:		Temperature	68°F
		Pressure	355 psia

6.2.16.1 (Continued)

Composition:

Component	Mol %
H_2	59.03
N_2/Ar (as N_2)	0.31
CO	16.18
CH_4	17.63
C_2H_6	0.22
CO_2	6.63

Products

SNG

Dry Gas	18,143.5 lb-mol/hr
Moisture	64.0 lb-mol/hr
Total Production Rate	18,207.5 lb-mol/hr
	or 343,733 lb/hr

Battery Limit Conditions: Temperature 100°F
Pressure 270 psia

Composition:

Component	Dry Mol %
H_2	1.24
N_2/Ar(as N_2)	0.75
CO	0.05
CH_4	87.53
CO_2	10.43

Process Condensate (~100% H_2O)

Total Flow Rate 163,562 lb/hr
326.8 gpm

6.2.16.1 (Continued)

Battery Limit Conditions:	Temperature	200°F
	Pressure	270 psia

Utility Requirements

Steam produced (650 psig sat'd)	817,200 lb/hr
Process Condensate Produced	163,562 lb/hr
Recycle compressor BHP	5,500 HP (4120 kW)
Electric Power	400 kW
Cooling Water (ΔT = 30°F)	293 gpm
Fuel Gas (Intermittent use for catalyst activation)	43.0 MM Btu/hr

Catalyst Requirements:	Initial	6,250 ft^3
	Annual	6,250 ft^3

6.2.16.2 PROCESS DESCRIPTION

The process flow sketch of the unit is presented on Drawing No. 835704-22-4-101, and the unit material balance and equipment list follow (Tables 6.2.16-1 and 6.2.16-2). The plot plan is shown on Drawing No. 835704-22-4-050.

Pure synthesis gas from the Rectisol unit is reacted in this unit to produce SNG. The main reactions taking place are conversion of carbon monoxide (CO) and carbon dioxide (CO_2) into methane (CH_4) by reaction with hydrogen (H_2).

$$CO + 3H_2 \longrightarrow CH_4 + H_2O$$
$$CO_2 + 4H_2 \longrightarrow CH_4 + 2H_2O$$

6.2.16.2 (Continued)

Both of these reactions are promoted by a nickel-containing catalyst and are strongly exothermic. A minor quantity of ethane present in the gas is also hydrogenated to methane as follows:

$$C_2H_6 + H_2 \longrightarrow 2CH_4$$

Heat released by the exothermic reactions is used to generate high pressure steam. The synthesis reactions are carried out in four wet methanation reactors in series followed by a damp methanator.

The fresh synthesis gas feed to the unit is preheated against the last wet methanator product and split into four streams. Feed to the first methanation reactor is flow-controlled in proportion to the recycle flow from the fourth wet methanator circulated by the recycle compressor. The mixture ratio is approximately five parts recycle gas to one part of fresh syngas. This ratio, the feed composition, and the inlet temperature to each of the wet reactors are controlled to give the same effluent temperature for all reactors. The combined fresh feed and recycle feed to the wet reactors are at 554°F.

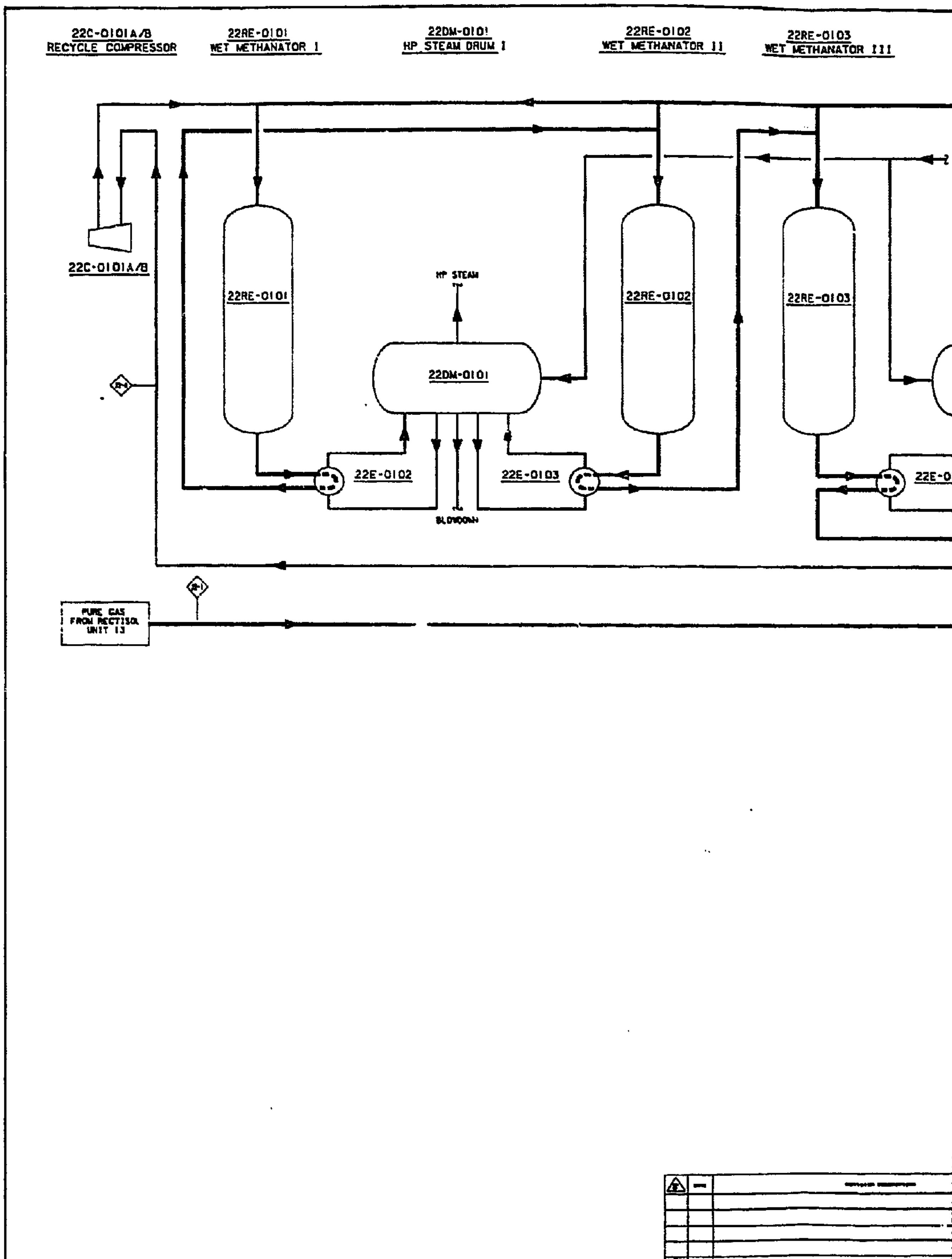
The reactors operate at pressures ranging from 350 to 320 psia in a descending order from first to fourth reactor. The effluent from the first reactor is cooled in a waste heat boiler producing high pressure steam. The cooled gas is rich in methane and water vapor, but contains very little CO, which makes it suitable for use as a diluting medium for additional fresh synthesis gas. The cooled gas is mixed with a controlled quantity of fresh syngas and fed to the second methanation reactor. The reactants in the fresh feed are methanated with the first reactor product acting as a heat sink to moderate the temperature at the outlet. The effluent is cooled again in a waste heat boiler, mixed with additional fresh syngas and fed to the third and fourth reactors/coolers where the process is repeated.

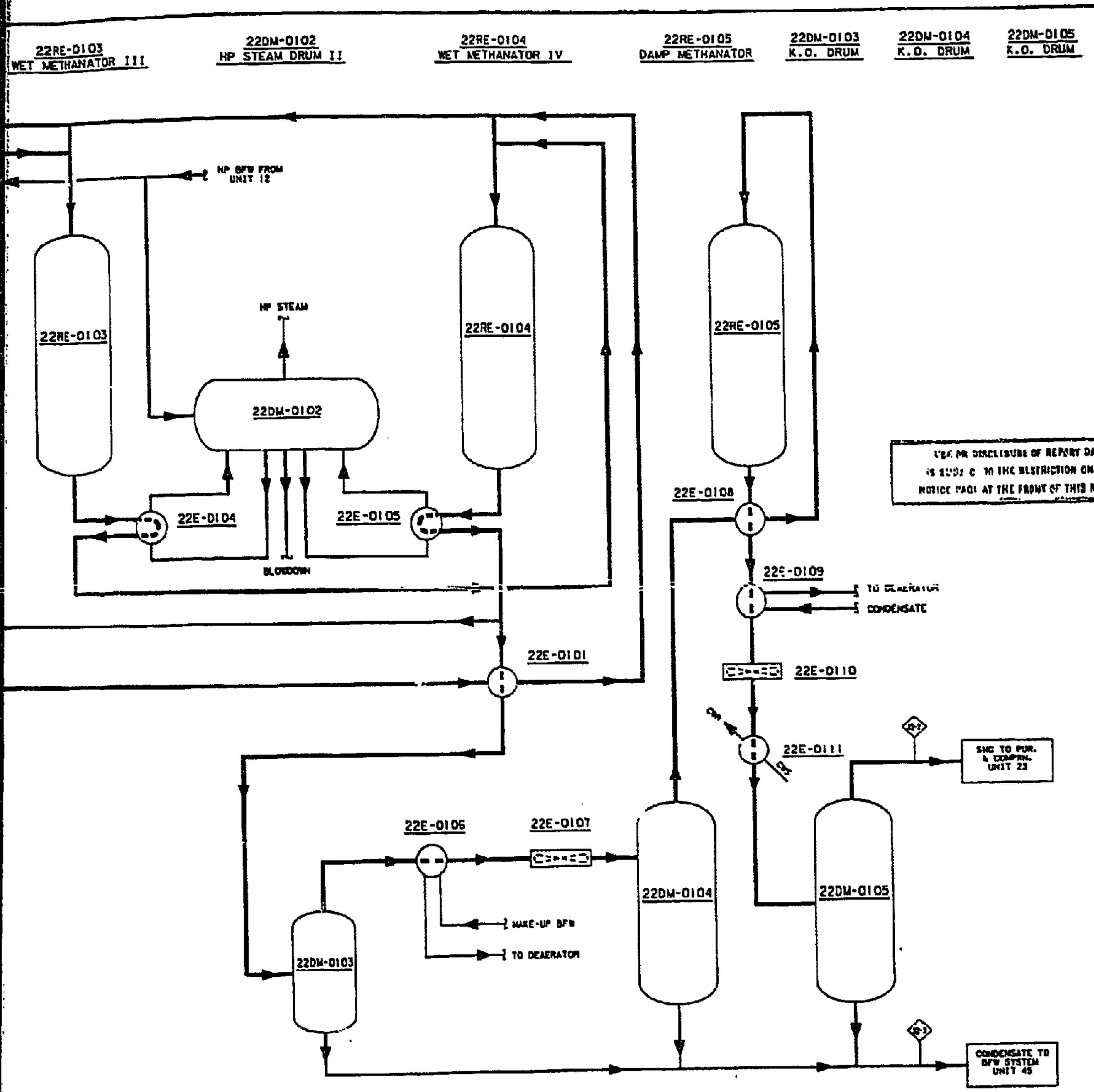
6.2.16.2 (Continued)

The effluent from the fourth waste heat boiler is divided into two streams: A recycle stream, which is recompressed and mixed with the fresh feed to the first reactor and a product stream which is processed further as follows.

The product stream from the fourth methanator is cooled against the fresh syngas feed to the unit, deaerator feedwater (BFW) and air cooled to 230°F. A portion of the product water from the methanation reaction condenses and is removed in a liquid/vapor separator drum. The gas is preheated in a feed-effluent exchanger and fed to the damp methanation reactor. The damp methanator reduces the H_2 and CO concentration to levels meeting pipeline specifications. The reactor product is cooled in a series of exchangers preheating the feed, heating cold condensate followed by air cooling and water cooling. The water from the methanation reaction is condensed and removed. The SNG product at 100°F flows to the SNG Compression and Purification unit (23). The reaction product condensate collected from the unit is sent to Unit 45 - BFW System for use as boiler feed water makeup.

A catalyst activation system is provided in the unit to reduce the nickel catalyst. Hydrogen is used as the reducing gas to activate the catalyst prior to unit startup and at subsequent periods after fresh catalyst is installed.


22C-0101A/B
RECYCLE COMPRESSOR
22RE-0101
WET METHANATOR I
22DM-0101
HP STEAM DRUM I
22RE-0102
WET METHANATOR II
22RE-0103
WET METHANATOR III
22C-0101A/B
22RE-0101
HP STEAM
22DM-0101
22RE-0102
22RE-0103
22E-0102
22E-0103
BLOWDOWN
PURE GAS FROM RECTISOL UNIT 13



USE OR DISCLOSURE OF REPORT DATA IS SUBJECT TO THE RESTRICTION ON THE NOTICE PAGE AT THE FRONT OF THIS REPORT

NOTES:

1. THIS DRAWING SHOWS ONE TRAIN OF A TWO TRAIN SECTION. CAPACITY OF EACH TRAIN IS 50% OF UNIT DESIGN. THE RECYCLE COMPRESSOR 22C-0101 IS PROVIDED WITH THREE (TWO OPERATING PLUS ONE SPARE) 50% MACHINES FOR TWO TRAINS.
2. THIS DRAWING IS A SIMPLIFIED FLOW DIAGRAM OF A PROPRIETARY UNIT. STREAM CONFIGURATIONS AND/OR EQUIPMENT ARE NOT COMPLETELY REPRESENTED.
3. THE FLOW QUANTITIES, TEMPERATURES AND PRESSURES REFERENCED BY DIAMONDS ARE SHOWN ELSEWHERE.

FLUOR

R. WHITE
C.C. ABATAY
R. O'BELMITO
R. McCARTHY
R. LANG

PROCESS FLOW DIAGRAM
METHANATION
UNIT 22

CROW TRIBE OF INDIANS — SYNFUELS FEASIBILITY STUDY

Scale	Drawing No.	Rev.
NONE	835704-22-4-101	1

003 3572210I

TABLE 6.2.16-1

MATERIAL BALANCE

METHANATION - UNIT 22

Stream Number	22-1		22-2	
Stream Name	Syngas Feed		SNG to Pur. & Comprn.	
Component	lb-mol/hr	Mol%	lb-mol/hr	Mol%
H_2	25,765.1	59.03	225.0	1.24
N_2	135.3	0.31	135.3	0.75
CO	7,062.1	16.18	9.1	0.05
CH_4	7,695.0	17.63	15,881.6	87.53
C_2H_6	96.0	0.22	-	-
CO_2	2,893.8	6.63	1,892.5	10.43
		100.00		100.00
Dry Gas, lb-mol/hr	43,647.3	100.00	18,143.5	99.65
H_2O Vapor, lb-mol/hr	-0-	-0-	64.0	0.35
Wet Gas, lb-mol/hr	43,647.3	100.00	18,207.5	100.00
Dry Gas, lb/hr	507,295		342,580	
H_2O vapor, lb/hr			1,153	
H_2O liquid, lb/hr			-	
TOTAL, lb/hr	507,295		343,733	
Pressure, psia	355		270	
Temperature, °F	68		100	

NOTE: Flow quantities, pressures, and temperatures shown are for the total unit design purposes, and are not necessarily the conditions which will be attain

TABLE 6.2.16-1

MATERIAL BALANCE

METHANATION - UNIT 22

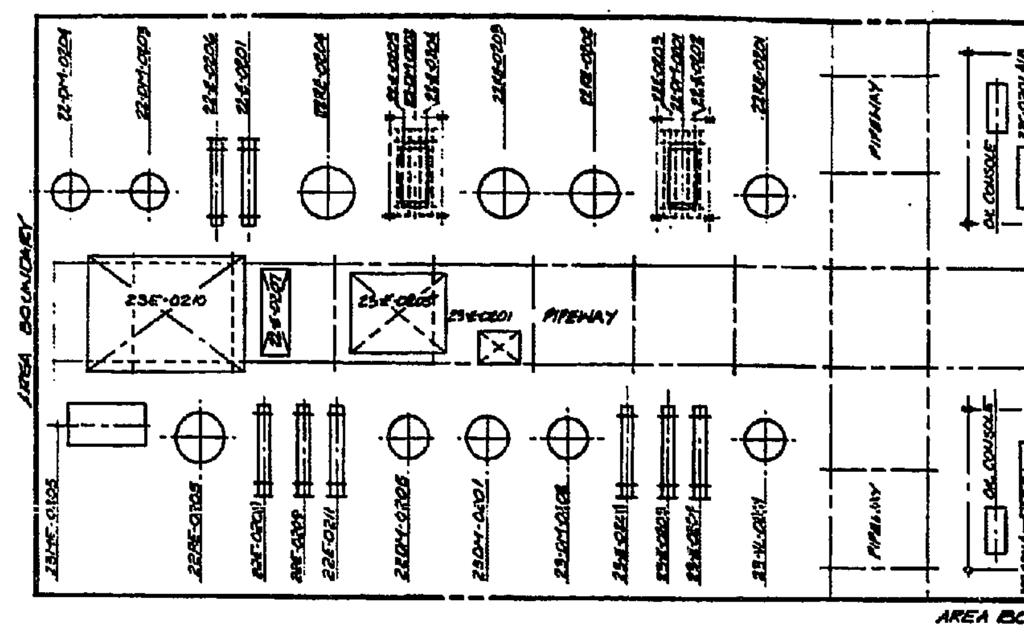
	22-2 SNG to Pur. & Comprn.		22-3 Condensate	22-4 Recycle Gas	
ol%	lb-mol/hr	Mol%		lb-mol/hr	Mol%
03	225.0	1.24		2,110.0	7.97
31	135.3	0.75		185.2	0.70
18	9.1	0.05		74.1	0.28
63	15,881.6	87.53		21,130.0	79.81
22	-	-		-	-
63	1,892.5	10.43		2,978.2	11.25
00		100.00			100.00
00	18,143.5	99.65		26,477.5	69.59
	64.0	0.35		11,570.3	30.41
00	18,207.5	100.00		38,047.8	100.00
	342,580		-	481,580	
	1,153		-	208,451	
	-		163,562	-	
	343,733		163,562	690,031	
	270		270	315	
	100		200	543	

ures shown are for the total unit on a stream-day basis, are to be used solely for process the conditions which will be attained during actual operations.

REDUCED PRINT SCALES

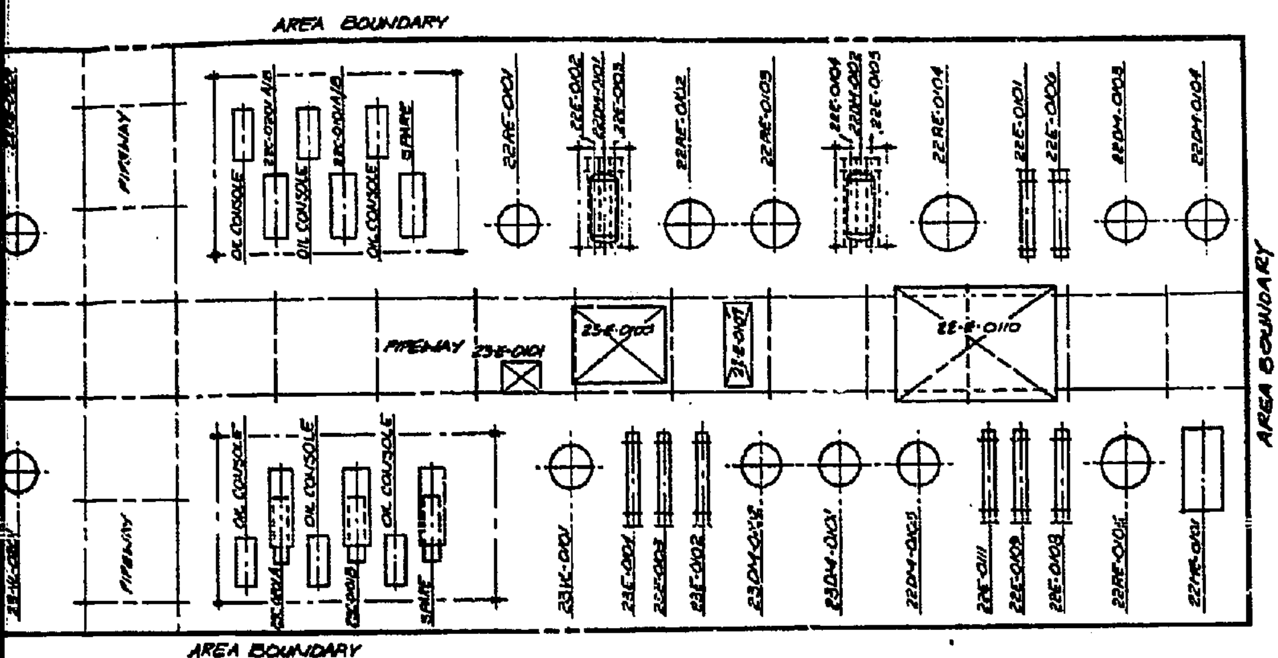
1/4"

3/8"



	DATE	REVISION DESCRIPTION
1.		APPROVED FOR STUDY

USE OR DISCLOSURE OF REPORT DATA
IS SUBJECT TO THE RESTRICTION ON THE
NOTICE PAGE AT THE FRONT OF THIS REPORT



REDUCED PRINT SCALE

ALL EQUIPMENT SIZES AND LOCATIONS ARE APPROXIMATE.

FLUOR

K.UYEKAWA

PLOT PLAN - UNITS 22 & 23
METHANATION
SNG PURIFICATION & COMPRESSION

1"=20'-0"

835704-22-4-050

1

TABLE 6.2.16-2

EQUIPMENT LIST

METHANATION - UNIT 22

Item No.	Equipment Name	Number Required Oper.	Number Required Spare
22RE-0101	Wet Methanator I	2	0
22RE-0102	Wet Methanator II	2	0
22RE-0103	Wet Methanator III	2	0
22RE-0104	Wet Methanator IV	2	0
22RE-0105	Damp Methanator	2	0
22DM-0101	HP Steam Drum I	2	0
22DM-0102	HP Steam Drum II	2	0
22DM-0103	K.O. Drum	2	0
22DM-0104	K.O. Drum	2	0
22DM-0105	K.O. Drum	2	0
22E-0101	Feed Gas Preheater	2	0
22E-0102	Waste Heat Boiler I	2	0
22E-0103	Waste Heat Boiler II	2	0
22E-0104	Waste Heat Boiler III	2	0
22E-0105	Waste Heat Boiler IV	2	0
22E-0106	Wet Methanator Product Cooler	2	0
22E-0107	Wet Methanator Product Trim Cooler	2	0
22E-0108	Damp Methanator Feed Preheater	2	0
22E-0109	Damp Methanator Product Cooler	2	0
22E-0110	Damp Methanator Product Air Cooler	2	0
22E-0111	Damp Methanator Product Trim Cooler	2	0
22C-0101 A/B	Recycle Compressor	2	1
22ME-0101	Catalyst Activation System	1	0

NOTE: Train No. 2 equipment numbers which are not shown are the same as indicated above except the train designation is 02 instead of 01.

Example:	Train No. 1	Train No. 2
	22RE-0101	22RE-0201

6.2.17 SNG PURIFICATION & COMPRESSION - UNIT 23

6.2.17.1 DESIGN BASIS

Purpose of Unit

The purpose of this unit is to remove the CO_2 in the SNG from the methanation unit to acceptable levels for pipeline transport and sales and compress the SNG product to the pressure required for pipeline transport.

Scope of Unit

The unit includes compression of SNG from methanation unit, high pressure Rectisol wash and the final compression and cooling of the product SNG to pipeline operating conditions. Flow metering of the gas and Btu analysis are carried out to provide in-plant information and to set the quantity of odorant to be injected into the product gas. Refrigeration and solvent regeneration associated with the Rectisol wash of SNG are included in the Rectisol Unit 13.

General Design Criteria

SNG compression consists of two 55 percent trains. Each train has two stages of compression, one prior to Rectisol wash and one after. The compressors are steam driven; a spare compressor common to both trains is provided.

There are two 55 percent parallel trains of SNG purification which interface with the two train operation of the Rectisol Unit 13. The CO_2 absorption is accomplished at a high pressure requiring methanation unit product gas to be compressed. This reduces hot regeneration requirements with little added compression cost (except for the additional CO_2

6.2.17.1 (Continued)

compression) since the SNG must be compressed to pipeline pressure in any case. The unit on-stream factor is compatible with the overall plant stream factor of 332 days per year.

Process Performance Objective

The Rectisol wash is designed to reduce the CO_2 in SNG to a level such that the SNG product has a HHV of 980 Btu/SCF. Under normal operating conditions, this corresponds to approximately 1.2 percent CO_2 in the SNG product. The compressors are designed to deliver 137.5 MM SCF/SD of SNG to the pipeline at 1435 psig and 115°F.

Feedstock

SNG from Methanation

Dry Gas	183,143.5 lb-mol/hr	
Moisture	64.0 lb-mol/hr	
Total Gas	18,207.5 lb-mol/hr	
	or 343,733 lb/hr	
Battery Limit Conditions:	T	100°F
	P	270 psia

Composition:

Components	Dry mol %
H_2	1.24
N_2/Ar (as N_2)	0.75
CO	0.05
CH_4	87.53
CO_2	10.43

6.2.17.1 (Continued)

Products

SNG to Pipeline

SNG Production Rate	15,100.9 lb-mol/hr 137.51 MM SCF/SD
Battery Limit Conditions:	T 115°F P 1435 psig

Composition:

Components	Mol %
H_2	1.49
N_2/Ar(as N_2)	0.90
CO	0.06
CH_4	96.34
CO_2	1.21
H_2O	-nil-

HHV = 980 Btu/SCF

Utility Requirements

The utilities associated with Rectisol wash are included in Rectisol Unit 13. The utilities for SNG compression are the following:

Operating (compressor) BHP[(1)]	18,800 HP
Cooling Water (ΔT = 30°F)	10,200 gpm (including turbine surface condenser)
Electric Power	250 kW

(1) 550 psig 750°F Steam consumed in compressor turbine driver — 145,700 lb/hr

6.2.17.1 (Continued)

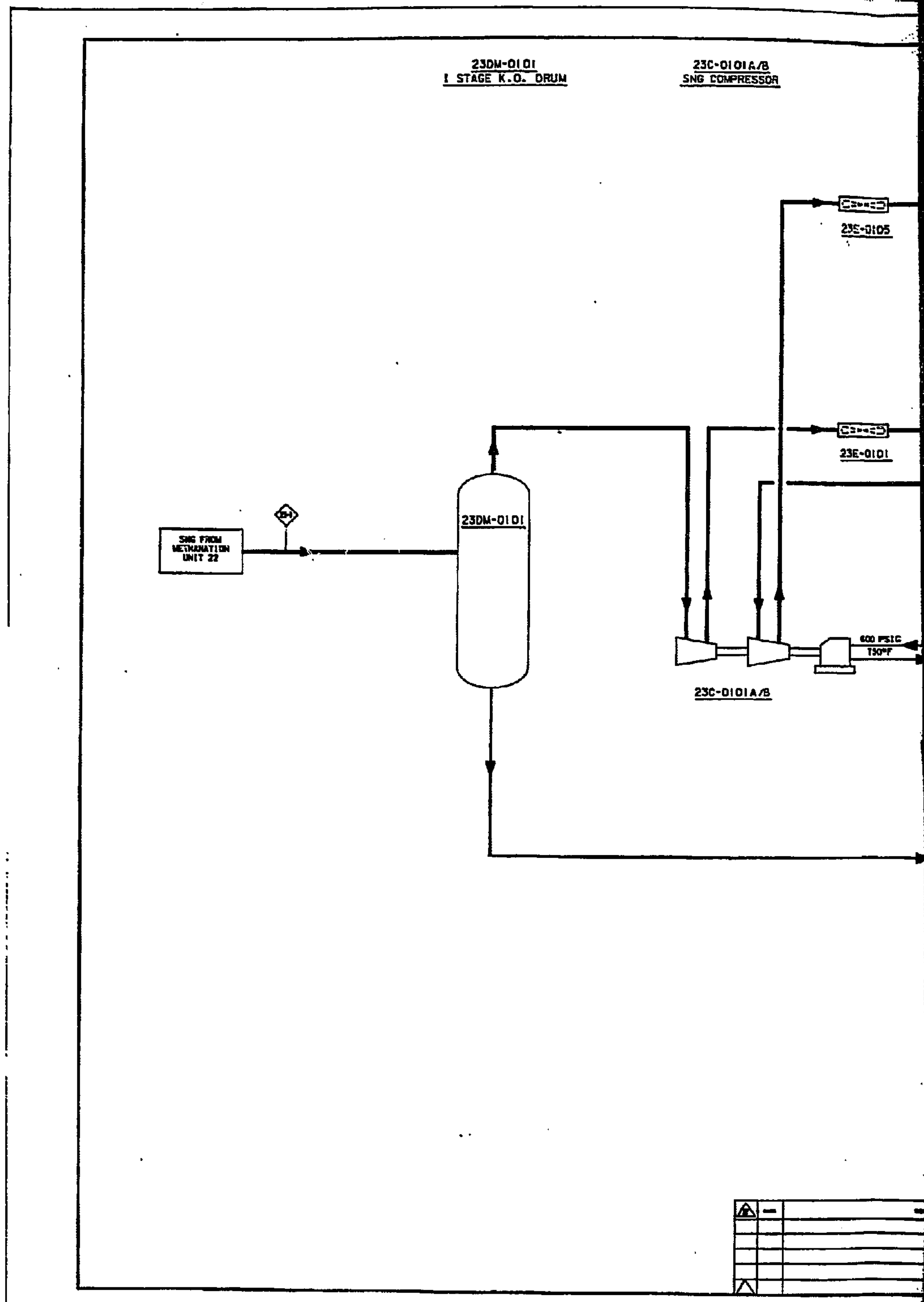
Condensate generated . 145,700 lb/hr

6.2.17.2 PROCESS DESCRIPTION

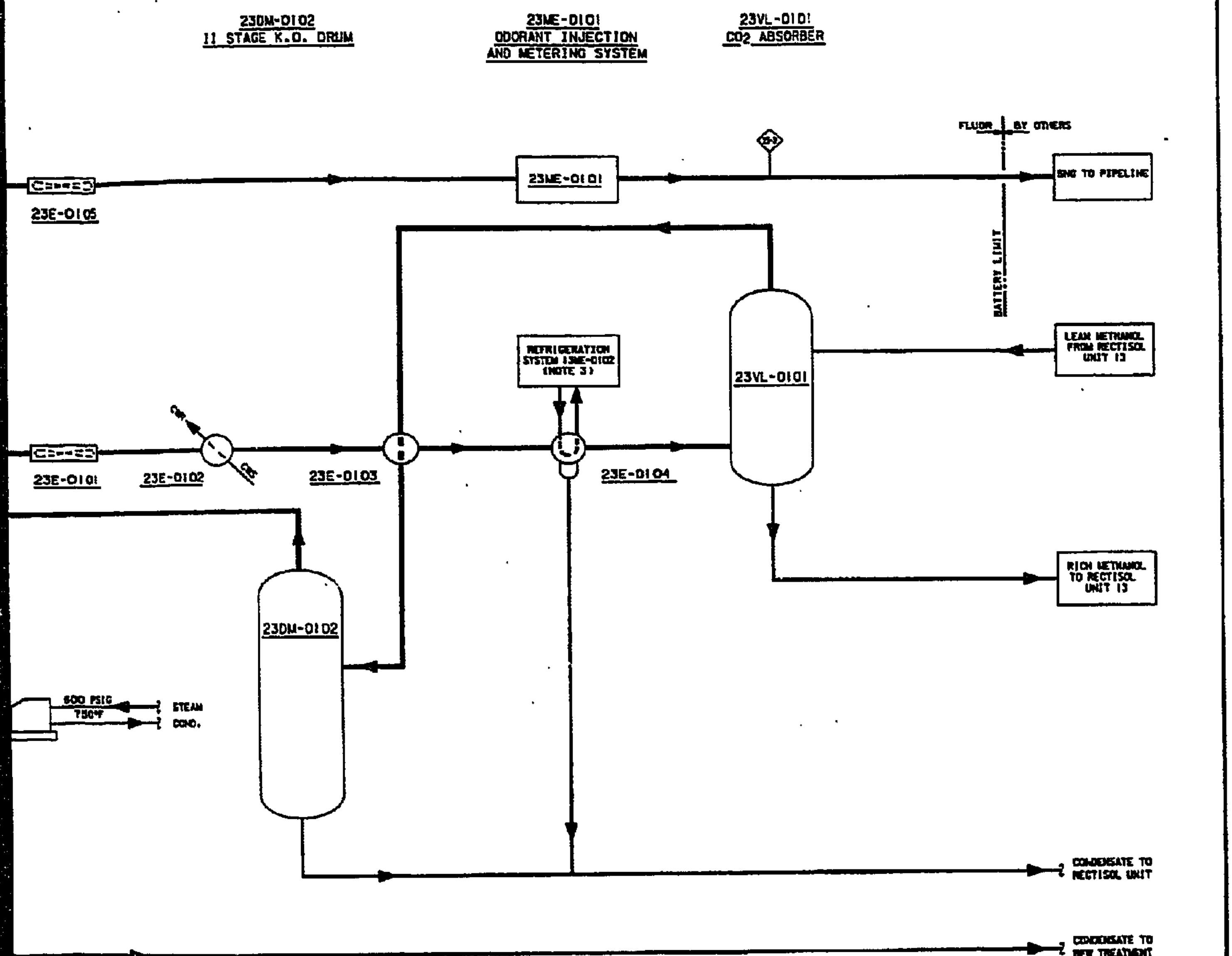
Drawing No. 835704-23-4-101 is the process flow sketch for the unit. The material balance and equipment list for the unit follow (Tables 6.2.17-1 and 6.2.17-2).

The methanated gas from Unit 22 contains CO_2 which is removed to acceptable levels and the SNG is delivered to the plant battery limit at the pipeline operating pressure.

The methanated gas fed to the unit free of any condensate carry-over is compressed to 685 psia in the first stage of the steam-driven centrifugal compressor. The compressor discharge is cooled in an air-cooler and water-cooler followed by a feed-effluent exchanger. The gas is cooled to subzero temperatures against propylene refrigerant and fed to the CO_2 absorber. The refrigeration system is part of the Rectisol Unit 13. In the absorber, CO_2 concentration in the SNG is reduced from 10.4 percent to 1.2 percent by absorption in fresh methanol solvent from the Rectisol unit. The CO_2 removal is carried out at high pressure. The rich methanol solvent which has absorbed a significant quantity of CH_4 along with the CO_2 is piped to the primary Rectisol absorber where most of the CH_4 flashes to join the fresh synthesis gas. SNG leaving the absorber is warmed by cooling the feed to absorber. The gas is compressed in the second stage of the compressor and air cooled to 115°F. The SNG product, at 1435 psig and a higher heating value of 980 Btu/SCF, is metered, odorized, and delivered to the plant battery limit. Condensate collected in the refrigerant cooler and the second stage knockout drum is piped to the Rectisol unit to recover any methanol present.



23DM-0101
I STAGE K.O. DRUM
23C-0101A/B
SNG COMPRESSOR
23E-0105
23E-0101
SNG FROM
METHANATION
UNIT 22
23DM-0101
600 PSIG
750°F
23C-0101A/B



NOTES:

1. THIS DRAWING SHOWS ONE TRAIN OF A TWO TRAIN SECTION. CAPACITY OF EACH TRAIN IS 50% OF UNIT DESIGN. SNG COMPRESSOR 23C-0101 IS PROVIDED WITH THREE (TWO OPERATING PLUS ONE SPARE) 50% MACHINES FOR TWO TRAINS.
2. THE FLOW QUANTITIES, TEMPERATURES AND PRESSURES REFERENCED BY DIAMONDS ARE SHOWN ELSEWHERE.
3. REFRIGERATION SYSTEM 13ME-0102 IS LOCATED IN UNIT 13 - RECTISOL.

USE OR DISCLOSURE OF REPORT DATA
IS SUBJECT TO THE RESTRICTION ON THE
NOTICE PAGE AT THE FRONT OF THIS REPORT

FLUOR

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PROCESS FLOW DIAGRAM
SNG PURIFICATION AND COMPRESSION
UNIT 23

CROW TRIBE OF INDIANS — SYNFUELS FEASIBILITY STUDY

NONE	835704-23-4-101	1

003 35723101

TABLE 6.2.17-1

MATERIAL BALANCE

SNG PURIFICATION & COMPRESSION - UNIT 23

Stream Number Stream Name	23-1 SNG from Methanation		23-2 SNG to Pipeline	
Component	lb-mol/hr	Mol %	lb-mol/hr	Mol %
H_2	225.0	1.24	225.0	1.49
N_2	135.3	0.75	135.3	0.90
CO	9.1	0.05	9.1	0.06
CH_4	15,881.6	87.53	14,548.8	96.34
CO_2	1,892.5	10.43	182.7	1.21
		100.00		100.00
Dry Gas, lb-mol/hr	18,143.5	99.65	15,100.9	-0-
H_2O vapor, lb-mol/hr	64.0	0.35	-0-	-0-
Wet Gas, lb-mol/hr	18,207.5	100.00	15,100.9	100.00
Dry Gas, lb/hr	342,580		245,948	
H_2O vapor, lb/hr	1,153		-	
TOTAL, lb/hr	343,733		245,948	
Pressure, psia	270		1,448	
Temperature, °F	100		115	

NOTES: 1. Flow quantities, pressures and temperatures shown are for the total unit on a stream-day basis, are to be used solely for process design purposes, and are not necessarily the conditions which will be attained during actual operations.

2. CO_2, CH_4 and H_2O removed from the SNG at a total rate of 97,090 lb/hr flow to the Rectisol Unit 13.

TABLE 6.2.17-2

EQUIPMENT LIST

SNG-PURIFICATION & COMPRESSION - UNIT 23

Item No.	Equipment Name	Number Required Oper.	Spare
23DM-0101	I Stage K.O. Drum	2	0
23DM-0102	II Stage K.O. Drum	2	0
23VL-0101	CO_2 Absorber	2	0
23E-0101	I Stage After Cooler	2	0
23E-0102	I Stage Trim Cooler	2	0
23E-0103	Crude/Pure SNG Exchanger	2	0
23E-0104	SNG Refrigeration Cooler	2	0
23E-0105	II Stage SNG After Cooler	2	0
23C-0101 A/B	SNG Compressor	2	1
23C-0101T A/B	SNG Compressor Turbine Driven	2	1
23C-0101E A/B	SNG Compressor Turbine Surface Condensor	2	1
23C-0101P A/B/C	SNG Compressor Surface Condensor Condensate Pump	2	4
23C-0101ME A/B/C	SNG Compressor Surface Condensor Ejector	2	4
23ME-0101	Odorant Injection & Metering System	2	0

NOTE: Train No. 2 equipment numbers which are not shown are the same as indicated above except the train designated is 02 instead of 01.

Example:

Train No. 1	Train No. 2
23DM-0101	23DM-0201

6.2.18 PARTIAL OXIDATION - UNIT 24

6.2.18.1 DESIGN BASIS

Purpose of Unit

The purpose of the Texaco Gasification Process (POX) unit is to produce additional synthesis gas by partial oxidation of the Lurgi liquids (tar, oil, phenols).

Scope of Unit

The unit is designed to gasify 282 ST/D of the liquid feed stock containing tar, oil, and phenols. The unit includes three major sections as follows:

(1) Texaco Gasification Process and Waste Heat Recovery
(2) Carbon Scrubbing
(3) Carbon Recovery

General Design Criteria

The unit is designed as a single train. The unit produces approximately 37.5 MM SCF/D of wet crude synthesis gas. The unit onstream factor is compatible with the overall plant stream factor of 332 days per year. All pumps in the unit are motor driven and spared.

6.2.18.1 (Continued)

Feedstock

Oil Feed

The composition, feed rate, and the battery limit conditions of the feed streams to the unit are as follows:

Component	Wt.%
Oil	64.54
Tar	21.52
Phenol	13.94
Total Flow Rate	23,487 lb/hr
Temperature	350 °F
Pressure	100 psig

Crude Gas

Composition	Mole %
H_2	41.34
N_2/Ar (as N_2)	0.54
CO	51.03
CO_2	6.53
CH_4	0.33
H_2S	0.20
COS	0.01
NH_3/HCN	0.02
Flow Rate	75051 lb/hr
Flow Rate	37.52 MM SCF/D
Temperature	355 °F
Pressure	450 psig

6.2.18.1 (Continued)

Ash

Flow Rate, lbs/hr	18

Utility Requirements

Consumption:

600 psig sat. steam	11,492 lb/hr
Oxygen @ 485 psig, 284°F	21,725 lb/hr
Makeup condensate @ 447 psig, 212°F	21,346 lb/hr
HP BFW @ 775 psig, 230°F	18,900 lb/hr
LP BFW @ 140 psig, 230°F	39,562 lb/hr
Power	171 kW
Cooling water, (ΔT = 30°F)	465 gpm

Production

100 psig sat. steam	38,375 lb/hr
650 psig sat. steam	18,333 lb/hr
Blowdown water to Biotreating	2,981 lb/hr

6.2.18.2 PROCESS DESCRIPTION

The process flow sketch for the Partial Oxidation Unit is presented on Drawing No. 835704-24-4-101. The material balance and equipment list for the unit follow (Tables 6.2.18-1 and 6.2.18-2). The plot plan is shown on Drawing No. 835504-24-4-050.

6.2.18.2 (Continued)

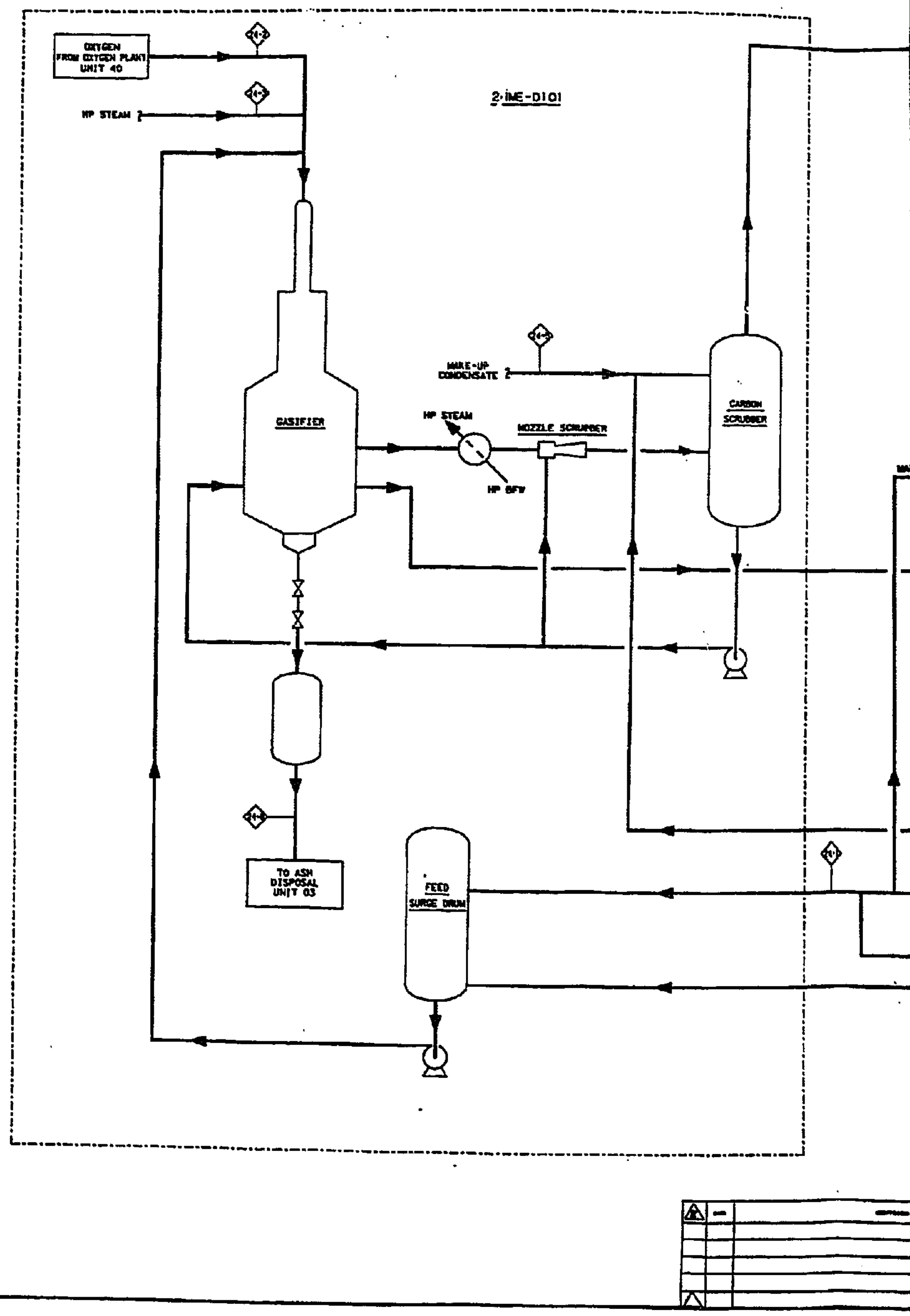
Phenols from the Phenosolvan unit and tar plus oil from the Tar Distillation unit are fed to the tar-oil slurry tank. The mixture of hydrocarbon liquids with oxygen and high pressure steam is fed to the Texaco gasifier through a special burner. The hydrocarbons are partially oxidized in the gasifier generating additional gas for synthesis of methane.

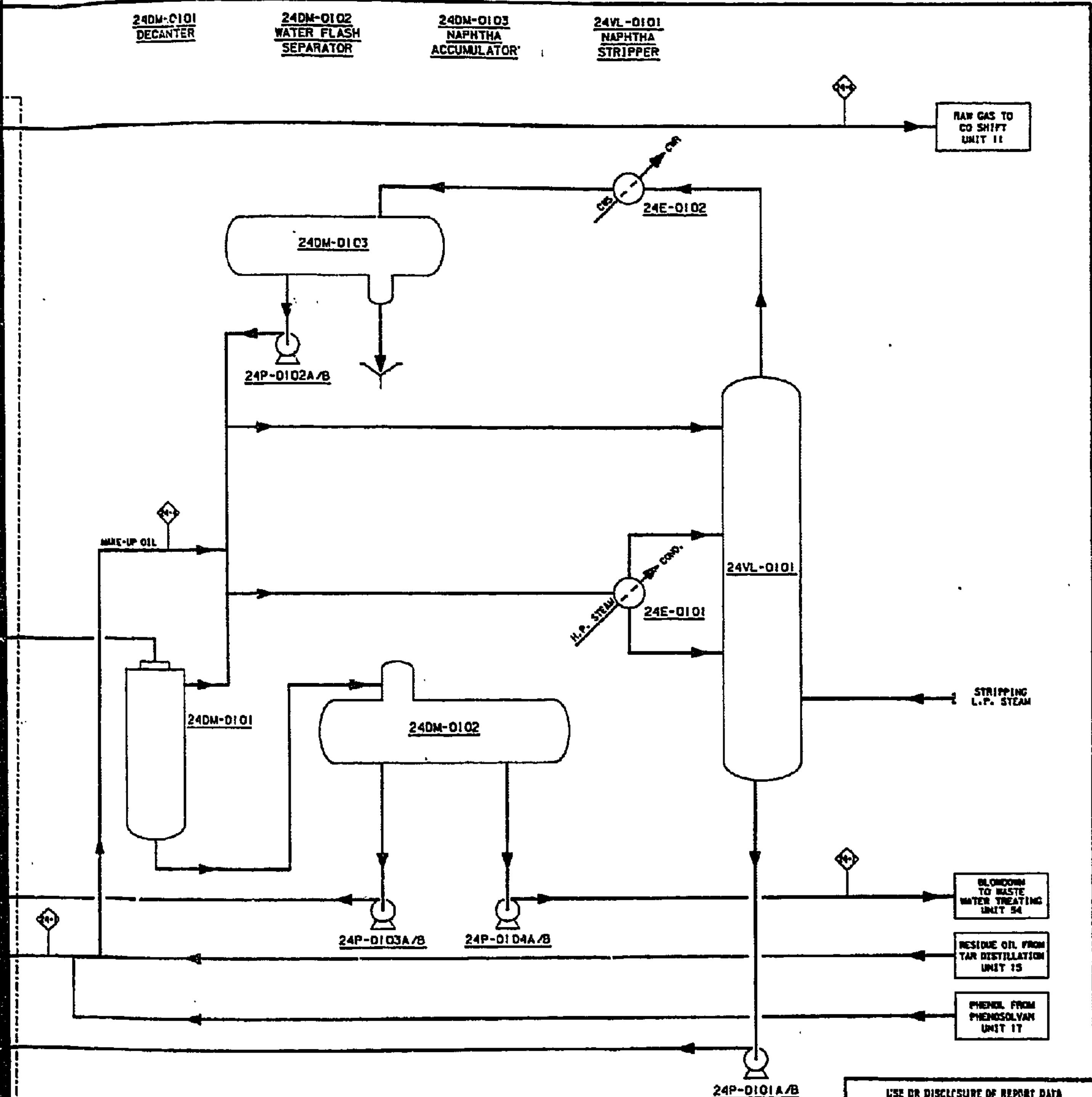
The hot raw gas from the gasifier is passed through a radiant waste heat boiler and a convective waste heat boiler producing, respectively, high pressure and medium pressure steam. The waste heat recovery section is followed by soot scrubbing which quenches the gas and removes the particulates from the raw gas in a scrubber. The particulate-free raw gas is mixed with the gas from Lurgi gasifiers before flowing to the CO Shift unit.

The major portion of the scrubber bottoms is circulated to the radiation sump section of the gasifier. The remaining portion is pumped back to the scrubber. The black water from the radiation sump is routed to a decanter where lighter hydrocarbons separate from heavier liquids as an overflow stream to the naphtha stripper. The naphtha stripper recovers hydrocarbons as the bottom product for recycle to the gasifier via the oil feed surge drum. The overhead vapors from the naphtha stripper are water cooled, and condensed. The condensate then flows to the naphtha accumulator. A portion of the condensate is returned to the column as reflux. The remaining condensate and the makeup oil is joined by the overflow from the decanter to serve as feed to the naphtha stripper.

The water and carbon particles contained in the bottoms of the decanter are separated in the water flash separator. The clear water is recycled to the scrubber and used to scrub the raw gas. Some water is purged to prevent buildup of dissolved solids in the system. Purging is accomplished by the blowdown pumps.

24ME-0101
GASIFICATION, WASTE HEAT RECOVERY AND CARBON RECOVERY SYSTEM





NOTES:

1. THIS DRAWING REPRESENTS A SINGLE TRAIN UNIT.
2. THIS DRAWING IS A SIMPLIFIED FLOW DIAGRAM OF A PROPRIETARY UNIT. STREAM CONFIGURATIONS AND/OR EQUIPMENT ARE NOT COMPLETELY REPRESENTED.
3. THE FLOW QUANTITIES, TEMPERATURES AND PRESSURES REFERENCED BY DIAMONDS ARE SHOWN ELSEWHERE.

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FLUOR

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PROCESS FLOW DIAGRAM
PARTIAL OXIDATION
UNIT 24

CROW TRIBE OF INDIANS — SYNFUELS FEASIBILITY STUDY

NONE | 835704-24-4-101 | 1

003 35724101

TABLE 6.2.18-1

MATERIAL BALANCE

PARTIAL OXIDATION - UNIT 24

Stream Number	24-1		24-2		24-3		24-4		24-5	
Stream Name	Feed		Oxidant		Steam		Makeup Oil		Makeup Condensate	
Component	lb-mol/hr	Mol%	lb-mol/hr	Mol%	lb-mol/hr	Mol%	lb-mol/hr	Mol%	lb-mol/hr	Mol%
H_2										
N_2			10.2	1.5						
CO										
CO_2										
H_2S										
COS										
NH_3/HCN										
O_2			670.0	98.5						
Total Dry Gas			680.2	100.0						
H_2O										
Total Wet Gas										
Dry Gas, lb/hr			21,725							
H_2O, lb/hr					11,492				21,346	
Tars, lb/hr	5,008						60			
Oils, lb/hr	15,023						118			
Naphtha, lb/hr							33			
Phenols, lb/hr	3,245									
Ash, lb/hr	(18 with tar)									
Total, lb/hr	23,276		21,725		11,492		211		21,346	
Pressure, psia	115		485		613		115		460	
Temperature, °F	350		284		489		350		212	

NOTE: Flow quantities, pressures and temperatures shown are for the total unit on a stream- for process design purposes, and are not necessarily the conditions which will be att

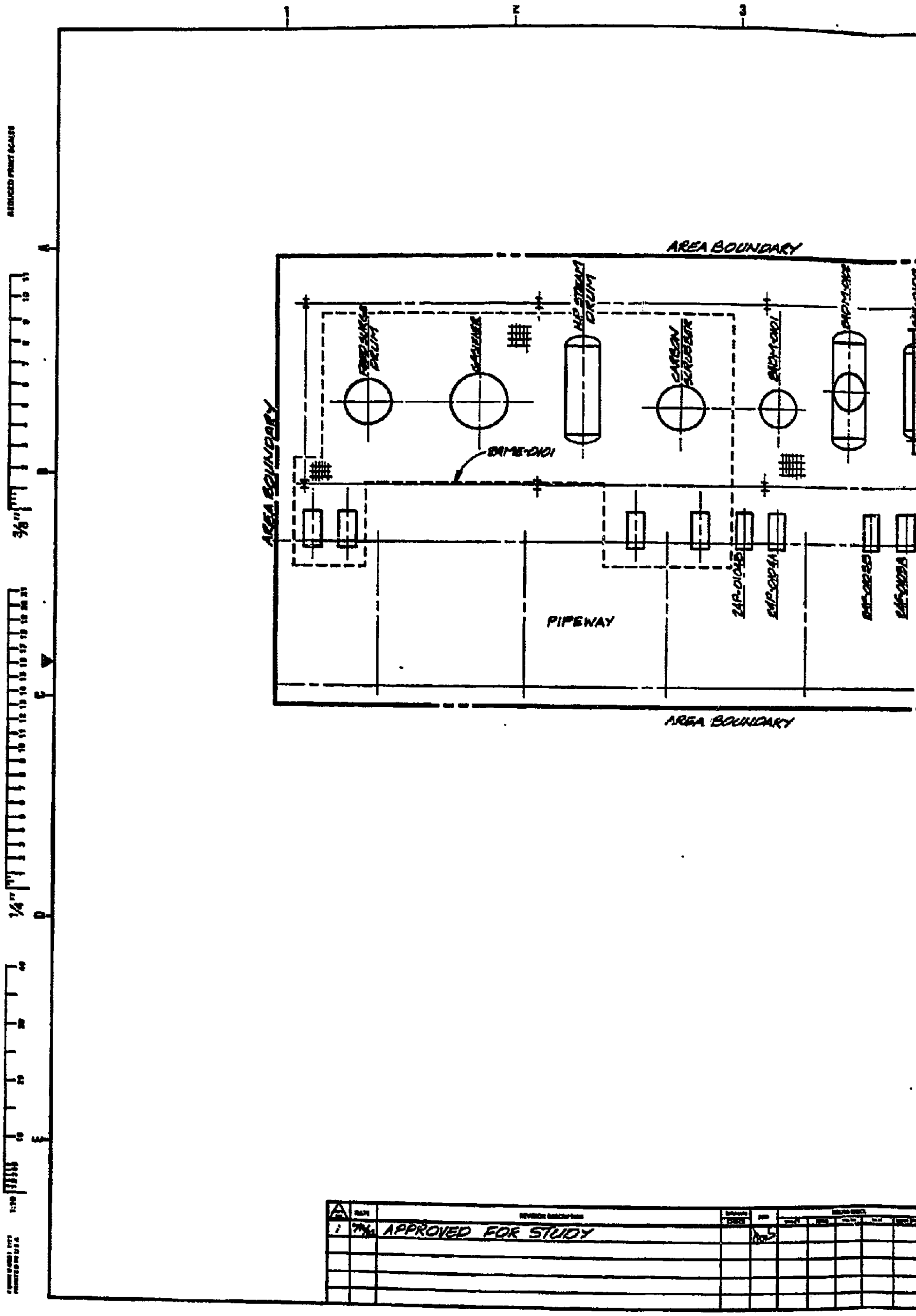
TABLE 6.2.18-1

MATERIAL BALANCE

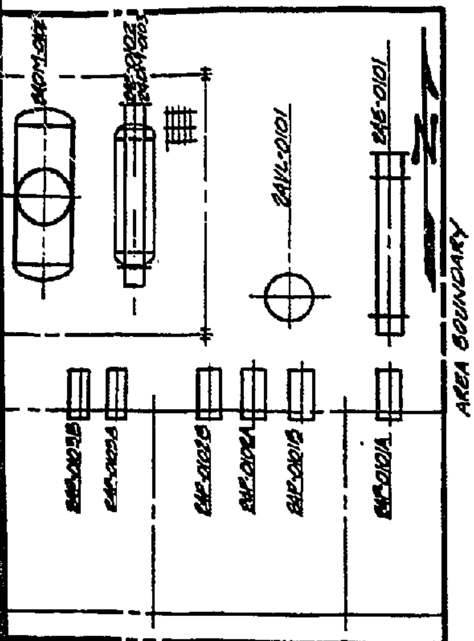
PARTIAL OXIDATION - UNIT 24

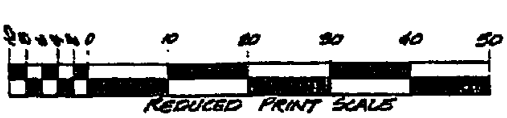
24-3		24-4		24-5		24-6		24-7		24-8	
Steam		Makeup Oil		Makeup Condensate		Raw Gas to Unit 11		Blowdown to Unit 54		Ash	
lb-mol/hr	Mol%	lb-mol/hr	Mol%	lb-mol/hr	Mol%	lb-mol/hr	Mol%	lb-mol/hr	Mol%	lb-mol/hr	Mol%
						1,161.1	41.34				
						15.1	0.54				
						1,433.0	51.03				
						183.4	6.53				
						5.5	0.20				
						0.3	0.01				
						0.6	0.02				
						2,808.4	100.0				
						1,316.2					
						1,124.6					
						51,338					
11,492				21,346		23,713		2,981			
		60									
		118									
		33									
										18	
11,492		211		21,346		75,051		2,981		18	
613		115		460		450		70		Ambient	
489		350		212		355		269		235	

tures shown are for the total unit on a stream-day basis, are to be used solely
not necessarily the conditions which will be attained during actual operations.


AREA BOUNDARY
HP STEAM DRUM
CARBON SCRUBBER
PIPEWAY
AREA BOUNDARY
APPROVED FOR STUDY

USE OR DISCLOSURE OF REPORT DATA IS SUBJECT TO THE RESTRICTION ON THE NOTICE PAGE AT THE FRONT OF THIS REPORT





ALL EQUIPMENT SIZES AND LOCATIONS ARE APPROXIMATE

DWG. NO.	REFERENCE DRAWING
00-6-050	SITE #1 PLOT PLAN

FLUOR

K. UYEKAWA
HUTTEN
T. CONTI
J. SMETS
R. LANG

PLOT PLAN - UNIT - 24
PARTIAL OXIDATION
CROW TRIBE OF INDIANS — MONTANA

Scale: 1"=10'-0"

Drawing No.: 835704-24-4-050 — Rev. 1

TABLE 6.2.18-2

EQUIPMENT LIST

PARTIAL OXIDATION - UNIT 24

Item No.	Equipment Name	Number Required Oper.	Number Required Spare
24DM-0101	Decanter	1	0
24DM-0102	Water Flash Separator	1	0
24DM-0103	Naphtha Accumulator	1	0
24VL-0101	Naphtha Stripper	1	0
24E-0101	Naphtha Stripper Feed Preheater	1	0
24E-0102	Naphtha Condenser	1	0
24P-0101 A/B	Recycle Pump	1	1
24P-0102 A/B	Naphtha Pump	1	1
24P-0103 A/B	Water Recycle Pump	1	1
24P-0104 A/B	Blowdown Pump	1	1
24ME-0101	Gasification, Waste Heat Recovery and Carbon Recovery System	1	0

6.2.19 PSA HYDROGEN PRODUCTION - UNIT 25

6.2.19.1 DESIGN BASIS

Purpose of Unit

The purpose of the pressure swing adsorption (PSA) unit is to produce a 99.5 percent volume H_2 stream for the Naphtha Hydrotreating unit and the Sulfur Recovery SCOT unit from pure syngas feed.

Scope of Unit

The unit includes a PSA system from the vendor (Union Carbide), H_2 compression and offgas compression to the required pressures.

General Design Criteria

The PSA unit is designed in one 100 percent train except for the H_2 and fuel gas compressors which are provided with 100 percent spare. The train shall include provision to maintain 50 percent of the plant in operation during malfunction of individual pieces of equipment. The onstream factor for the unit is compatible with the overall plant stream factor of 332 days per year. The compressors in the unit are motor driven. The design run length between scheduled maintenance is expected to be two years.

Process Performance Objectives

The PSA unit is designed to recover 72 percent of the H_2 in feed gas and produce hydrogen with a minimum purity of 99.5 vol. percent and a CO + CO_2 concentration of less than 10 ppmv.

6.2.19.1 (Continued)

Feedstock

Pure Syngas Feed

Feed Rate 277.6 lb mol/hr
or 3,226 lb/hr

Battery Limit Conditions: Temperature 68°F
Pressure 355 psia

Composition:

Component	Mol %
H_2	59.03
N_2/Ar(as N_2)	0.31
CO	16.18
CH_4	17.63
C_2H_6	0.22
CO_2	6.63

Products

Product H_2

Production Rate 118.6 lb-mol/hr
or 255 lb/hr

Composition:

99.5% H_2
0.5% N_2

6.2.19.1 (Continued)

Battery Limit Conditions:	Temperature	380°F
	Pressure	900 psia

Purge Gas (To Fuel Gas)

Flow Rate	159.0 lb-mol/hr	
Battery Limit Conditions:	Temperature	100°F
	Pressure	75 psia

Composition:

Component	Mol %
H_2	28.9
N_2/Ar(as N_2)	0.1
CO	28.2
CH_4	30.8
C_2H_6	0.4
CO_2	11.6
	100.0

Utility Requirements

Electric Power	600 kW
Cooling Water (ΔT = 30°F)	24.6 gpm
Fuel Gas Produced	30.46 MM Btu/hr

6.2.19.2 PROCESS DESCRIPTION

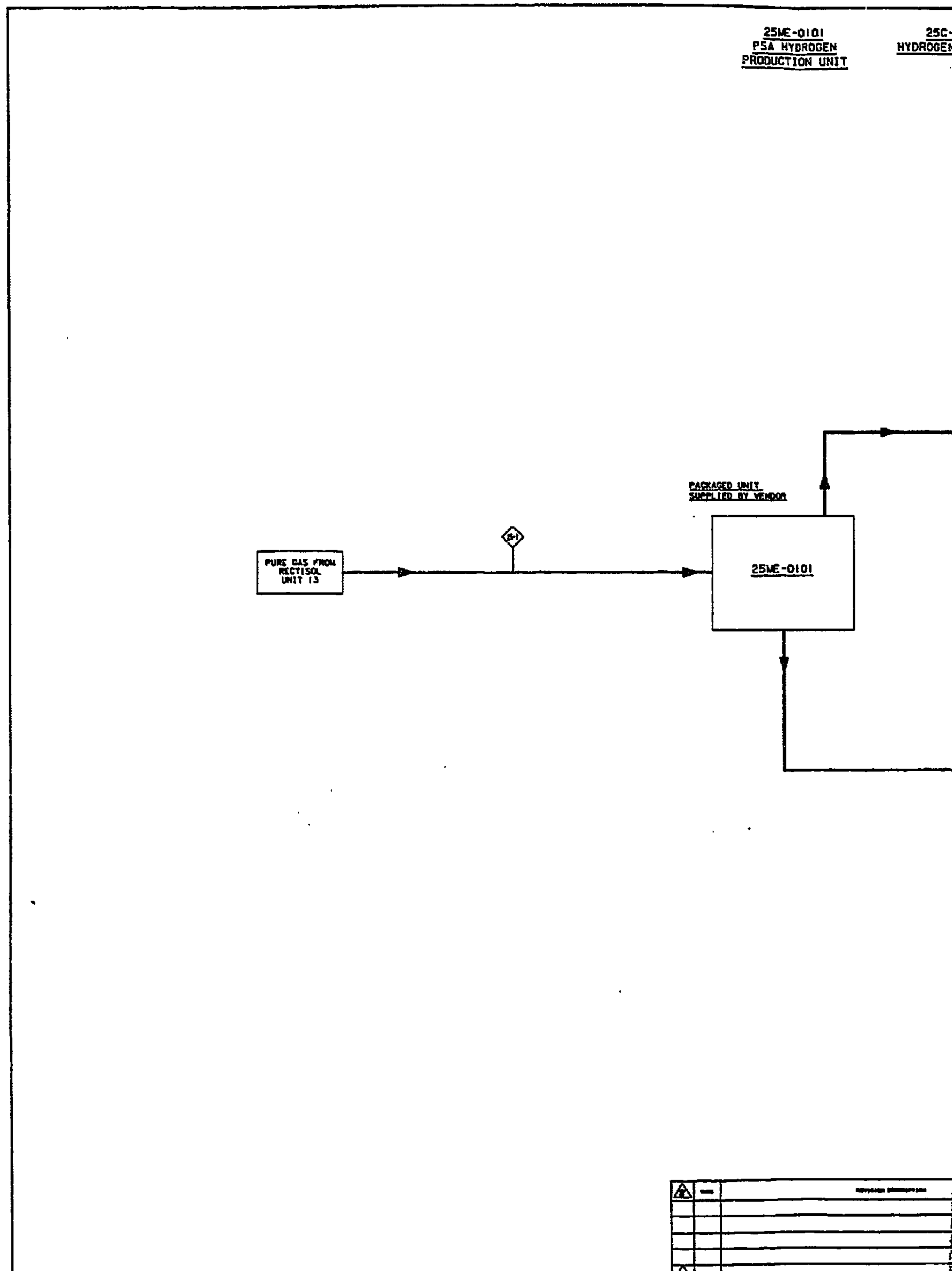
The process flow sketch of the unit is shown on Drawing No. 835704-25-4-101 and the unit material balance and equipment list follow (Tables 6.2.19-1 and 6.2.19-2). The plot plan is shown on Drawing No. 835704-25-4-050.

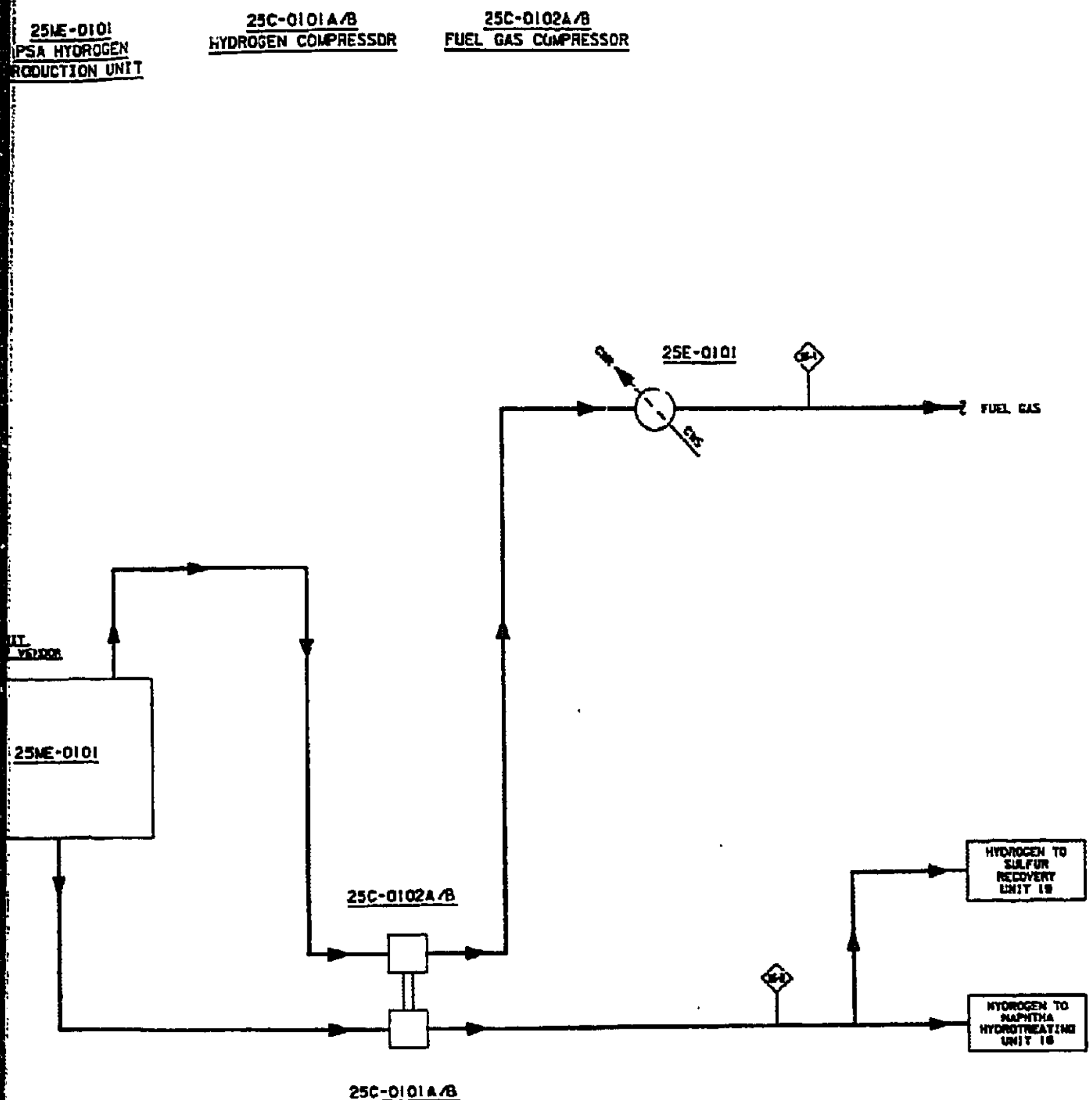
A slipstream of pure synthesis gas from the Rectisol Unit is fed to this unit to produce pure hydrogen (99.5% H_2) for use in Naphtha

6.2.19.2 (Continued)

Hydrotreating unit and the SCOT section of the Sulfur Recovery unit. Pressure swing adsorption (PSA) process is used to remove the nonhydrogen components in the feed gas by adsorption on molecular sieves. The hydrogen product is compressed in a reciprocal compressor and delivered to the Naphtha Hydrotreating unit. A small quantity of hydrogen flows to the Sulfur Recovery unit for use as reducing gas.

Purge gas from the PSA unit is produced at near-ambient conditions, and has significant fuel value. The gas is compressed, cooled and sent to the plant fuel gas system.


25ME-0101
PSA HYDROGEN
PRODUCTION UNIT
HYDROGEN
PACKAGED UNIT
SUPPLIED BY VENDOR
PURE GAS FROM
RECTISOL
UNIT 13
25ME-0101



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IS SUBJECT TO THE RESTRICTION ON THE
NOTICE PAGE AT THE FRONT OF THIS REPORT

NOTES:

1. THIS DRAWING REPRESENTS A SINGLE TRAIN UNIT.
2. THIS DRAWING IS A SIMPLIFIED FLOW DIAGRAM OF A PROPRIETARY UNIT. STREAM CONFIGURATIONS AND/OR EQUIPMENT ARE NOT COMPLETELY REPRESENTED.
3. THE FLOW QUANTITIES, TEMPERATURES AND PRESSURES REFERENCED BY DIAMONDS ARE SHOWN ELSEWHERE.

USE OR DISCLOSURE OF REPORT DATA
IS SUBJECT TO THE RESTRICTION ON THE
NOTICE PAGE AT THE FRONT OF THIS REPORT

FLUOR

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PROCESS FLOW DIAGRAM
PSA HYDROGEN PRODUCTION
UNIT 25

CROW TRIBE OF INDIANS
SYNFUELS FEASIBILITY STUDY

NONE
835704-25-4-101
1
003 357251 01

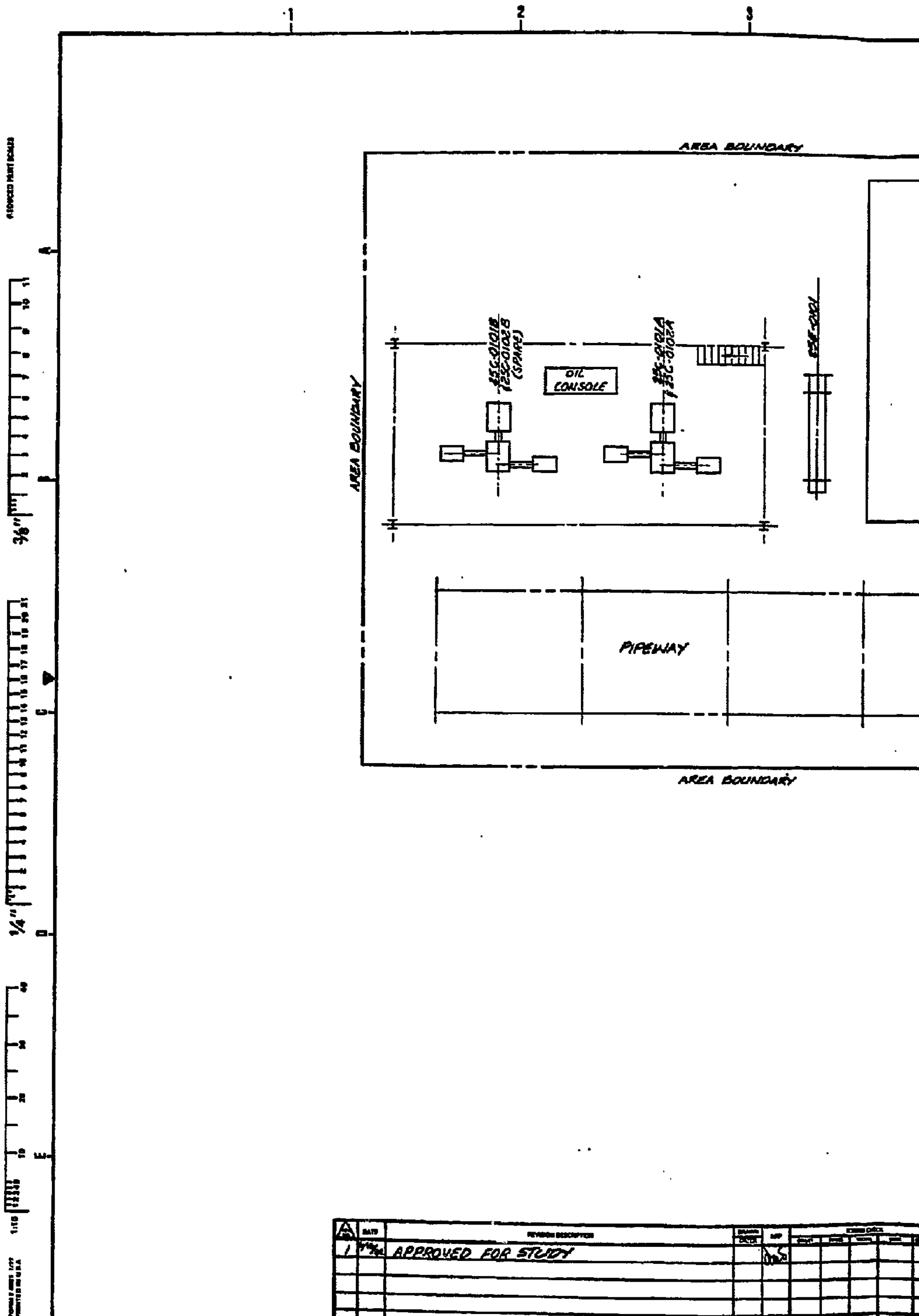
TABLE 6.2.19-1

MATERIAL BALANCE

PSA HYDROGEN PRODUCTION - UNIT 25

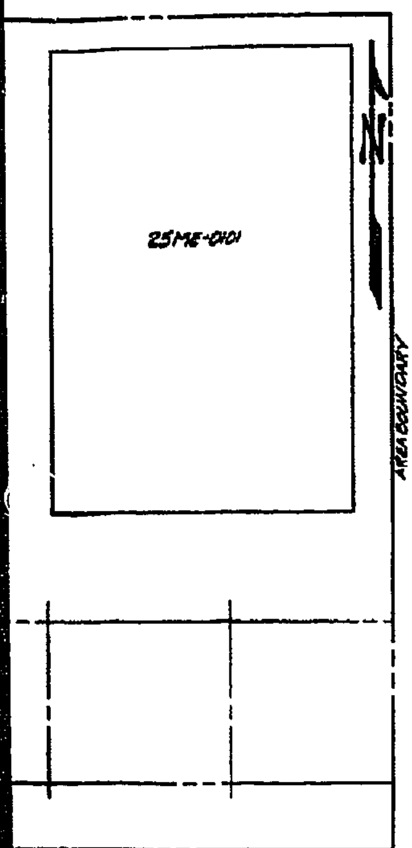
Stream Number	25-1		25-2		25-3	
Stream Name	Syngas Feed		Hydrogen Product		Purge Gas To Fuel Gas	
Component	lb-mol/hr	Mol %	lb-mol/hr	Mol %	lb-mol/hr	Mol %
H_2	164.0	59.03	118.0	99.5	46.0	28.93
N_2	0.8	0.31	0.6	0.5	0.2	0.13
CO	44.9	16.18			44.9	28.24
CH_4	48.9	17.63			48.9	30.75
C_2H_6	0.6	0.22			0.6	0.38
CO_2	18.4	6.63			18.4	11.57
TOTAL, lb-mol/hr	277.6	100.00	118.6	100.00	159.0	100.00
TOTAL, lb/hr	3,226		255		2,971	
Pressure, psia	355		900		75	
Temperature, °F	68		380		100	

NOTE: Flow quantities, pressures and temperatures shown are for the total unit on a stream-day basis, are to be used solely for process design purposes, and are not necessarily the conditions which will be attained during actual operations.



AREA BOUNDARY
AREA BOUNDARY
OIL CONSOLE
PIPEWAY
AREA BOUNDARY
1
2
3
1/4"
3/8"
APPROVED FOR STUDY

USE OR DISCLOSURE OF REPORT DATA IS SUBJECT TO THE RESTRICTION ON THE NOTICE PAGE AT THE FRONT OF THIS REPORT



REDUCED PRINT SCALE

ALL EQUIPMENT SIZES AND LOCATIONS ARE APPROXIMATE

DWG NO	REFERENCE DRAWING
00-5-050	31TG31 PLOT PLAN

FLUOR

KUYEKAWA
HUTTEN
T. CONTI
J. SMETS
R. LANG

PLOT PLAN - UNIT 25
PSA HYDROGEN PRODUCTION
CROW TRIBE OF INDIANS — MONTANA

Scale: 1"=10'-0"

Drawing number: 835704-25-4-050

Revision: 1

TABLE 6.2.19-2

EQUIPMENT LIST
PSA HYDROGEN PRODUCTION - UNIT 25

Item No.	Equipment Name	Number Required Oper.	Number Required Spare
25C-0101 A/B	Hydrogen Compressor	1	1
25C-0102 A/B	Fuel Gas Compressor with Intercooler	1	1
25E-0101	Fuel Gas After Cooler	1	0
25ME-0101	PSA Hydrogen Production (Package) Unit	1	0