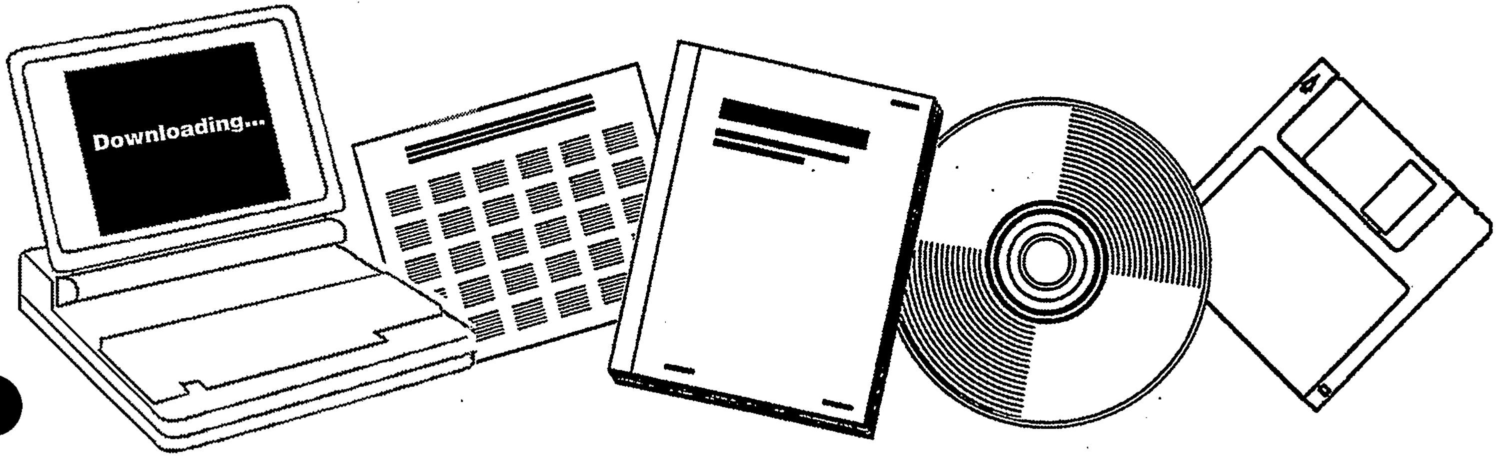
HYGAS: 1964 TO 1972. PIPELINE GAS FROM COAL--HYDROGENATION (IGT HYDROGASIFICATION PROCESS). VOLUME 4

INSTITUTE OF GAS TECHNOLOGY, CHICAGO, ILL

JUL 1975



U.S. Department of Commerce
National Technical Information Service

One Source. One Search. One Solution.

NTIS

Providing Permanent, Easy Access to U.S. Government Information

National Technical Information Service is the nation's largest repository and disseminator of government-initiated scientific, technical, engineering, and related business information. The NTIS collection includes almost 3,000,000 information products in a variety of formats: electronic download, online access, CD-ROM, magnetic tape, diskette, multimedia, microfiche and paper.

Search the NTIS Database from 1990 forward

NTIS has upgraded its bibliographic database system and has made all entries since 1990 searchable on **www.ntis.gov.** You now have access to information on more than 600,000 government research information products from this web site.

Link to Full Text Documents at Government Web Sites

Because many Government agencies have their most recent reports available on their own web site, we have added links directly to these reports. When available, you will see a link on the right side of the bibliographic screen.

Download Publications (1997 - Present)

NTIS can now provides the full text of reports as downloadable PDF files. This means that when an agency stops maintaining a report on the web, NTIS will offer a downloadable version. There is a nominal fee for each download for most publications.

For more information visit our website:

www.ntis.gov

U.S. DEPARTMENT OF COMMERCE
Technology Administration
National Technical Information Service
Springfield, VA 22161

FE381T9P4

HYGAS(T.M): 1964 TO 1972
PIPELINE GAS FROM COAL — HYDROGENATION
(IGT HYDROGASIFICATION PROCESS)

R&D REPORT NO. 22 FINAL REPORT

VOLUME 4

PART VIII : COMMERCIAL PLANT DESIGN

PART IX : PROCESS ECONOMICS

OCR Contract No.

14-01-0001-381 July 1964
14-01-0001-381 (1) June 1967
14-01-0001-381 (2) March 1972

Performance Period: July 1964 to September 1972

Prepared for

ENERGY RESEARCH AND DEVELOPMENT ADMINISTRATION

Washington, D.C. 20545

NOTICE

Available from the National Technical Information Service, U. S. Department of Commerce, Springfield, Virginia 22161

Printed in the United States of America
USERDA Technical Information Center, Oak Ridge, Tennessee

GENERAL DISCLAIMER

TE-381-T9-P4
Distribution Category UC-90c

Institute of Gas Technology
IIT Center
Chicago, Illinois 60616

HYGAS™: 1964 TO 1972
PIPELINE GAS FROM COAL — HYDROGENATION
(IGT HYDROGASIFICATION PROCESS)

Final Report

by

IGT Process Research Division

Prepared for

ENERGY RESEARCH AND DEVELOPMENT ADMINISTRATION

Washington, D.C. 20545

July 1975

PART VIII

COMMERCIAL PLANT DESIGN

TABLE OF CONTENTS, PART VIII

TABLE OF CONTENTS, PART VIII
(Continued)

LIST OF FIGURES, PART VIII

LIST OF FIGURES, PART VIII
(Continued)

LIST OF FIGURES, PART VIII
(Continued)

LIST OF TABLES, PART VIII

8.0. Summary of Commercial Plant Design

The design of a HYGAS demonstration plant is discussed in Part VIII: the design is based on technology existing in December 1969. The design study was undertaken by Procon Incorporated and IGT for the Department of the Interior, Office of Coal Research, and the American Gas Association. The design that evolved could be the basis for one train of the full-scale commercial facility that is illustrated in Figure 8-1.

The 1969 plant design incorporated the electrothermal process for manufacture of hydrogen-rich gas because that process was most advanced at that time. Extensive research and development by IGT for OCR and others has since proved that two alternative systems for producing hydrogen-rich gas — namely the steam-oxygen and steam-iron processes — may be even more promising based on current economics. Facilities for extensive testing of these processes at the HYGAS pilot plant are being readied as this report is written.

Because the design discussed was an early design and progress in the hydrogasification of coal to high-Btu gas for pipeline use has been substantial since 1969, a critique of the design also is presented together with suggested areas of improvement.* Progress has been achieved in most of the improvement areas discussed since this work was performed. The purpose of the early demonstration plant design in 1969 was 1) to provide a guide for future designs, both in the arrangement and sequence of processing steps, 2) to pinpoint problem areas where pilot-scale study should be concentrated, and 3) to pinpoint problem areas where a paucity of existing data might indicate additional basic research would be required; all of these goals have been accomplished.

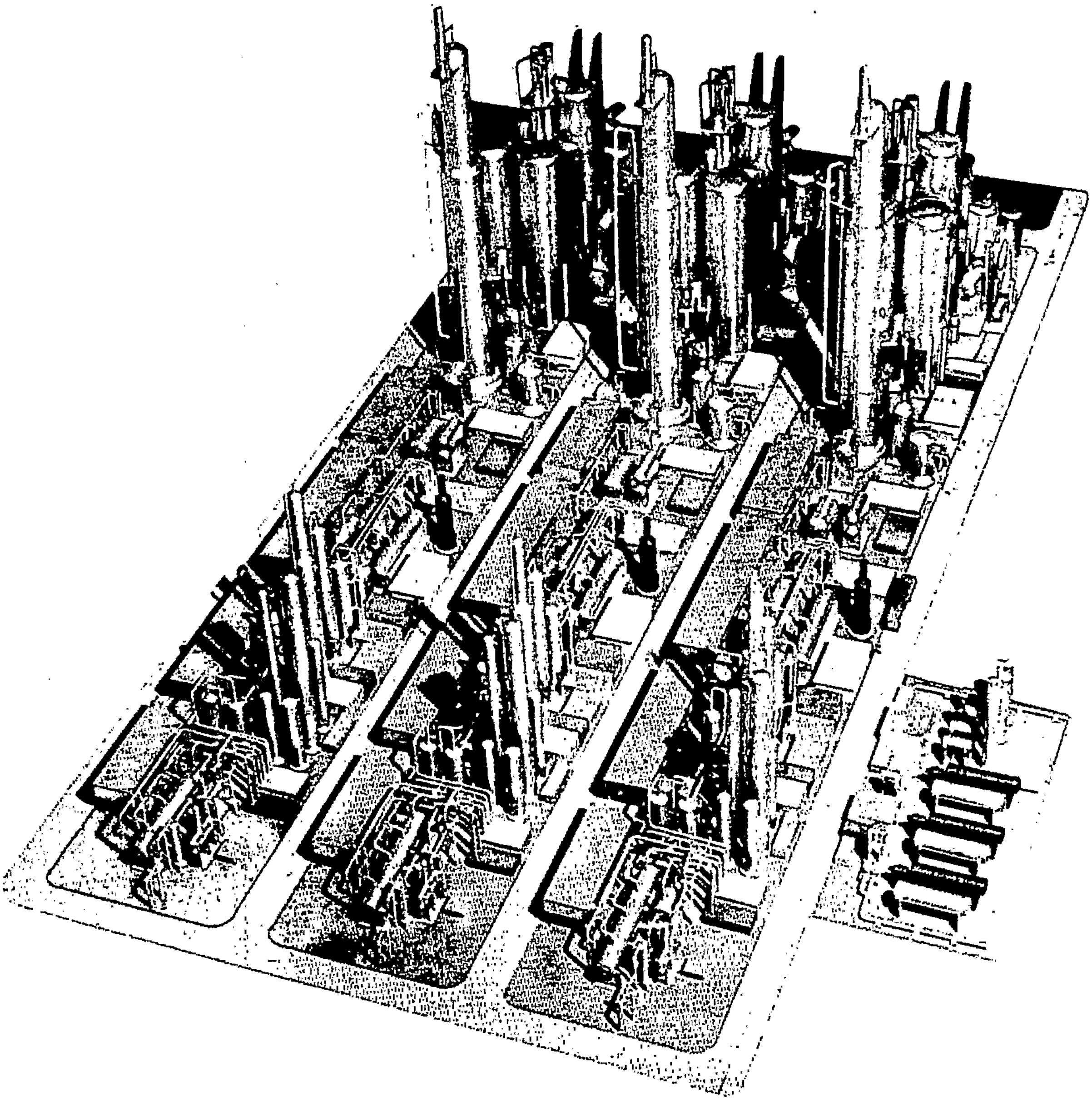
The primary impact of this design, however, has been to show that it is feasible to design large-scale plants for the hydrogasification of coal to pipeline-quality gas (SNG), with new American approaches to gasification, coupled with existing technology. As any of the desirable improvements discussed herein become available, both the economics and design of coal gasification plants will benefit. This design project indicates, however, that commercial conversion of coal to SNG is feasible today with improved American technology, without awaiting advances to optimum designs.

8.1 Introduction

IGT, in conjunction with Procon Incorporated, undertook the preliminary design of a demonstration-size HYGAS coal hydrogasification plant under the sponsorship of OCR and A. G. A. in December 1969. The design is described here and — because it is the initial design — it is also critically analyzed to identify possible design improvements.

---

* The Section 8. 7 — Suggested Areas for Design Improvement — discussed in this report is based on "Analysis of a HYGAS Coal Gasification Plant Design," presented by F. C. Schora and C. W. Matthews in November 1972 before the 65th Annual Meeting of the A. I. Ch. E. at New York City.



P7506 1357

Figure 8-1. PHOTOGRAPH OF THE MODEL OF A COMPLETE, COMMERCIAL HYGAS PLANT THAT COULD CONVERT 15,000 TO 18,000 TONS OF COAL DAILY INTO 250 MILLION CUBIC FEET OF SNG FOR PIPELINE USE. THE DEMONSTRATION PLANT DISCUSSED HEREIN COULD COMPRISE ONE PROCESS TRAIN SIMILAR TO THE THREE THAT COMPRISE THE PLANT SHOWN.

The three versions of the HYGAS Process under development differ in the methods used to generate hydrogen, which is consumed in the hydrogasification section. The three hydrogen-rich gas producing systems under active investigation at the HYGAS pilot plant site are electrothermal, steam-oxygen, and steam-iron processes. At the time the preliminary design of a demonstration-size plant was undertaken, the electrothermal process for the production of hydrogen-rich gas was most advanced at IGT; therefore, the electrothermal hydrogen-rich gas process was used in this design study. All three hydrogen-rich gas producing processes are discussed in detail in Part IV of this HYGAS final report.* The steam-oxygen gasifier has been installed (1974) in the base of the hydrogasification section of the pilot plant; steam-oxygen was studied earlier in a 6-inch-diameter test unit. The construction of a steam-iron process development unit was begun during 1974 at the HYGAS pilot plant under a separate OCR contract. The electrothermal process has been batch-tested in a 2-MW unit at the HYGAS pilot plant.

Although the hydrogen-rich gas production systems were at various stages of development when the preliminary demonstration plant design was undertaken, several benefits were attributed to designing the plant at that time. Benefits include providing a guide to future designs in both the arrangement and sequence of the processing steps, and shortening the engineering time in the final detailed design. Commercially available processes and practices have been selected, where practicable, so that unexpected problems and costs associated with these processes can be spotlighted. Another advantage is that an early design study identifies areas where information should be obtained during pilot plant operation.

8.2 Basic Criteria for Coal Gasification

The product gas from coal gasification plants must be interchangeable with natural gas, which is transported in long-distance pipelines. The gross heating value of the gas must be greater than 900 Btu/SCF. Carbon monoxide content of the gas should be less than 0.1%, and its water content should be less than 7 lb/million SCF. A wide range of gas compositions produced from liquid and solid fuels have been shown to be interchangeable with natural gas.

A high thermal efficiency in coal gasification is needed to minimize the cost of synthetic gas. Even so, the cost would be high if compared with the artificially low wellhead price of natural gas during recent years. In the HYGAS Process a high thermal efficiency is attained by the selection of pressure, temperature, residence time, and gas composition in three reaction stages. About two-thirds of the total methane is produced within the hydrogasifier. The HYGAS reactor pressure ranges from 1000 to 1500 psi, temperature in the three reaction zones increases from about 1200° to 1900°F, and the solids residence time in the reaction zones varies from 10 seconds to 85 minutes. The heat evolved by formation of methane within the gasifier is used for additional gasification by reaction of steam with carbon.

---

* This report covers HYGAS work performed under OCR Contract No. 14-01-0001-381 and the amended contract, 14-01-0001-381(1) and 14-01-0001-381(2), between July 29, 1964, and September 1, 1972.

For coal gasification plants to be environmentally acceptable, the plant design must minimize water usage and limit potential pollutant emissions such as ash, sulfur, ammonia, cyanides, phenolics, oils, and waste heat.

Virtually all streams data and process flow diagrams published in this HYGAS report are from Report for Phase II Preliminary Design Report and Budget Investment Estimate: Preliminary Engineering Study HYGAS Demonstration Plant (volume 1), submitted to IGT by Procon on April 3, 1972. In order to completely satisfy the contractual reporting requirement of the OCR-IGT contracts involved in this HYGAS work (OCR Contracts 14-01-0001-381, 14-01-0001-381(1), and 14-01-0001-381(2)), the balance of content of volume 1 of the Procon report is published here as Appendix 8-A. For the same reason, the complete content of volume 2 of the Procon report, covering preliminary equipment data and piping and instrument diagrams, is included here as Appendix 8-B.

8.3 Demonstration Plant Design Constraints

The demonstration plant size of 80 million SCF/day capacity was selected as the largest commercially feasible single-train plant. Provisions were made for expansion to 250 million SCF/day capacity. An Eastern bituminous coal was chosen for feed because its caking properties would require pretreatment facilities within the plant. The ultimate analysis (dry basis) of the fresh coal feed and pretreated char was:

	Feed (wt %)	Pretreated Char (wt %)
Carbon	71.50	71.34
Hydrogen	5.02	4.02
Sulfur	4.42	3.83
Nitrogen	1.23	1.00
Oxygen	6.53	7.51
Ash	11.30	12.30
Totals	100.00	100.00

Heating value: 13,186 Btu/lb of dry coal (HHV)
12,670 Btu/lb of dry coal (LHV)

Although as-mined coal was assumed to be the feed, an especially dirty coal would require cleaning. No site was identified for the gasification plant, which was assumed to be located near the mine mouth. About 120 million tons of recoverable coal reserves are needed to supply feed for a 250 million SCF/day plant for a 20-year period. The HYGAS Process with an electrothermal gasifier was chosen as the gasification process. Conventional on-site electrical power generation supplied power for the electrothermal gasifier. Insofar as possible, the processing steps included in the design represent the state-of-the-art practice that prevailed in 1969 at the time the design was undertaken.

8.4 HYGAS Demonstration Plant Description

The Procon Incorporated block flow diagrams presented in Figures 8-2 through 8-5 show, respectively, the preliminary design of the overall plant, of the solids handling units, of the gas treatment units, and a block flow material balance of utilities and miscellaneous units. Table 8-1 presents stream conditions, compositions, and materials balances for the various gas treatment units shown in Figure 8-4.

On some of the process flow diagrams and many of the flowstreams reports contained in this discussion, there is – in addition to stream properties and quantities – a value termed "fuel equivalent." This is a measure of the potential methane content of the stream under consideration. "Fuel equivalent" is a handy index to process efficiency through the many steps between the reactor and the product gas pipeline. Losses of high fuel equivalent deserve careful scrutiny; those of small fuel equivalent can be neglected at this stage of design.

The fuel equivalent is calculated as the sum of

Moles H_2 x 0.25
Moles CO x 0.25
Moles CH_4 x 1.00
Moles C_2H_6 x 1.75

In the stoichiometry of the shift reaction, one mole of carbon monoxide is equivalent to one mole of hydrogen. The methanation reaction requires four moles of H_2 equivalent to make one mole of CH_4. By similar logic, one mole of ethane is worth two moles of methane minus one mole of hydrogen. The pertinent reactions are:

$$CO + H_2O \rightarrow CO_2 + H_2$$

$$CO + 3H_2 \rightarrow CH_4 + H_2O$$

$$C_2H_6 + H_2 \rightarrow 2\ CH_4$$

Regarding the accuracy of figures, there may be some mismatching of stream flow quantities from one flow sheet to another. These (if any) are minor and are believed to be well within acceptable limits for this study. Additionally, in most cases the numerical values for various data have been rounded off to practical significant figures. For this reason the values shown for the various constituents of stream flows may not add up to exactly 100%.

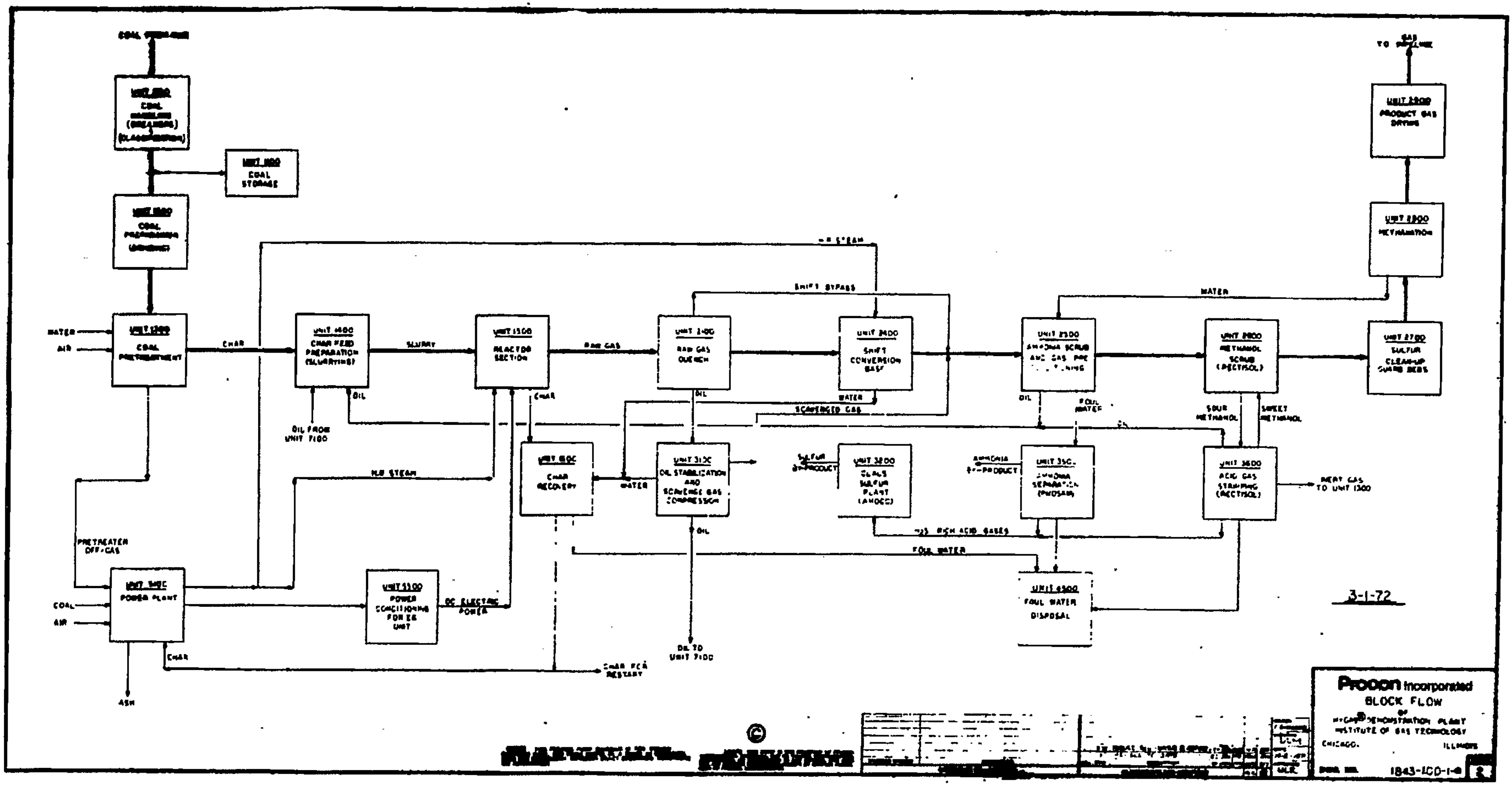


Figure 8-2. BLOCK FLOW DIAGRAM OF THE EARLY, CONCEPTUAL HYGAS DEMONSTRATION PLANT

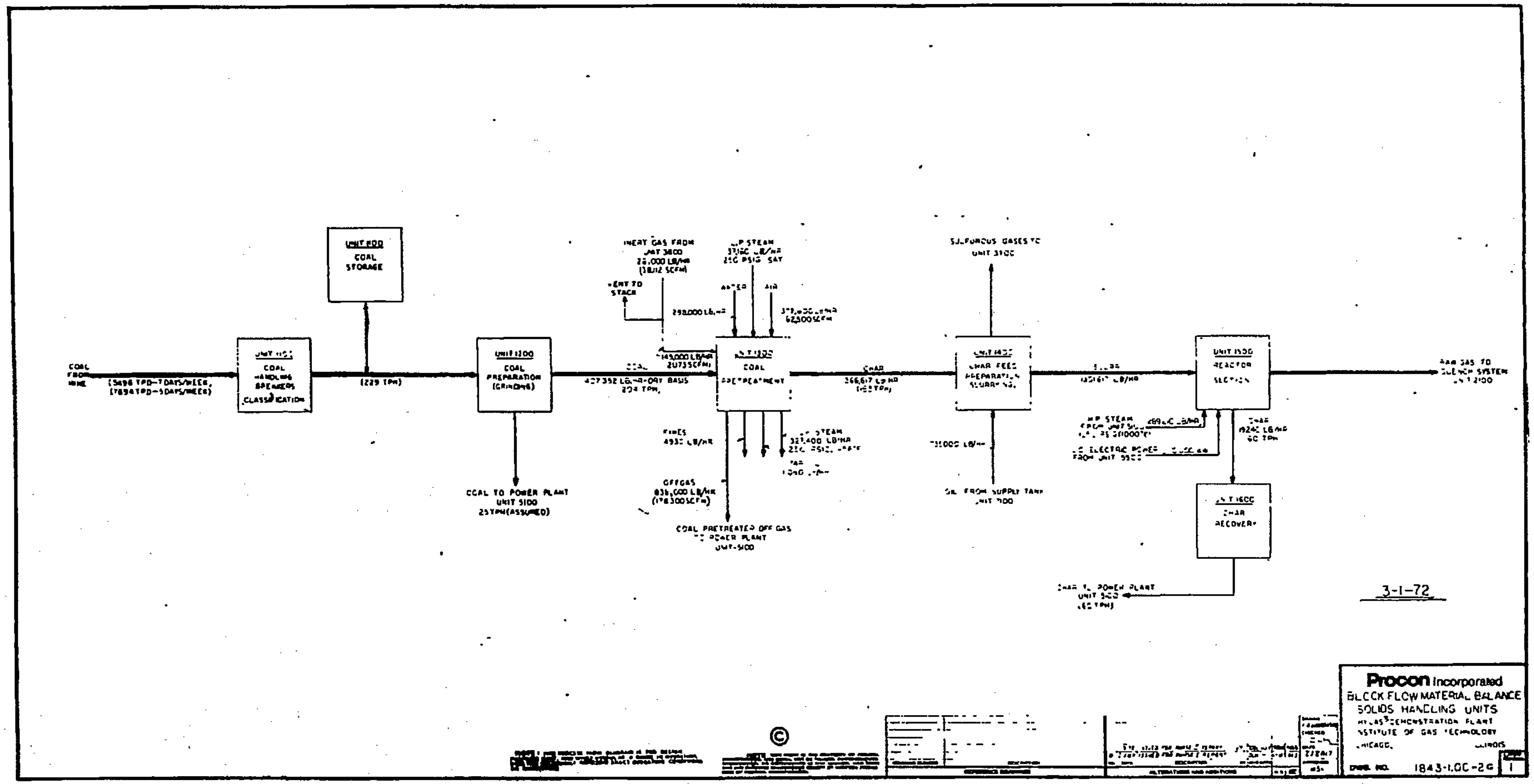


Figure 8-3. BLOCK FLOW MATERIAL BALANCE OF SOLIDS HANDLING UNITS IN THE HYGAS DEMONSTRATION PLANT

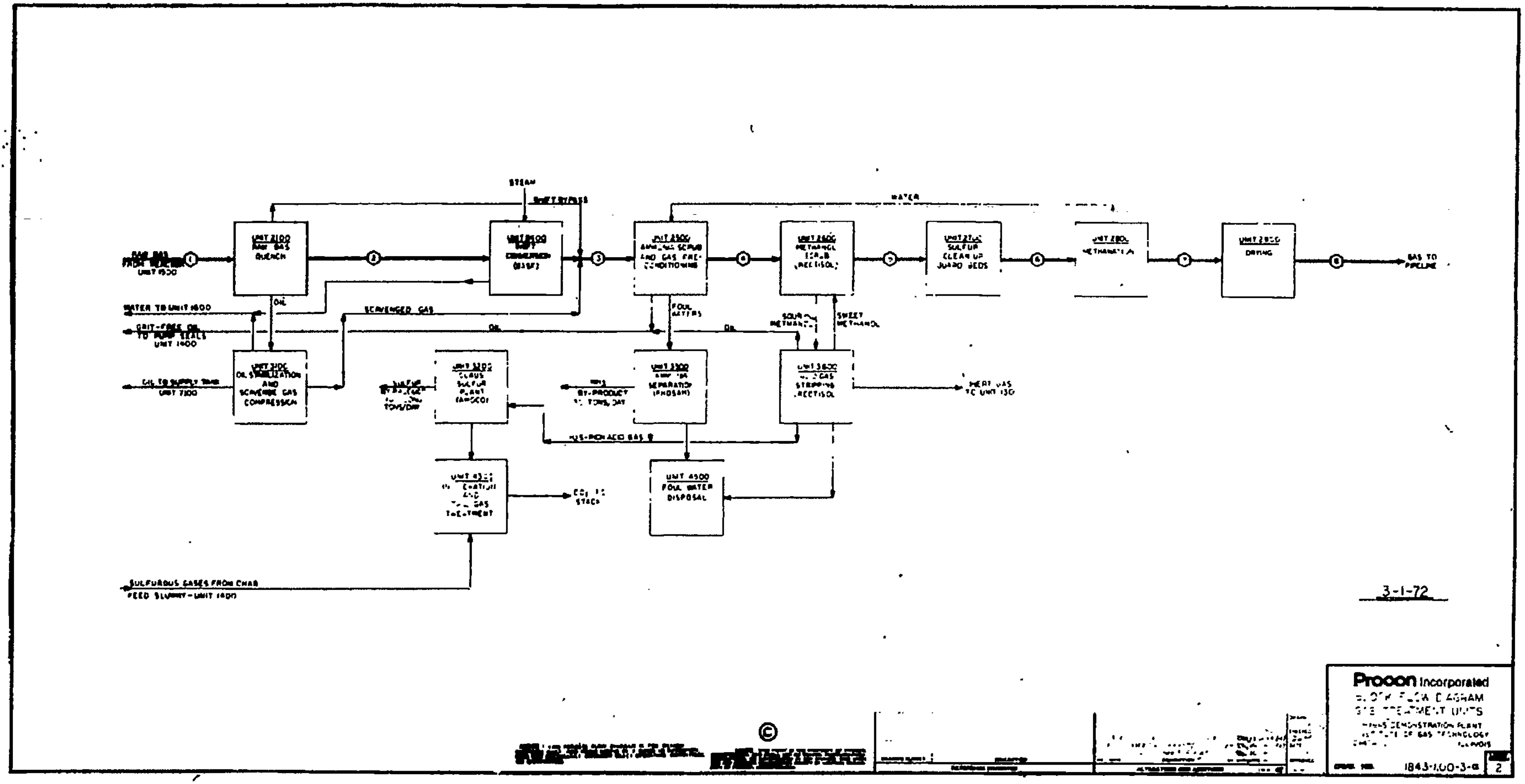


Figure 8-4. BLOCK FLOW DIAGRAM OF GAS TREATMENT UNITS IN THE HYGAS DEMONSTRATION PLANT

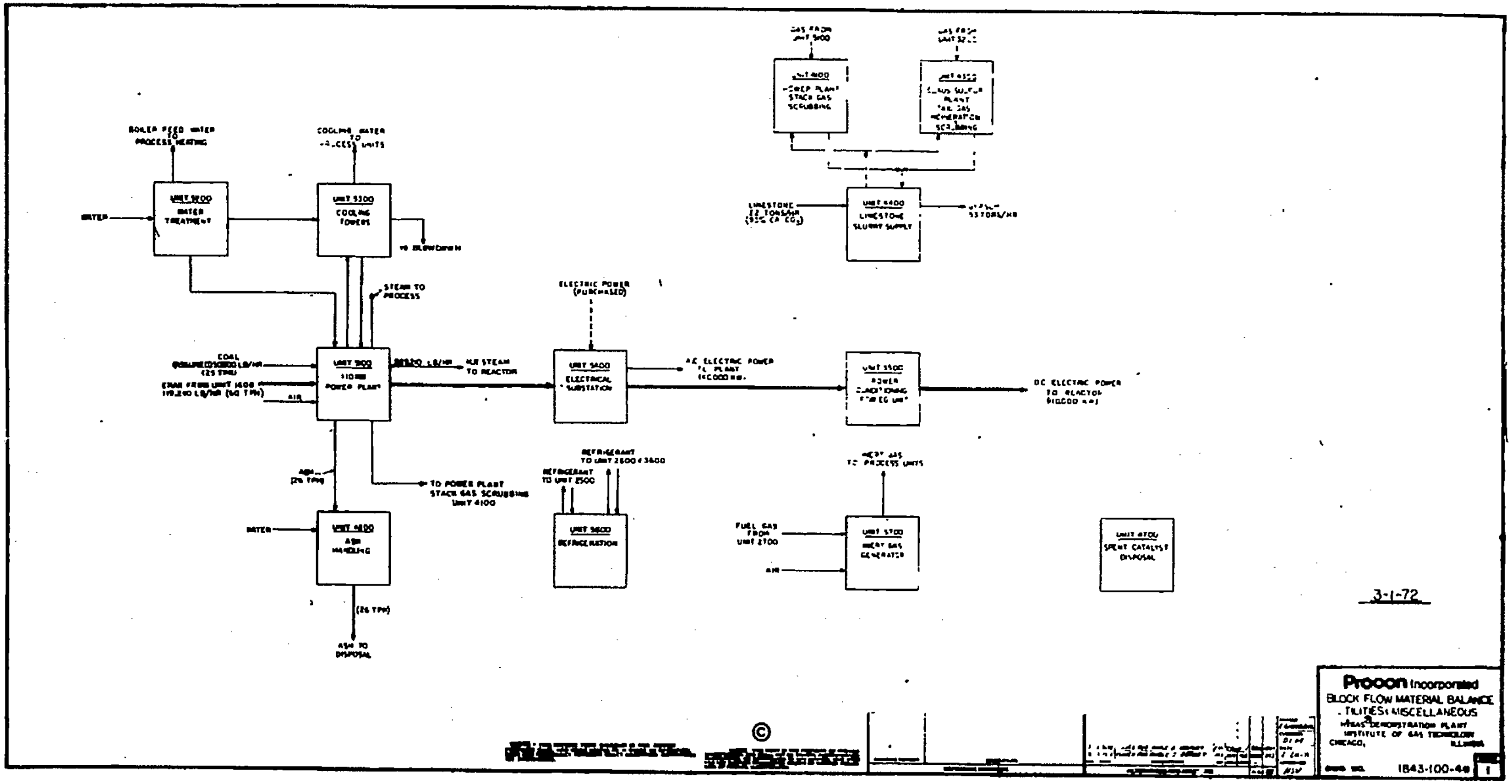


Figure 8-5. BLOCK FLOW MATERIAL BALANCE OF UTILITIES AND MISCELLANEOUS UNITS OF THE HYGAS DEMONSTRATION PLANT

Table 8-1. GAS TREATMENT STREAM DATA FOR UNITS OF THE HYGAS DEMONSTRATION PLANT SHOWN IN FIGURE 8-4

Stream Description	Raw Gas		Quenched Gas		Shifted Gas		NH_3 Free Gas		Sweetened Gas		Sweet Gas		Wet Product Gas		Gas to Pipeline	
Stream Number	1501	1	2104	2	2501	3	2506	4	2601	5	2701	6	2809	7	2901	8
Weight Flow, M lbs/hr		1356.9		297.7		565.9		483.7		182.0		182.0		131.2		131.2
MM SCFD, dry basis		247.		134.		224.		220.		156.		156.		79.7		79.7
Density at Op. Conditions		4.06		3.15		2.33		3.42		1.82		1.81		2.57		2.57
Temperature, °F		600		175		360		70		117		117		110		110
Pressure, psia		1215		1198		1134		1097		106		1053		1055		1050
Molecular Weight		42.8		20.20		20.41		20.03		10.6		10.6		14.95		14.95
Fuel Equivalent		8584		5912		8580		8533		8225		8225		8225		8225
Component	mol %	mol/hr	mol %	mol/hr	mol %	mol/hr	mol %	mol/hr	mol %	mol/hr	mol %	mol/hr	mol %	mol/hr	mol %	mol/hr
CO_2	17.47	3792.6	16.12	2349.4	26.92	6565.0	26.75	6440.0	0.15	25.6	0.15	25.6	0.11	9.5	0.11	9.5
H_2S	1.53	342.8	1.14	173.0	1.24	312.5	1.14	280.3	--	0.1 ppm	--	0.02 ppm	--	--	--	--
NH_3	0.77	165.8	0.04	6.5	0.48	117.6	--	15 ppm	--	--	--	--	--	--	--	--
Other Contaminants	0.06	14.1	0.03	4.4	0.02	5.9	--	--	--	0.6	0	0.6	--	--	--	--
Subtotal	19.83	4315.3	17.33	2533.3	28.66	7001.0	27.89	6720.3	0.15	26.2	0.15	26.2	0.11	9.5	0.11	9.5
H_2	29.20	6340.2	30.12	4434.4	37.62	9178.7	38.12	9176.2	53.27	9123.4	53.27	9123.4	7.99	699.8	7.99	699.8
CO	25.88	5620.5	26.64	3883.6	11.41	2782.9	11.55	2781.0	16.00	2776.8	16.00	2776.8	0.016	1.4	0.016	1.4
CH_4	24.18	5249.5	24.72	3604.2	21.51	5247.3	21.65	5211.0	30.38	5195.4	30.38	5195.4	91.9	8049.3	91.9	8049.3
C_2H_6	0.91	196.9	0.8	130.3	0.80	195.6	0.79	190.0	0.20	32.4	0.20	32.4	--	--	--	--
Subtotal Gas	100.00	17407.1	100.00	12052.5	100.00	17404.5	100.00	17358.2	100.00	17128.0	100.00	17128.0	100.00	8760.0	100.00	8760.0
Steam	--	4500.0	--	90.1	--	3062.8	--	5.6	--	--	--	--	--	10.5	--	1.0
Aromatics and Oils	--	5371.7	--	69.2	--	266.9	--	70.4	--	--	--	--	--	--	--	--
Total, mols/hr	--	31594.1	--	14745.1	--	27734.5	--	24154.5	--	17154.2	--	17154.2	--	8770.5	--	8761.0

8.4.1 Solids Handling

The solids handling units are designated in Figure 8-6. Figure 8-3 provides a block-flow material balance of solids handling units. Coal mining, cleaning, and transportation to the demonstration plant were not part of the design.

8.4.1.1 Coal Handling and Storage

From the nearby mine, 6-inch x 0 size run-of-mine coal is delivered to hoppers in the coal-breaker house by conveyor. A rotary breaker reduces the coal to 1-1/2 inch x 0 size. Rock and other unbroken material that is larger than this size is separated, and the crushed coal is transported by conveyor to a traveling stacker, which delivers it to the live storage pile.

The coal rate from the mine is 1350 tons/hr on a 1 shift (7 hours) day, 5 days/week basis. Gravity flow from the live storage supply is enough to feed the plant during 2-day weekends. Longer mine shutdown periods of up to 30 days are accommodated by bulldozing from dead storage to reclaim stations.

8.4.1.2 Coal Preparation

Reclaimed coal feeds through hoppers to the reclaiming conveyor located beneath the live storage pile in the reclaim tunnel. This conveyor moves coal to elevated receiving hoppers in the coal-conditioning house. Tramp iron is magnetically removed before the coal discharges into hoppers. Uncrushed coal is transferred as needed from these hoppers to the power plant for fuel. Coal for processing is crushed to 4 mesh x 0 size and is conveyed to the coal-pretreatment plant. The process flow diagram of Figure 8-7 provides details of the equipment and process flow, and Figure 8-3 provides a block-flow material balance.

8.4.1.3 Coal Pretreatment

Pretreatment destroys agglomerating tendencies of bituminous coals by mildly oxidizing the coal surfaces in air at a temperature of 800°F. The pretreater coal feed contains 8% moisture and is received from the coal-crushing plant at a rate of 204 tons/hr. Char product from pretreatment is produced at a rate of 184 tons/hr.

Process flow diagrams of this section appear in Figures 8-8 and 8-9 Stream conditions, compositions and material balances are provided in Table 8-2.

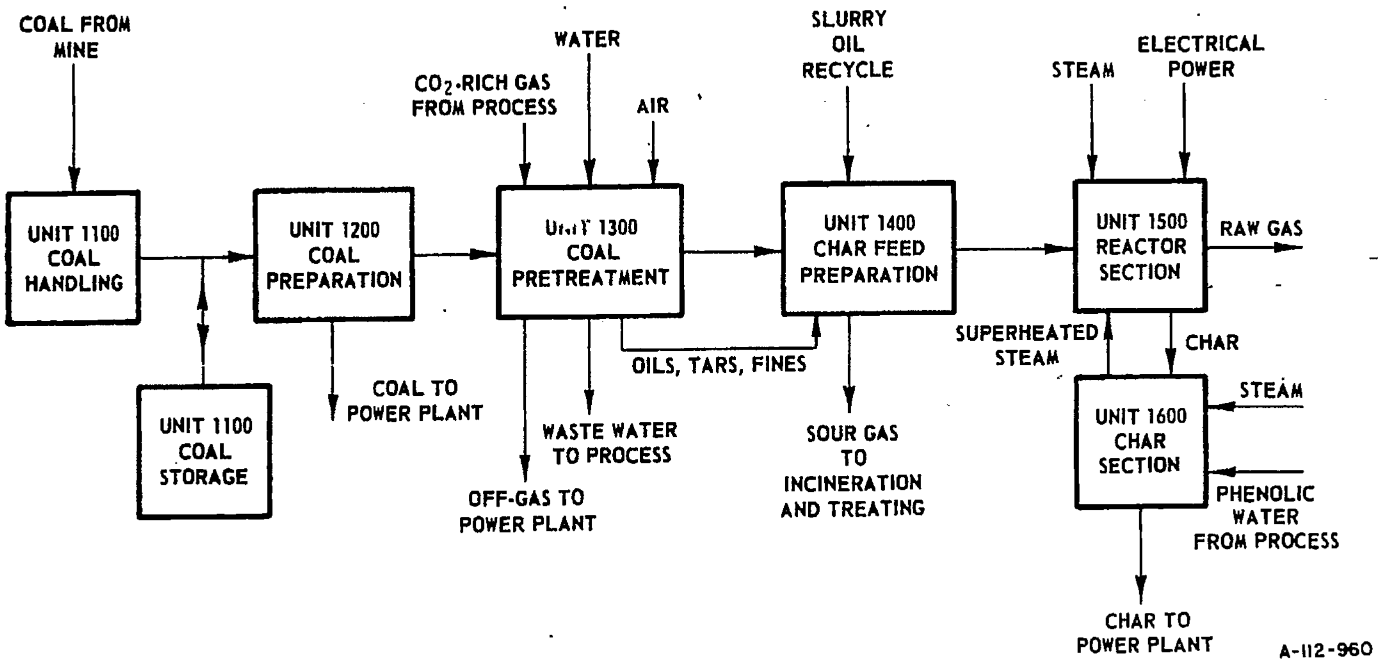


Figure 8-6. HYGAS DEMONSTRATION PLANT FEED SOLIDS HANDLING

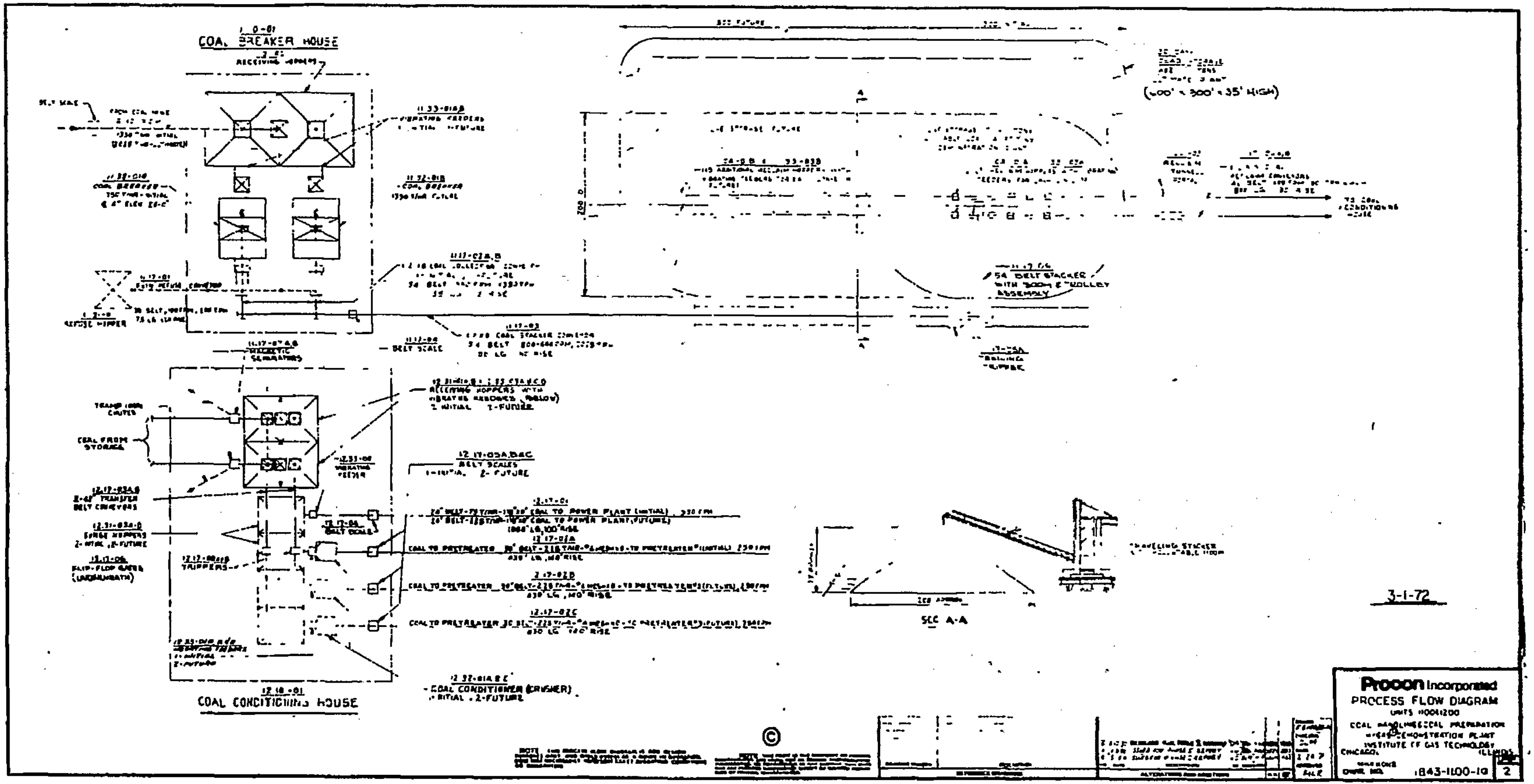


Figure 8-7. COAL HANDLING AND COAL PREPARATION PROCESS FLOW DIAGRAM FOR THE HYGAS DEMONSTRATION PLANT

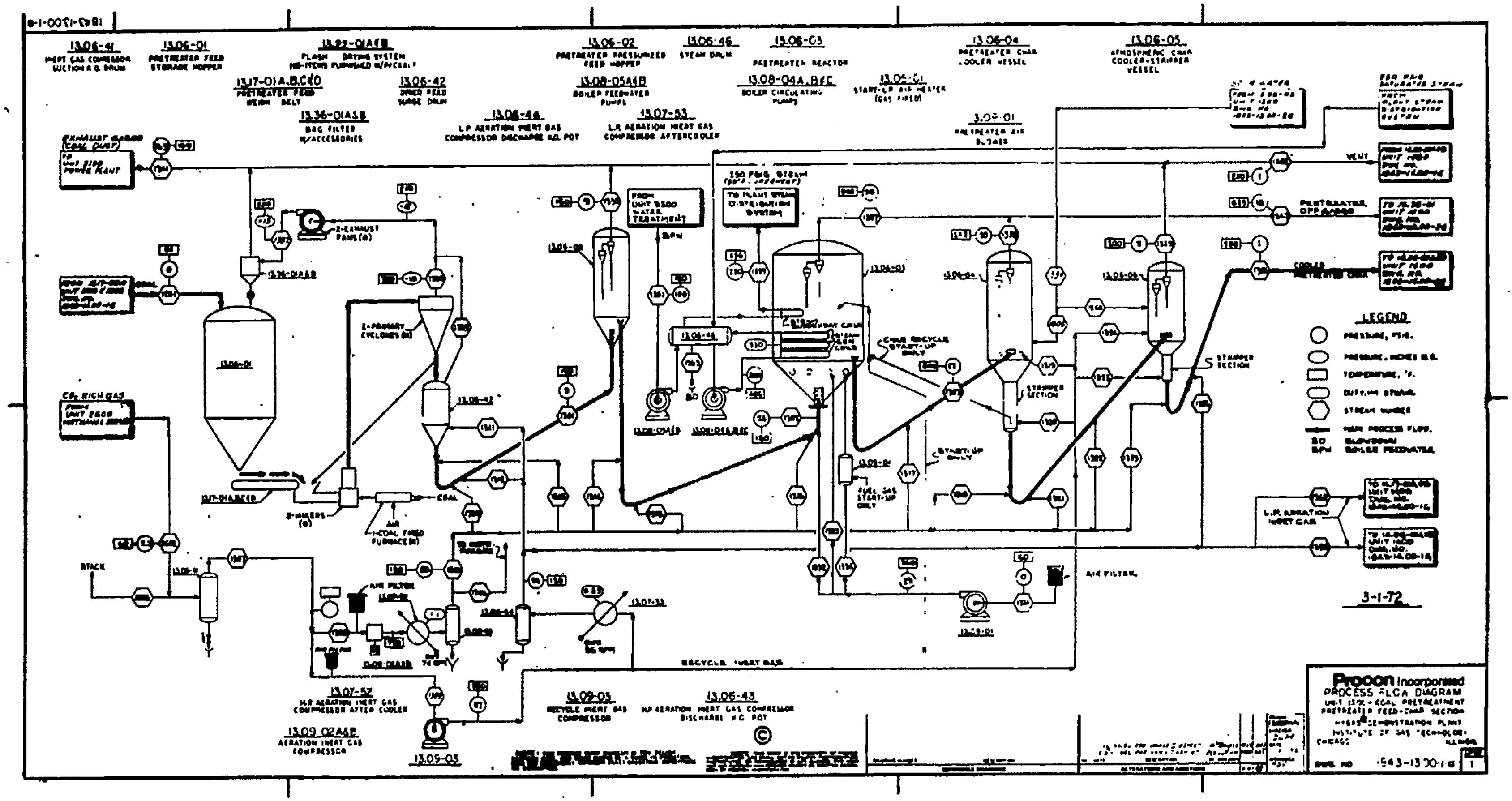


Figure 8-8. COAL PRETREATMENT, PRETREATER CHAR SECTION PROCESS FLOW DIAGRAM FOR THE HYGAS DEMONSTRATION PLANT

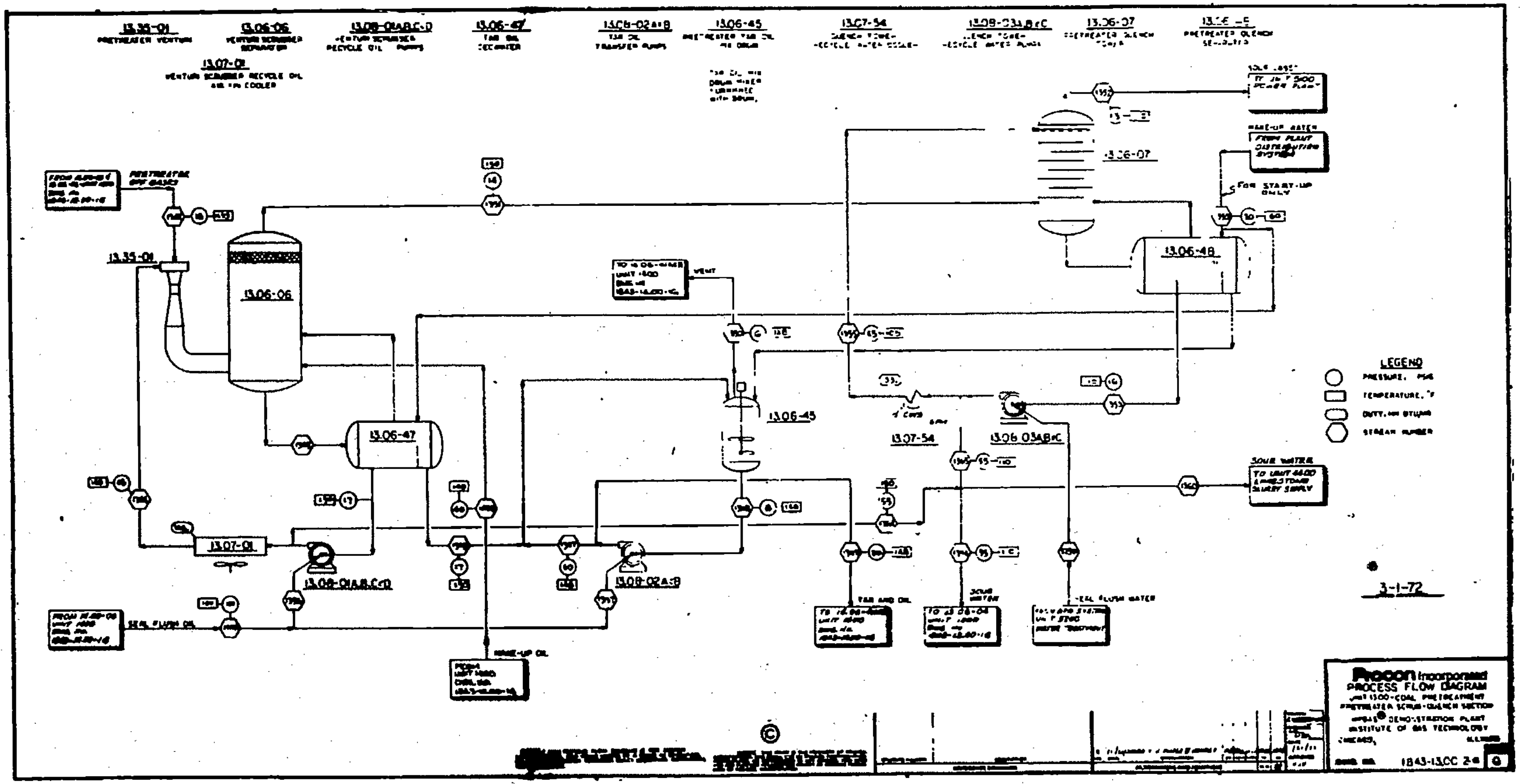


Figure 8-9. COAL PRETREATMENT, PRETREATER SCRUB-QUENCH SECTION PROCESS FLOW DIAGRAM FOR THE HYGAS DEMONSTRATION PLANT

Table 8-2, Part 1. COAL PRETREATMENT STREAM CONDITIONS, COMPOSITIONS AND MATERIALS BALANCES FOR THE HYGAS DEMONSTRATION PLANT

Stream Description	Component Molecular Weight	Dried Coal Feed	Dried Coal Feed to Pretreater Reactor	Dry Char Out of Reactor	Cooled Char to Char Feed Hopper - Unit 1400	Condensate in Pretreater Char Cooler	Inert Gas	Inert Gas	Inert Gas	Inert Gas	Inert Gas	Inert Gas	Inert Gas	Inert Gas	Inert Gas	Inert Gas	Inert Gas	Inert Gas
Stream Number		1301	1302	1303	1304	1305	1306	1307	1308	1309	1310	1311	1312	1313	1314	1315	1316	1317
Weight Flow, M lbs/hr		426.3	543.9	367.9	371.4	70.0	115.78	145.22	19.9	125.32	15.93	1.815	13.27	0.951	0.951	3.457	3.457	1.253
MM SCFD, dry basis		--	--	--	--	--	24.3	30.6	4.2	26.4	3.4	0.4	2.8	0.2	0.2	0.7	.0.7	0.3
Density at Op. Conditions		38.0	38.0	29.0	23.0	61.3	0.1278	0.1278	0.1278	0.1278	0.6894	0.3353	0.3353	0.6894	0.6894	0.6894	0.6894	0.6894
Temperature, °F		150	150	800	200	140	60	60	60	60	120	120	120	120	120	120	120	120
Pressure, psia		24	38	44.3	14.3	69	16.5	16.5	16.5	16.5	99.3	48.3	48.3	99.3	99.3	99.3	99.3	99.3
Molecular Weight		--	--	--	--	18.14	43.21	43.21	43.21	43.21	43.21	43.21	43.21	43.21	43.21	43.21	41.21	43.21
Fuel Equivalent		--	--	--	--	--	--	--	--	--	--	--	--	--	--	--	--	--
Component, mols/hr																		
Coal (lbs/hr)	--	407352	407352	--	--	--	--	--	--	--	--	--	--	--	--	--	--	--
Char (lbs/hr)	--	--	--	366617	366617	--	--	--	--	--	--	--	--	--	--	--	--	--
Surface Moist. (lbs/hr)	--	4073	4073	--	--	--	--	--	--	--	--	--	--	--	--	--	--	--
Total Sol	--	411425	411425	366617	366617	--	--	--	--	--	--	--	--	--	--	--	--	--
N_2 (mols/hr)	28.18	--	3394	--	--	--	--	--	--	--	--	--	--	--	--	--	--	--
O_2	32.0	--	889	--	--	--	--	--	--	--	--	--	--	--	--	--	--	--
Other Gases (HCN, COS, NH_3, CS_2)	--	--	--	--	--	0.4	--	--	--	--	--	--	--	--	--	--	--	--
SO_2	64.07	--	--	--	--	0.71		-	--	--	--	--	--	--	--	--	--	--
H_2O	18.02	--	30	--	--	3846.6	--	--	--	--	--	--	--	--	--	--	--	--
CO_2	44.01	331.1	175.18	27.91	99.83	6.61	2574.7	3235.3	443.3	2792	354.72	40.4	295.5	21.18	21.18	77.0	77.0	27.91
H_2S	34.03	5 ppm	--	--	--	--	5 ppm	5 ppm	5 ppm	5 ppm	5 ppm	5 ppm	5 ppm	5 ppm	5ppm	5 ppm	5 ppm	5 ppm
H_2	2.02	3.0	1.99	0.25	0.898	--	23.3	29.2	4.0	25.2	3.2	0.37	2.68	0.19	0.19	0.7	0.7	0.25
CO	28.01	0.2	0.114	0.02	0.07	0.075	1.86	2.14	0.3	1.84	0.24	0.03	0.2	0.014	0.014	0.05	0.05	0.02
CH_4	16.04	0.8	0.408	0.07	0.232	--	5.96	7.34	1.0	6.34	0.82	0.09	0.7	0.048	0.048	0.18	0.18	0.07
C_2H_6	30.07	8.8	4.662	0.74	2.646	--	67.33	85.87	11.77	74.1	9.42	1.1	7.84	0.562	0.562	2.05	2.05	0.74
CH_3OH	32.04	0.1	0.046	0.01	0.024	0.89	0.75	0.95	0.13	0.82	0.1	0.01	0.08	0.006	0.006	0.02	0.02	0.01
C_6H_6 Benzene	78.12	--	--	--	--	0.38	--	--	--	--	--	--	--	--	--	--	--	--
C_7H_8 Toluene	92.13	--	--	--	--	0.075	--	--	--	--	--	--	--	--	--	--	--	--
$C_{10}H_8$ Naphthalene	128.16	--	--	--	---	0.1	--	--	--	--	--	--	--	--	--	--	--	--
C_6H_5OH Phenol	94.11	--	--	--	2.61	2.36	--	--	--	--	--	--	--	--	--	--	--	--
C_7H_7OH Cresol	108.13	--	--	--	0.27	0.25	--	--	--	--	--	--	--	--	--	--	--	--
Total, mols/hr		344	4495	29	106.6	3858.5	2675	3360.8	460.5	2900.3	368.3	42	387	22	22	80	80	29

Table 8-2, Part 2. COAL PRETREATMENT STREAM CONDITIONS, COMPOSITIONS AND MATERIALS BALANCES FOR THE HYGAS DEMONSTRATION PLANT

Stream Description	Component Molecular Weight	Inert Gas	Inert Gas	Inert Gas	Inert Gas	Inert Gas	Inert Gas	Inert Gas	Inert Gas	Inert Gas	Outlet Gas Stream of Reactor	Gases Out of Pretreater Char Cooler	Gases Out of Atmospheric Char Cooler	Inert Gas	Inlet Air Stream of Reactor	Inlet Air Stream of Reactor	Inlet Air Stream of Reactor	Inlet Air Stream of Reactor
Stream Number		1318	1319	1320	1321	1322	1323	1324	1325	1326	1327	1328	1329	1330	1331	1332	1333	1334
Weight Flow, M lbs/hr		0.173	15.06	8.815	1.40	2.981	5.228	14.43	0.636	3.846	428.2	154.87	31.5	14.86	377.6	124.6	102.0	151.0
MM SCFD, dry basis		0.04	15.8	1.9	0.3	0.63	1.1	3.0	0.13	0.81	115.1	18.0	5.1	3.1	118.0	39.0	32.0	37.0
Density at Op. Conditions		0.6894	0.2792	0.2792	0.6894	0.6894	0.2792	0.2792	0.6894	0.3353	0.0719	0.1193	0.0753	0.1538	0.0764	0.1426	0.1426	0.1424
Temperature, °F		120	280	280	120	120	280	280	120	120	800	252	200	150	60	360	360	360
Pressure, psia		99.3	51.3	51.3	99.3	99.3	51.3	51.3	99.3	48.3	34.3	34.3	16.3	23.3	14.3	43.3	43.3	43.3
Molecular Weight		43.21	43.21	43.21	43.21	43.21	43.21	43.21	43.21	43.21	28.35	26.58	32.73	43.21	29	29	29	29
Fuel Equivalent		--	--	--	--	--	--	--	--	--	--	--	--	--	--	--	--	--
Component, mole/hr																		
Char (lbs/hr)	--	--	--	--	--	--	--	--	--	--	4930	--	--	--	--	--	--	--
Tar and Oil (lbs/hr)	--	--	--	--	--	--	--	--	--	--	1060	--	--	--	--	--	--	--
Total Sol. Liq.	--	--	--	--	--	--	--	--	--	--	5990	--	--	--	--		--	--
N_2 (mole/hr)	28.18	--	--	--	--	--	--	--	--	--	10286	--	--	--	10286	3394	2778	4114
O_2	32.0	--	--	--	--	--	--	--	--	--	145	--	--	--	2696	889	728	1079
H_2O Vapor	18.02	--	--	--	--	--	--	--	--	--	2238	3846.6	400	--	90	30	24	30
Other Gases (HCN, COS, CS_2, NH_3)	--	--	--	--	--	--	--	--	--	--	--	0.4	0.08	--	--	--	--	--
CO_2	44.01	3.85	1672.1	196.38	31.19	66.42	116.48	321.6	14.15	85.68	1277	1903	540.24	331.1	--	--	--	--
H_2S	34.08	5 ppm	5 ppm	5 ppm	5 ppm	5 ppm	5 ppm	5 ppm	5 ppm	5 ppm	--	--	--	5 ppm	--	--	--	--
SO_2	64.07	--	--	--	--	--	--	--	--	--	177	0.71	0.06	--	--	--	--	--
H_2	2.02	0.035	15.1	1.77	0.28	0.60	1.05	2.9	0.128	0.77	--	17.12	4.865	3.0	--	--	--	--
CO	28.01	0.003	1.1	0.13	0.02	0.05	0.08	0.2	0.010	0.06	653	1.325	0.358	0.2	--	--	--	--
CH_4	16.04	0.009	3.8	0.45	0.07	0.15	0.27	0.7	0.032	0.20	62.5	4.32	1.2	0.8	--	--	--	--
C_2H_6	30.07	0.102	44.4	5.21	0.83	1.76	3.09	8.5	0.376	2.27	41.5	50.35	1.282	8.8	--	--	--	--
C_3H_8	44.09	--	--	--	--	--	--	--	--	--	72	--	--	--	--	--	--	--
CH_3OH	32.04	0.001	0.5	0.06	0.01	0.02	0.03	0.01	0.004	0.02	--	1.46	0.311	0.1	--	--	--	--
C_6H_6 Benzene	78.12	--	--	--	--	--	--	--	--	--	--	0.38	0.4	--	--	--	--	--
C_7H_8 Toluene	92.13	--	--	--	--	--	--	--	--	--	--	0.075	0.005	--	--	--	--	--
$C_{10}H_8$ Naphthalene	128.16	--	--	--	--	--	--	--	--	--	--	0.1	0.01	--	--	--	--	--
Total, mole/hr		4	1737	204	32.4	69	121	334	14.7	89	14892	5826	961.4	334	13072	4313	3530	5229

A7506 1658A

Table 8-2, Part 3. COAL PRETREATMENT STREAM CONDITIONS, COMPOSITIONS AND MATERIALS BALANCES FOR THE HYGAS DEMONSTRATION PLANT

Stream Description	Component Molecular Weight	Inert Gas Out of Dried Feed Surge Drum	Vent From Flash Drier	Vent From Flash Drier And Inert Gas	Inert Gas	250 Psig Superheated Steam	Sour Water to Atmospheric Char Cooler	Exhaust Gases to Power Plant	Combustion Gases, Tar, Oils and Coal Fines	Liquid Outlet of Venturi Scrubber Separator	Cooling Water – First Quench	Tar, Oil and Coal Fines to Mix Drum	Inert Gas to Instrument Purges	Tar and Oil to Mixer	Tar and Oil	Tar and Oil to Char Slurry Mix Tank	Vent of Pretreater Tar Oil Mix Drum	Cooled Gases to Quench Tower
Stream Number		1335	1336	1337	1338	1339	1340	1341	1342	1343	1344	1345	1346	1347	1348	1349	1350	1351
Weight Flow, M lbs/hr		1.815	298.0	299.8	0.665	327.72	7.28	348.0	583.1	15061.3	14948.3	42.23	3.975	175.2	219.0	43.8	--	516.25
MM SCFD, dry basis		0.38	82.1	82.5	0.14	165.4	--	91.0	--	--	--	--	0.84	--	--	--	--	134.0
Density at Op. Conditions		0.0914	0.0537	0.0572	0.6894	0.4858	61.3	0.0604	0.0778	61.1	61.3	34.3	0.6894	59 est	59 est	59 est	--	0.1318
Temperature, °F		170	200	200	120	456	140	189	629	150	140	150	120	148	148	148	148	150
Pressure, psia		14.3	13.94	14.84	99.3	265	69	14.8	32.3	31	60	31.3	99.3	74.3	22.3	34.3	22.3	28.3
Molecular Weight		43.21	27.3	27.3	43.21	18.02	18.14	28.3	27.85	18.17	18.14	--	43.21	--	--	--	--	30.47
Fuel Equivalent		--	--	--	--	--	--	--	--	--	--	--	--	--	--	--	--	--
Component, mols/hr																		
Char (lbs/hr)		--	*	*	--	--	--	--	4930	4930	traces	4930	--	19720	24650	4930	--	--
Tar and Oil (lbs/hr)		--	--	--	--	--	--	--	1060	1060	traces	1060	--	4240	5300	1060	--	--
O_2 (mols/hr)	32.0	--	1495.3	1495.3	--	--	--	1495.3	145	--	--	--	--	--	--	--	--	145
N_2	28.18	--	7204.6	7204.6	--	--	--	7204.6	10286	--	--	--	--	--	--	--	--	10286
NH_3	17.03	--	--	--	--	--	0.04	0.04	0.19	66.32	66	--	--	--	--	--	--	--
CO_2	44.01	40.43	318.2	358.63	14.824	--	0.7	1260.21	3180	1402.68	1396	--	88.56	--	--	--	--	3173.62
H_2S	34.08	5 ppm	--	--	5 ppm	--	--		--	--	--	--	5 ppm	--	--	--	--	0.8
H_2	2.02	0.37	--	0.37	0.134	--	--	8.51	17.12	--	--	--	0.80	--	--	--	--	17.12
Other Gases (HCN, COS, CS_2)	--	--	--	--	--	--	0.04		0.21	74.35	74	--	--	--	--	--	--	0.15
CO	28.01	0.03	--	0.03	0.010	--	0.005	0.632	654.4	15.07	15	--	0.06	--	--	--	--	654.3
SO_2	64.07	--	9.3	0.3	--	--	0.06	9.36	117	148.7	148	--	--	--	--	--	--	116.3
H_2O	18.02	--	1888.6	1888.6	--	18186	400.	2288.6	6084.6	824694.6	821310	0.9	--	3.6	4.5	0.9	traces	2205.4
CH_3OH	32.04	0.01	--	0.01	0.004	--	0.15	0.421	1.46	305.46	304	--	0.03	--	--	--	--	--
CH_4	16.04	0.09	--	0.09	0.034	--	--	2.141	66.82	0.01	--	0.01	0.20	0.08	0.1	0.02	traces	66.84
C_2H_6	30.07	1.07	--	1.07	0.394	--	--	24.94	91.85	0.05	--	0.05	2.35	0.4	0.5	0.1	traces	91.8
C_3H_8	44.09	--	--	--	--	--	--	--	72	0.1	--	0.1	--	0.88	1.1	0.22	traces	71.9
C_6H_6 Benzene	78.12	--	--	--	--	--	0.04	0.04	0.38	84.29	81	2.9	--	29	36.25	7.25	traces	78.
C_7H_8 Toluene	92.13	--	--	--	--	--	0.005	0.005	0.075	17.22	15	2.15	--	20.12	25.15	5.03	traces	19.16
C_8H_{10} O-Xylene	106.16	--	--	--	--	--	--	--	--	2.6	--	2.6	--	21.6	27.0	5.4	--	6.9
C_9H_{12} Cumene	120.19	--	--	--	--	--	--	--	--	0.7	--	0.7	--	5.2	6.5	1.3	--	1.3
$C_{10}H_8$ Naphthalene	128.16	--	--	--	--	--	0.01	0.01	0.1	22.36	22	0.25	--	1.4	1.75	0.35	--	0.24
C_6H_5OH Phenol	94.11	--	--	--	--	--	0.25	0.25	--	494.64	492	0.25	--	2.04	2.55	0.51	--	1.56
C_7H_7OH Cresol	108.13	--	--	--	--	--	0.02	0.02	--	49.3	49	0.05	--	0.2	0.25	0.05	--	0.1
Light Oil (429°F)	144	--	--	--	--	--	--	--	--	162.2	--	162.2	--	664	830	166.0	--	3.8
Heavy Oil (677°F)	220	--	--	--	--	--	--	--	--	54.6	--	54.6	--	218.4	273.	54.6	--	--
Total, mols/hr		42	10916	10958	15.4	18186	401.3	12295.	20717	828095	823972	226.8	92	963.3	1204.15	240.83	--	16940

* 4000 to 10000 lbs/hr of coal fines

Table 8-2, Part 4. COAL PRETREATMENT STREAM CONDITIONS, COMPOSITIONS AND MATERIALS BALANCES FOR THE HYGAS DEMONSTRATION PLANT

Stream Description	Component Molecular Weight	Sour Gases	Cooling Water Second Quench	Sour Water to Char Coolers	Cooling Water – Second Quench	Seal Flush Oil	Seal Flush Oil	Inert Gas Purge to Unit 1400	Make-Up Water (Start-up Only)	Sour Water to Unit 4400	BFW to Steam Drum	Inert Gas Purge for Belt Conveyors Unit 1400	Blow Down	Water Condensate From Venturi Scrubber Separator	Water Condensate From Quench Tower
Stream Number		1352	1353	1354	1355	1356	1357	1358	1359	1360	1361	1362	1363	1364	1365
Weight Flow, M lbs/hr		488.02	6694.9	77.28	6668.2	--	--	1.34	--	20.08	305.09	1.525	14.53	70.7	26.66
MM SCFD, dry basis		133.9	--	--	--	--	--	0.28	--	--	--	0.32	--	--	--
Density at Op. Conditions		0.1409	61.8	61.3	61.9	50 est	50 est	0.3353	62.37	61.3	60.57	0.3353	53.6	61.1	61.8
Temperature, °F		110	110	140	105	100	100	120	60	140	180	120	406	150	110
Pressure, psia		27.3	30.3	69	60	115	115	48.3	34.3	69	115	48.3	265	70	69
Molecular Weight		31.57	18.14	18.14	18.14	--	--	43.21	18.02	18.14	18.02	43.21	18.02	18.14	18.14
Fuel Equivalent		--	--	--	--	--	--	--	--	--	--	--	--	--	--
Component, mols/hr															
Seal Flush Oil		--	--	--	--	0-20 gpm	0-10 gpm	--	--	--	--	--	--	--	--
O_2 (mols/hr)	32.0	145	--	--	--	--	--	--	--	--	--	--	--	--	--
N_2	28.18	10286	--	--	--	--	--	--	--	--	--	--	--	--	--
NH_3	17.03	--	--	0.23	--	--	--	--	--	0.09	--	--	--	0.32	--
CO_2	44.01	3171	632.32	7.31	630	--	--	29.84	--	1.9	--	33.98	--	6.68	2.53
H_2S	34.08	0.8	--	--	--	--	--	5 ppm	--	--	--	5 ppm	--	--	--
H_2	2.02	17.12	--	--	--	--	--	0.27	--	--	--	0.31	--	--	--
Other Gases (HCN, COS, CS_2)		0.15	--	0.25	--	--	--	--	--	0.1	--	--	--	0.35	--
CO	28.01	654.3	7.43	0.08	7.4	--	--	0.02	--	0.02	--	0.02	--	0.07	0.03
SO_2	64.07	116	67.37	0.77	67.1	--	--	--	--	0.2	--	--	--	0.7	0.27
H_2O	18.02	739.4	368126	4246.6	36660	--	--	--	--	1103.1	16928	--	806.5	3883.7	1466.
CH_3OH	32.04	--	--	1.04	--	--	--	0.01	--	0.42	--	0.01	--	1.46	--
CH_4	16.04	66.83	--	--	--	--	--	0.07	--	--	--	0.08	--	--	--
C_2H_6	30.07	91.75	--	--	--	--	--	0.79	--	--	--	0.90	--	--	--
C_3H_8	44.09	71.78	--	--	--	--	--	--	--	--	--	--	--	--	--
C_6H_6 Benzene	78.12	73.5	35.14	0.42	35	--	--	--	--	0.11	--	--	--	0.39	0.14
C_7H_8 Toluene	92.13	16.25	7.43	0.08	7.4	--	--	--	--	0.02	--	--	--	0.07	0.03
C_8H_{10} O-Xylene	106.16	4.1	--	--	--	--	--	--	--	--	--	--	--	--	--
C_9H_{12} Cumene	120.19	0.7	--	--	--	--	--	--	--	--	--	--	--	--	--
$C_{10}H_8$ Naphthalene	128.16	0.11	7.43	0.11	7.4	--	--	--	--	0.03	--	--	--	0.11	0.03
C_6H_5OH Phenol	94.11	0.4	224.9	2.61	224	--	--	--	--	0.68	--	--	--	2.39	0.9
C_7H_7OH Cresol	108.1	0.005	22.49	0.27	22.4	--	--	--	--	0.07	--	--	--	0.25	0.09
Light Oil (429°F)	144	traces	--	--	--	--	--	--	--	--	--	--	--	--	--
Heavy Oil (677°F)	220	--	--	--	--	--	--	--	--	--	--	--	--	--	--
Total, mols/hr		15455	369130	4259.8	367661	--	--	31.0	--	1106.8	16928	35.3	806.5	3896.5	1470.

Table 8-2, Part 5. COAL PRETREATMENT STREAM CONDITIONS, COMPOSITIONS AND MATERIALS BALANCES FOR THE HYGAS DEMONSTRATION PLANT

Stream Description	Component Molecular Weight	Vent From Char Feed Hoppers	Make-Up Oil From Unit 1400	CO_2 Rich Gas From Unit 2600	Seal Flush Water From Unit 5200	Seal Flush Oil
Stream Number		1402	1406	2602	5250	1408
Weight Flow, M lbs/hr		1.34	46.1	261.0	--	--
MM SCFD, dry basis		0.28	--	55	--	--
Density at Op. Conditions		0.0933	57.	0.1278	60.57	50 est
Temperature, °F		200	140	60	180	100
Pressure, psia		15.3	54.3	16.5	115	115
Molecular Weight		43.21	133.2	43.2	18.02	--
Fuel Equivalent		--	--	--	--	--
Component, mols/hr						
NH_3	17.03	--	0.13	--	--	--
CO_2	44.01	29.84	0.3	5810	--	--
H_2S	34.08	5 ppm	0.8	5 ppm	--	--
H_2	2.02	0.27	traces	52.5	--	--
Other Gases (HCN, COS, CO_3)	--	--	0.29	--	--	--
CO	28.01	0.02	traces	4.0	--	--
H_2O	18.02	--	5.4	--	0-15 gpm	--
CH_3OH	32.04	0.01	--	1.7	--	--
CH_4	16.04	0.07	0.03	13.3	--	--
C_2H_6	30.07	0.79	traces	154.2	--	--
C_6H_6 Benzene	78.12	--	80.9	--	--	--
C_7H_8 Toluene	92.13	--	21.3	--	--	--
C_8H_{10} O-Xylene	106.16	--	9.5	--	--	--
C_9H_{12} Cumene	120.19	--	2.0	--	--	--
$C_{10}H_8$ Naphthalene	128.16	--	0.5	--	--	--
C_6H_5OH Phenol	94.11	--	4.2	--	--	--
C_7H_7OH Cresol	108.13	--	0.4	--	--	--
Light Oil (429°F)	144	--	166	--	--	--
Heavy Oil (677°F)	220	--	54.6	--	--	--
Seal Flush Oil	--	--	--	--	--	0-30 gpm
Total, mols/hr		31	346.35	6035.7	--	--

Coal is delivered by conveyors to the pretreater feed storage hoppers. From there it is withdrawn to parallel flash drying systems, which reduce the moisture content from 8 to 1%. After separation from the drying gas, the coal is pneumatically transported by inert gas into an elevated, pressurized feed hopper. Then it moves by a standpipe into the pretreater reactor.

Residence time of the solids in the fluidized-bed pretreater reactor is 30 minutes at 800°F and 20 psig. One-third of the air injected enters with feed solids through a "birdcage" in the bottom cone of the reactor; the remaining air is introduced through two concentric air rings beneath the fluid bed. Heat is removed from the vessel by steam coils. Saturated 250-psig steam is generated from 180°F boiler feedwater at 290,000 lb/hr. An additional 37,000 lb/hr of 250-psig steam from the process joins the generated steam for superheating. The pretreated 800°F char is transferred to the char cooler vessel where it is cooled to 250°F by vaporization of injected sour water from the pretreater gas quenching system. The char is then transferred to the atmospheric char cooler in which it is further cooled to 200°F by water injection and inert gas stripping before being transferred to the char feed hoppers.

Aeration gas used in the various lines and hoppers is supplied as part of the carbon dioxide-rich vent gas from the acid-gas scrubbing section. This gas contains about 2% ethane.

Gases from the pretreater reactor and the char cooler contain tars, oils, nitrogen, carbon oxides, sulfur dioxide, light hydrocarbon gases, and steam. Quenching takes place in two stages. In the first stage, the gas is quenched to 150°F by washing with oily water in a venturi scrubber.

In the second-stage quench tower the gas is cooled to 110°F. Both stages use recirculated water for cooling. The recirculated water in the first stage is cooled by fin-fan air coolers, and the second-stage recirculated water is cooled by exchange with cooling water. Condensed oil from both stages is combined to reduce tar viscosity and transferred to the slurry preparation section. Part of the condensed water flows to the char cooler sprays, and the rest is used in slurrying limestone for the power plant stack-gas cleaning system.

Gas from the second-stage quench plus the vented gases from the pretreater, storage hoppers, dryers, and coolers, flow on to the boilers where the low-heating-value of the gas is recovered and its sulfur dioxide-content is treated by the boiler stack scrubbing system.

8.4.1.4 Char Feed Preparation

Pretreated char at 200°F is stored under inert gas in feed hoppers, which provide 10 hours of storage. In mix tanks, the char is stirred into recycled slurry oil to yield a slurry containing 33 weight percent solids. Slurry is continuously circulated through the feed pump suction lines back to the slurry tank. The feed pumps boost the reactor feed slurry pressure to 1450 psig. The slurry flows through two heat exchangers; in series, to raise its temperature to about 360°F. This helps to heat balance the gasification reactor.

As the 200°F pretreated char is stirred into the mix tank slurry oil, some sour gas flashes from the oil. The gas is compressed, cooled, and sent to incineration and final gas scrubbing.

A process flow diagram of this section is presented in Figure 8-10, and stream conditions, composition and materials balances appear in Table 8-3.

8.4.1.5 Reaction Section

The reaction section, including solids flow, is discussed as a unit under subheading 8.4.2 which follows this section.

8.4.1.6 Char Recovery

A process flow diagram of this section is shown in Figure 8-11 and Table 8-4 provides stream conditions, compositions, and materials balances.

Char enters the char recovery section at 1230°F and a pressure of about 1230 psig. From the cyclone separator, it is transferred by steam into a char-water slurrying tank. In the slurry tank, the char temperature is reduced to 573°F and the evolved steam is condensed by addition of the slurry water. The char-water slurry is reduced in pressure in four stages to minimize erosion in the letdown system. At each pressure-reduction stage, the temperature is decreased by the flashing of water. The steam evolved is condensed and returned to the slurry. After final letdown to atmospheric pressure, the solids content of the slurry is adjusted to 25 weight percent by the addition of phenol-containing process water.

The slurry is pumped to four parallel centrifuges, which yield a char concentrate containing 25 weight percent water. The char cake is dried with hot air in pulverizers and is pneumatically conveyed to the boiler fuel bunkers. Centrifuge liquor containing 6% char fines enters a settler, from which a thickened fines slurry is withdrawn and returned to the centrifuges. Part of the settler overflow water is recycled to the slurry mix tank, and the remainder flows to waste-water treatment facilities.

Phenolic process water is added to the char slurry because our laboratory has observed that spent EG char is capable of effectively reducing the phenolic content of water. Slurrying the char with phenolic process water reduces the phenolic load on the waste-water treating system.

8.4.2 Gasifier Reactor and Reaction Section

A process flow diagram of streams entering and emerging from this section is presented in Figure 8-12. Conditions, compositions and materials balances for these streams appear in Table 8-5. A second process flow diagram of streams within the pressure vessel of this reaction section is presented in Figure 8-13. Conditions, compositions and materials for the internal streams appear in Table 8-6.

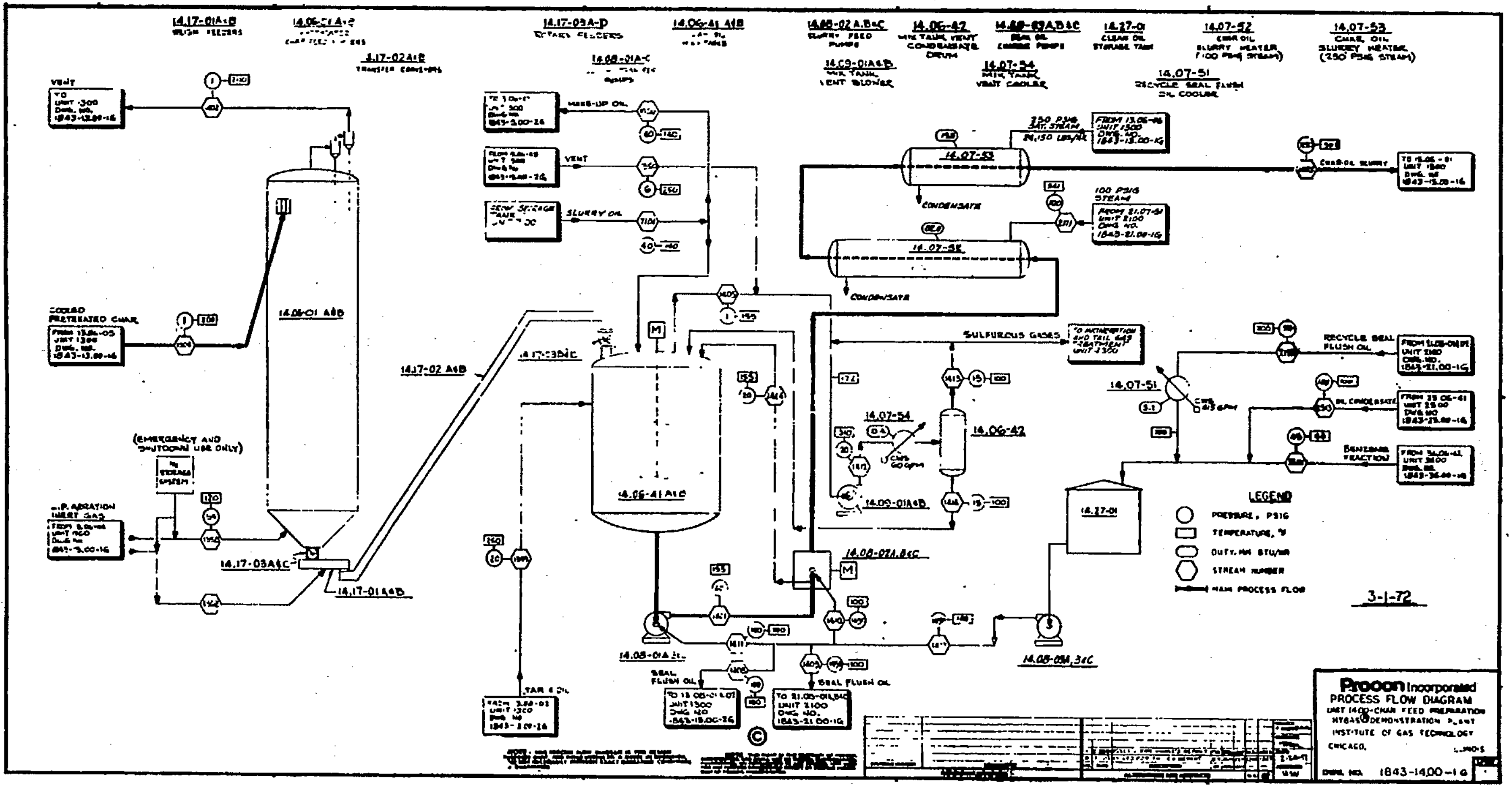


Figure 8-10. CHAR FEED PREPARATION PROCESS FLOW DIAGRAM FOR THE HYGAS DEMONSTRATION PLANT

Table 8-3. CHAR FEED PREPARATION STREAM CONDITIONS, COMPOSITIONS AND MATERIALS BALANCES FOR THE HYGAS DEMONSTRATION PLANT

Stream Description	Component Molecular Weight	Char Slurry From Mix Tanks	Inert Gas Vent From Char Feed Hoppers	Char Slurry to Reactor	Recycle Char Slurry to Mix Tanks	Vent From Mix Tanks	Make-Up Oil to Unit 1300	Seal Oil to Unit 1300/1400 Slurry Pumps	Seal Oil to Unit 1300 Slurry Pumps	Seal Oil to Unit 2100 Slurry Pumps	Seal Oil to Unit 1400 Slurry Pumps	Seal Oil to Unit 1400 Slurry Transfer Pumps	Combined Vents From 13.06-45 & 14.06-41 A&B	Vent to Unit 3100	Condensate Return to Mix Tanks
Stream Number		1401	1402	1403	1404	1405	1406	1407	1408	1409	1410	1411	1412	1413	1414
Weight Flow, M lbs/hr		1383.0	1.34	1107.6	276.5	2.47	46.1	102.0	12.0	80.0	4.0	6.0	2.47	1.86	0.61
MM SCFD, dry basis		--	0.28	--	--	0.45	--	--	--	--	--	--	0.45	0.21	--
Density at Op. Conditions		59 est	0.093	59 est	59 est	0.102	57 est	50 est	50 est	50 est	50 est	50 est	0.183	0.209	58.9
Temperature, °F		155	200	155	155	155	140	100	100	100	100	100	310	100	100
Pressure, psia		75	15.3	1265	35	15.3	54.3	1465	115	1465	1415	115	34.3	29.3	29.3
Molecular Weight		--	43.21	--	--	44.04	--	--	--	--	--	--	44.04	42.8	50.2
Fuel Equivalent		--	--	--	--	--	--	--	--	--	--	--	--	--	--
Component, mols/hr															
Char (lbs/hr)	--	464000	--	366617	92453		--	--	--	--	--	--	--	--	--
Slurry Oil (lbs/hr)	155.7	919000	--	735000	184000		46100	--	--	--	--	--	--	--	--
Tar and Oil (lbs hr)	--	--	--	1060	--		--	--	--	--	--	--	--	--	--
Coal Fines (lbs/hr)	--	--	--	4930	--	--	--	--	--	--	--	--	--	--	--
CO_2	44.01	--	29.84	--	--	36.4	--	--	--	--	--	--	36.4	36.1	0.3
H_2S	34.08	--	--	--	--	3.8	--	--	--	--	--	--	3.8	3.57	0.23
CO	28.01	--	0.02	--	--	0.03	--	--	--	--	--	--	0.03	0.03	--
H_2O	18.02	--	--	--	--	5.75	--	--	--	--	--	--	5.75	0.28	5.47
Other Gases (CS_2, H_2, NH_3)	--	--	0.27	--	--	1.42	--	--	--	--	--	--	1.42	1.25	0.17
CH_3OH	32.04	--	0.01	--	--	0.01	--	--	--	--	--	--	0.01	--	0.01
CH_4	16.04	--	0.07	--	--	0.35	--	--	--	--	--	--	0.35	0.35	--
C_2H_6	30.07	--	0.79	--	--	0.92	--	--	--	--	--	--	0.92	0.92	--
C_6H_6 Benzene	78.12	--	--	--	--	6.2	--	255 gpm	30 gpm	200 gpm	10 gpm	15 gpm	6.2	1.0	5.2
C_7H_8 Toluene	92.13	--	--	--	--	0.62	--	--	--	--	--	--	0.62	0.04	0.58
C_8H_{10} O-Xylene	106.16	--	--	--	--	0.1	--	--	--	--	--	--	0.1	--	0.1
C_9H_{12} Cumene	120.19	--	--	--	--	0.02	--	--	--	--	--	--	0.02	--	0.02
C_6H_5OH Phenol	94.11	--	--	--	--	0.06	--	--	--	--	--	--	0.06	--	0.06
Light Oil (429°F)	144	--	--	--	--	0.01	--	--	--	--	--	--	0.01	--	0.01
Total, mols/hr		--	31.0	--	--	55.69	--	--	--	--	--	--	55.69	43.54	12.15

Table 8-3, Cont. CHAR FEED PREPARATION STREAM CONDITIONS, COMPOSITIONS AND MATERIALS BALANCES FOR THE HYGAS DEMONSTRATION PLANT

Stream Description	Component Molecular Weight	Cooled Char to Char Feed Hopper	Tar and Oil to Char Slurry Mix Tank	Vent of Tar & Oil Mix Drum	Inert Gas Purge	Inert Gas Purge for Belt Conveyor	Recycle Seal Oil From Unit 2100	Oil Condensate From Unit 2500	Benzene Fraction From Unit 3600	Slurry Oil From Unit 7100
Stream Number		1304	1349	1350	1358	1362	2199	2513	3601	7101
Weight Flow, M lbs/hr		371.4	43.8	--	1.34	1.525	70.0	18.2	5.7	735.0
MM SCFD, dry basis		--	--	--	0.28	0.32	--	--	--	--
Density at Op. Conditions		23.0	59 est	--	0.3353	0.3353	50 est	51.0	51.12	57 est
Temperature, °F		200	148	148	120	120	130	100	64	140
Pressure, psia		14.3	34.3	22.3	48.3	48.3	1015	1120.7	60	75
Molecular Weight		--	--	--	43.21	43.21	--	74.32	76.4	--
Fuel Equivalent		--	--	--	--	--	--	--	--	--
Component, mols/hr										
Char (lbs/hr)	--	366617	--	--	--	--	--	--	--	--
Slurry Oil (lbs/hr)	155.7	--	--	--	--	--	--	--	--	735000
Tar and Oil (lbs/hr)	--	--	1060	--	--	--	--	--	--	--
Coal Fines (lbs/hr)		--	4930	--	--	--	--	--	--	--
CO_2	44.01	99.83	--	--	29.84	33.98	--	6.3	0.5	--
H_2S	34.08	--	--	--	5 ppm	5 ppm	--	5.0	0.4	--
H_2	2.02	0.898	--	--	0.27	0.31	--	2.1	--	--
Other Gases (HCN, COS, CS_2)	--	--	--	--	--	--	--	1.18	--	--
CO	28.01	0.07	--	--	0.02	0.02	--	1.2	--	--
H_2O	18.02	--	0.9	traces	--	--	--	1.0	--	--
CH_3OH	32.04	0.024	--	--	0.01	0.01	--	--	3.4	--
CH_4	16.04	0.232	0.02	traces	0.07	0.08	--	38.3	--	--
C_2H_6	30.07	2.646	0.1	traces	0.79	0.9	--	5.4	0.1	--
C_3H_8	44.09	--	0.22	traces	--	--	--	--	--	--
C_6H_6 Benzene	78.12	--	7.25	traces	--	--	175 gpm	138.6	65.8	--
C_7H_8 Toluene	92.13	--	5.03	traces	--	--	--	5.62	4.4	--
C_8H_{10} O-Xylene	106.16	--	5.4	--	--	--	--	19.71	--	--
C_9H_{12} Cumene	120.19	--	1.3	--	--	--	--	4.16	--	--
$C_{10}H_8$ Naphthalene	128.16	--	0.35	--	--	--	--	0.083	--	--
C_6H_5OH Prenol	94.11	2.61	0.51	--	--	--	--	0.95	--	--
C_7H_7OH Cresol	108.13	0.27	0.05	--	--	--	--	0.05	--	--
Light Oil (429°F)	144	--	166.0	--	--	--	--	21.5	--	--
Heavy Oil (677°F)	220	--	54.6	--	--	--	--	0.11	--	--
Total, mols/hr		106.6	240.83	--	31.0	35.3	--	251.3	74.6	--

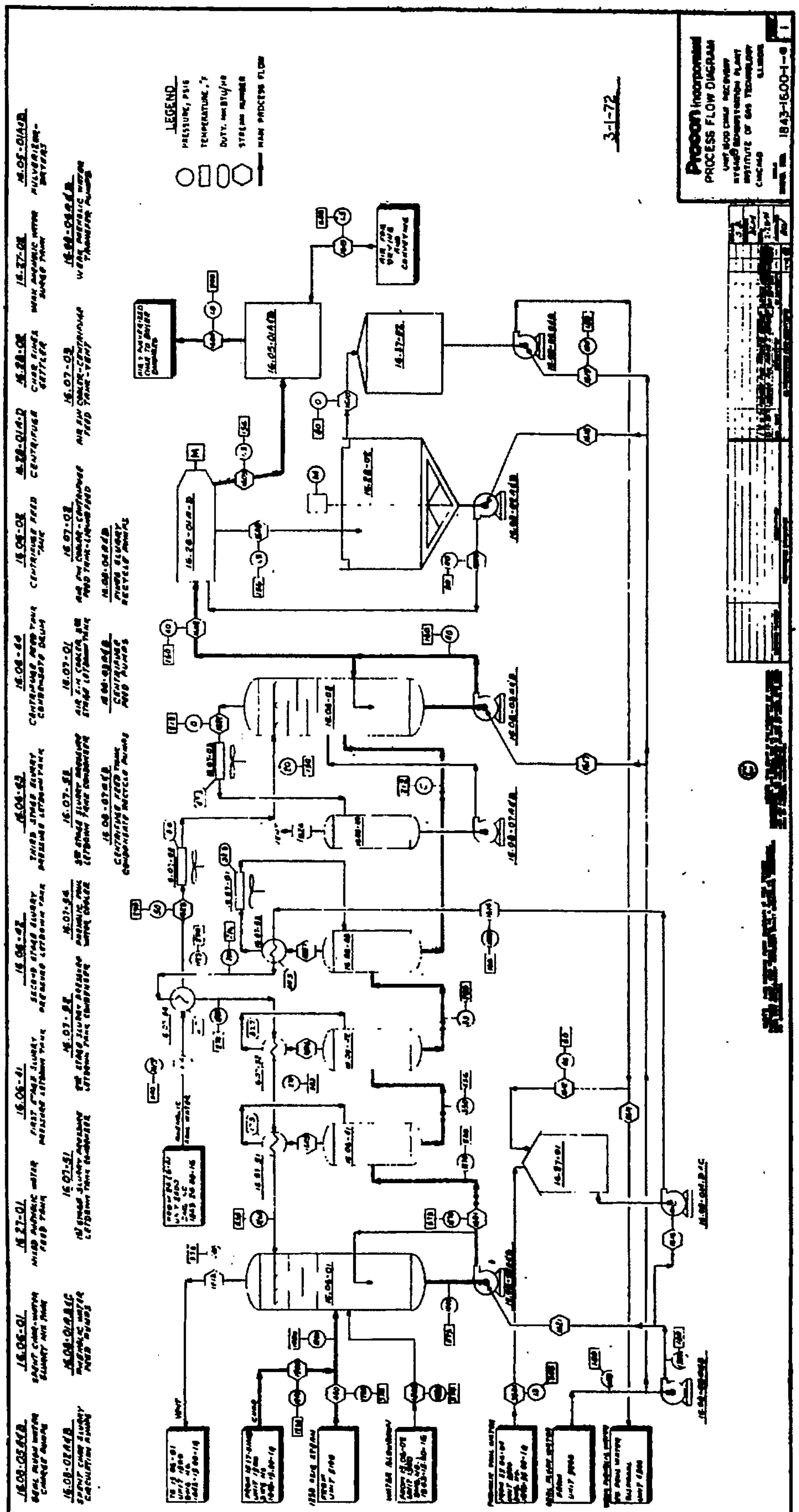


Figure 8-11. CHAR RECOVERY SECTION PROCESS FLOW DIAGRAM FOR THE HYGAS DEMONSTRATION PLANT

Table 8-4. CHAR RECOVERY SECTION STREAM CONDITIONS, COMPOSITIONS AND MATERIALS BALANCES FOR THE HYGAS DEMONSTRATION PLANT

Stream Description	Component Molecular Weight	Slurry Out of Char-Water Mix Tank	Cooled Slurry to Centrifuge	Humid Char Cake Out of Centrifuge	Air + Pulverizing Char to Boiler Burners	870 psig Steam	350 psig Steam	35 psig Steam	Water Flow Out of Centrifuge	Underflow of Thickening Loop	Overflow to Surge Tank	Foul Water to Unit 4500	Weak Phenolic Water to Feed Tank	Vent From Char-Water Slurry Mix Tank	Recirculating Water To Slurry Mix Tank	Air For Drying and Conveying	Seal Flush Water	Seal Flush Water
Stream Number		1601	1602	1603	1604	1605	1606	1607	1608	1609	1610	1611	1612	1613	1614	1615	1616	1617
Weight Flow, M lbs/hr		477.367	570.169	149.436	413.436	25.734	47.864	68.0	449.212	28.721	420.491	260.523	159.97	0.631	334.899	263.766	--	--
MM SCFD, dry basis		--	--	--	--	13.0	24.19	34.36	--	--	--	--	--	0.3	--	82.18	--	--
Density at Op. Conditions		46.31	59.85	40 est	15 est	1.95	0.8	0.1152	61.05	60.61	62.22	62.22	62.22	2.212	60.86	0.0375	60.569	60.569
Temperature, °F		573	160	156	200	530	436	280	156	80	80	80	80	513	166	680	180	180
Pressure, psia		1285	55	15.8	15.3	885	365	50	15.8	55	14.3	55	55	1255	1315	16	1315	115
Molecular Weight		--	--	--	--	18.0	18.0	18.0	--	--	--	--	--	19.54	17.98	28.64	18.02	18.02
Fuel Equivalent		--	--	--	--	--	--	--	--	--	--	--	--	--	--	--	--	--
Component, mol/hr																		
Char (lbs/hr)	--	119254	119254	119242	119242	--	--	--	7155	7143	12	7	5	--	5	--	--	--
N_2 (mole/hr)	28.18	--	--	--	7150	--	--	--	--	--	--	--	--	--	--	--	--	--
O_2	32.00	--	--	--	1890	--	--	--	--	--	--	--	--	--	--	7150	--	--
H_2O	18.02	19872.05	24947.19	1656.22	1749.22	1426.48	2653.16	3769.33	24485.97	1195.0	23290.97	13717.54	9573.43	27.78	18574.83	93.0	0.5 gpm	0.5 gpm
CO_2	44.01	4.87	5.97	0.44	2.2	--	--	--	5.81	0.28	5.53	5.26	2.27	--	4.87	1.8	--	--
H_2S	34.08	2.97	5.17	0.38	0.38	--	--	--	5.03	0.24	4.79	2.82	1.97	--	2.97	--	--	--
NH_3	17.03	42.99	46.56	3.53	3.53	3.08	5.74	8.16	45.24	2.21	43.03	25.34	17.69	0.35	43.34	--	--	--
Other Gases (HCN, COS, CS_2, H_2)	--	--	--	--	--	--	--	--	--	--	--	--	--	4.15	4.15	--	--	--
C_6H_6 Benzene	78.12	0.486	1.286	0.096	0.096	0.035	0.065	0.092	1.25	0.06	1.19	0.699	0.491	0.005	4.91	--	--	--
C_7H_8 Toluene	92.13	0.302	0.802	0.06	0.06	0.022	0.04	0.057	0.782	0.04	0.742	0.437	0.305	0.003	0.305	--	--	--
C_6H_5OH Phenol	90.1	4.869	5.449	5.403	5.403	--	--	--	0.048	0.002	0.046	0.027	0.019	--	4.869	--	--	--
Light Oil	144	0.067	0.177	0.014	0.014	--	--	--	0.171	0.008	0.163	0.096	0.067	--	0.067	--	--	--
Total, mols/hr	--	19928.6	25012.6	1666.1	10800.9	1429.6	2659.0	3777.6	24544.3	1197.8	23346.5	13750.2	9596.2	32.288	18635.9	9034.8	0-5 gpm	0-5 gpm

Table 8-4, Cont. CHAR RECOVERY SECTION STREAM CONDITIONS, COMPOSITIONS AND MATERIALS BALANCES FOR THE HYGAS DEMONSTRATION PLANT

Stream Description	Component Molecular Weight	Seal Flush Water	Seal Flush Water	1250 psig Steam From Unit 1500	Seal Flush Water	Steam Out of Centrifuge • Feed Tank	Cooled Foul Water From Unit 2400	Vent of Vapor-Liquid Separator	Spent Char From Unit 1500	Water Blow Down	Foul Water From Unit 2400	Sour Water From Unit 3600
Stream Number		1618	1619	1620	1621	1622	1623	1624	1506	1508	2404	3524
Weight Flow, M lbs/hr		--	--	23.85	--	46.369	92.88	0.078	119.249	--	92.88	174.931
MM SCFD, dry basis		--	--	12.05	--	23.41	--	0.03	--	--	--	--
Density at Op. Conditions		60.569	60.569	2.894	60.569	0.0373	57.9	0.0559	15 est	44.44	55.6	59.1
Temperature, °F		180	180	573	180	212	290	160	1230	574	340	245
Pressure, psia		115	115	1265	1315	14.3	1123	14.3	1255	1265	1133	28
Molecular Weight		18.02	18.02	18.02	18.02	18.01	18.26	26.03	--	18.02	18.26	19.35
Fuel Equivalent		--	--	--	--	--	--	--	--	--	--	--
Component, mole/hr												
Char (lb/hr)	--	--	--	--	--	--	--	--	119249	--	--	--
H_2O (mols/hr)	18.02	0-5 gpm	0-5 gpm	1325	0-5 gpm	2567.2	5076.1	0.96	--	Normally No Flow	5076.1	9001.4
CO_2	44.01	--	--	--	--	--	1.1	--	--		1.1	2.6
H_2S	34.08	--	--	--	--	--	2.2	--	--		2.2	1.0
NH_3	17.03	--	--	--	--	4.79	3.6	0.03	--		3.6	25.65
Other Gases (HCN, COS, CS_2, H_2)	--	--	--	--	--	2.01	2.01	2.01	--		2.01	4.15
C_6H_6 Benzene	78.12	--	--	--	--	0.133	0.8	--	--		0.8	--
C_7H_8 Toluene	92.13	--	--	--	--	0.083	0.5	--	--		0.5	--
C_6H_5OH Phenol	90.1	--	--	--	--	--	0.58	--	--		0.58	4.85
Light Oil	144	--	--	--	--	--	0.11	--	--		0.11	--
Total, mols/hr	--	0-5 gpm	0-5 ppm	1325	0-5 gpm	2574.2	5087.0	3.0	119249.0	--	5087.0	9040.7

A7506 1660A

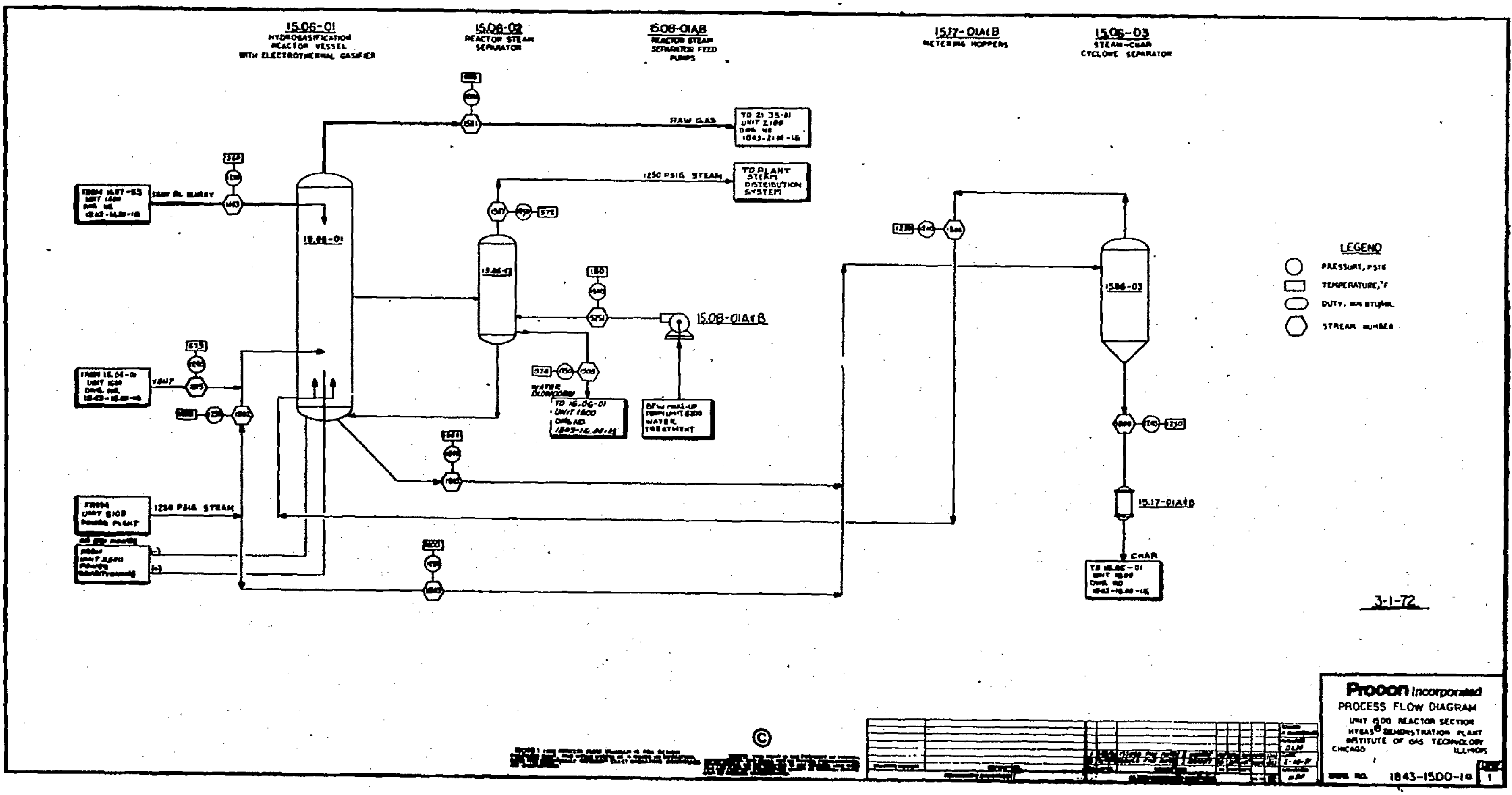


Figure 8-12. REACTION SECTION PROCESS FLOW DIAGRAM FOR STREAMS ENTERING AND EMERGING FROM REACTOR FOR HYGAS DEMONSTRATION PLANT

Table 8-5. REACTION SECTION CONDITIONS, COMPOSITIONS AND MATERIALS BALANCES FOR STREAMS ENTERING AND EMERGING FROM REACTOR FOR HYGAS DEMONSTRATION PLANT

Stream Description	Component Molecular Weight	Reactor Effluent Gases	Steam to H.G. Reactor	Steam to Steam-Char Cyclone Separator	Steam to E.G. Reactor	Spent Char From Reactor	Spent Char From Steam-Char Cyclone	Steam From Separator	Water Blowdown	Char-Slurry Feed to Reactor	Vent From Spent Char-Water Slurry Mix Tank	Make-Up BFW From Unit 5200
Stream Number		1501	1502	1503	1504	1505	1506	1507	1508	1403	1613	5251
Weight Flow, M lbs/hr		1355.894	72.224	216.986	216.986	119.249	119.249	99.2	--	1106.547	0.631	99.2
MM SCFD, dry basis		287.3	36.5	109.6	109.6	--	--	50.1	--	--	0.3	--
Density at Op. Conditions		4.572	1.461	1.461	1.230	15	15	2.763	44.44	80 est	2.212	60.57
Temperature, °F		600	1000	1000	1230	1900	1230	574	574	360	~ 537	180
Pressure, psia		1215	1265	1265	1255	1255	1255	1265	1265	1265	1255	1315
Molecular Weight		42.8	18.02	18.02	18.02	--	--	18.02	18.02	--	19.54	18.02
Fuel Equivalent		--	--	--	--	--	--	--	--	--	--	--
Component, mols/hr												
Char (lb/hr)	--	--	--	--	--	119249	119249	--	--	371547	--	--
Slurry Oil (lb/hr)		--	--	--	--	--	--	--	--	735000	--	
H_2O (mols/hr)	18.02	4500.0	4008	12041	12041	--	--	5505	--	--	27.78	5505
CO_2	44.01	3792.6	--	--	--	--	--	--	--	--	--	--
H_2S	34.08	332.7	--	--	--	--	--	--	Normally No Flow	--	--	--
HCN	27.03	8.4	--	--	--	--	--	--		--	--	--
COS	60.07	3.8	--	--	--	--	--	--		--	--	--
CS_2	76.13	1.9	--	--	--	--	--	--		--	--	--
NH_3	17.03	165.8	--	--	--	--	--	--		--	--	--
H_2	2.02	6340.2	--	--	--	--	--	--		--	--	--
CO	28.01	5620.5	--	--	--	--	--	--		--	--	--
CH_4	16.04	5249.5	--	--	--	--	--	--		--	--	--
C_2H_6	30.07	196.9	--	--	--	--	--	--		--	--	--
C_6H_6 Benzene	78.12	1411.6	--	--	--	--	--	--		--	0.005	--
C_7H_8 Toluene	92.13	331.5	--	--	--	--	--	--		--	0.003	--
C_8H_{10} O-Xylene	106.16	160.7	--	--	--	--	--	--		--	--	--
C_9H_{12} Cumene	120.19	35.2	--	--	--	--	--	--		--	--	---
$C_{10}H_8$ Naphthalene	128.16	8.1	--	--	--	--	--	---		--	--	--
C_6H_5OH Phenol	98.1	70.1	--	--	--	--	--	--		--	--	--
C_7H_8OH Cresol	108.1	7.0	--	--	--	--	--	---		--	--	--
Light Oil (429°F)	144	2523.7	--	--	--	--	--	--		--	--	--
Heavy Oil (677°F)	226	823.8	--	--	--	--	--	--		--	--	--
Total, mols, hr		31584.0	4008	12041	12041	--	--	5505		--	32.288	5505

	WT %
C	71.36
H	4.03
O	7.51
N	0.99
S	3.81
Ash	12.30
Total Solids	371,547 Lb/Hr
Light Oil	735,000 Lb/Hr

	WT %
C	71.36
H	4.03
O	7.51
N	0.99
S	3.81
Ash	12.30
Total	371,547 Lb/Hr

	WT %
C	80.89
H	0.94
O	--
N	0.42
S	1.59
Ash	16.16
Total	282,875 Lb/Hr

	WT %
C	75.04
H	0.87
O	--
N	0.56
S	2.11
Ash	21.42
Total	213,466 Lb/Hr

	WT %
C	57.76
H	0.67
O	--
N	0.99
S	2.75
Ash	37.83
Total	119,249 Lb/Hr

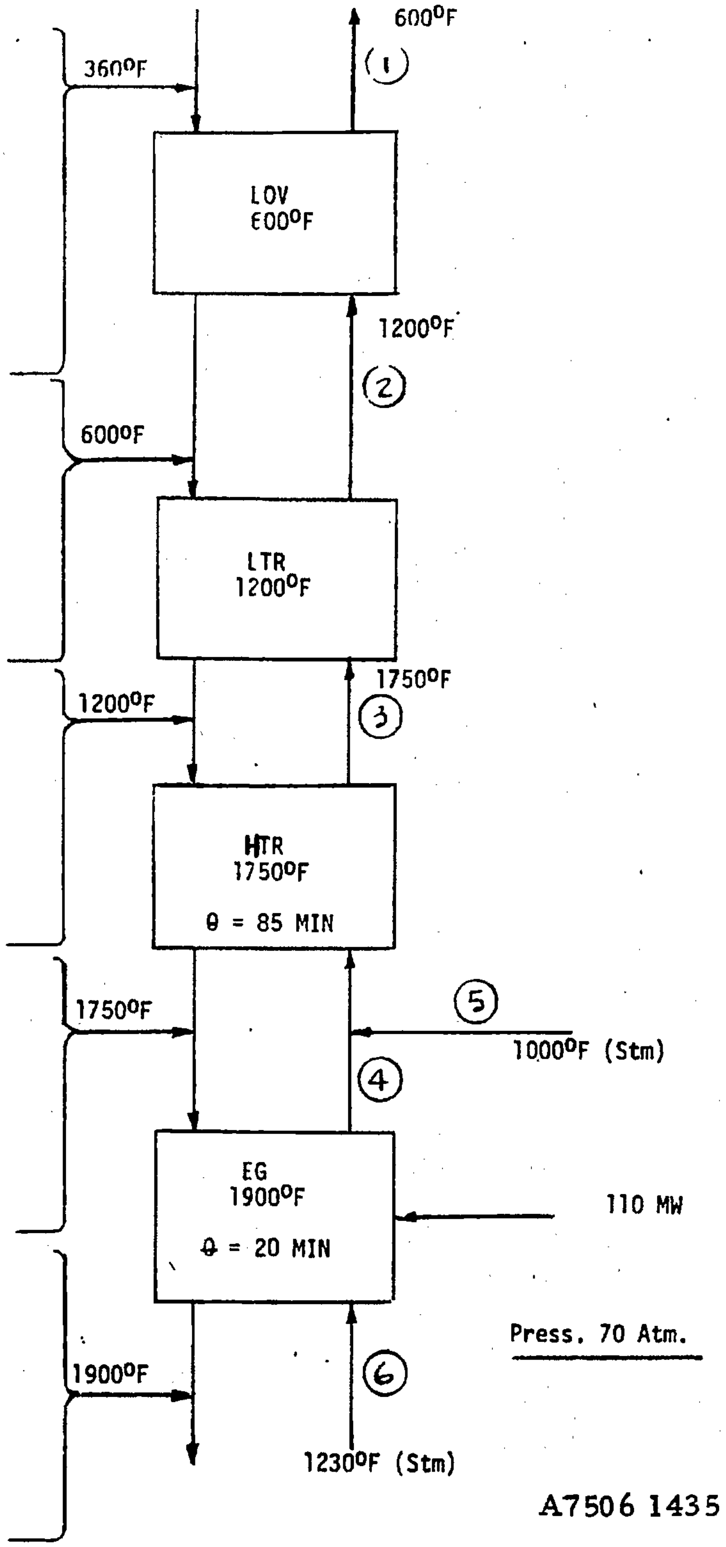


Figure 8-13. REACTION SECTION PROCESS FLOW DIAGRAM FOR STREAMS INSIDE REACTOR FOR HYGAS DEMONSTRATION PLANT

Table 8-6. REACTION SECTION CONDITIONS, COMPOSITIONS AND MATERIALS BALANCES FOR STREAMS WHILE CONTAINED IN REACTOR FOR HYGAS DEMONSTRATION PLANT

STREAM	①		②		③		④		⑤	⑥
TEMP, °F PRESS, PSI	600°F 1200 PSI		1200°F 1210 PSI		1750°F 1230 PSI		1900°F 1250 PSI		1000°F 1250 PSI	1230°F 1250 PS
	MOL/HR	MOL %	MOL/HR	MOL %	MOL/HR	MOL %	MOL/HR	MOL %	MOL/HR	MOL/HR
CO	5621	17.80	5621	21.35	5053	24.19	4805	26.79	--	--
CO_2	3793	12.01	3793	14.41	3446	16.49	1570	8.75	--	--
H_2	6340	20.07	6340	24.09	3718	17.80	6335	35.31	--	--
H_2O	4500	14.25	4500	17.09	4087	19.56	4147	23.12	4008	12,041
CH_4	5250	16.62	5250	19.94	4553	21.79	1046	5.83	--	--
C_2H_6	197	0.62	197	0.75	--	--	--	--	--	--
C_6H_6	65	0.21	65	0.25	--	--	--	--	--	--
NH_3	166	0.53	166	0.63	--	--	--	--	--	--
H_2S	333	1.05	333	1.27	36	0.17	36	0.20	--	--
L.Oil	5319	16.84	59	0.22	--	--	--	--	--	--
Total	31,584	100.00	26,324	100.00	20,893	100.00	17,939	100.00	4008	12,041
SCF/Hr. ACF/Hr. ACF/Sec.	11,970,340 296,560 82		9,976,800 382,040 106		7,918.450 397,200 110		6,798,880 358,430 100		1,519,030 49,430 14	4,563,540 176,410 49

Total Coal Required = 371,547 lb/hr = 4460 ton/day
Total Slurry = 1,106,547 lb/hr
Total Steam = 289,200 lb/hr = 6.94 million lb/day
Total Power = 110 MW

A7506 1662

The gasification reactor contains four fluidized beds through which char passes downward, counter to the upward flow of gas and steam. Figure 8-14a briefly identifies typical operating conditions. Figure 8-14b is a simplified conceptual preliminary design drawing of the reaction section. Figure 8-14c is an engineering drawing of the conceptual HYGAS demonstration plant reactor section for the electrothermal case as executed for IGT by Bechtel Corp.

Pretreated char feed at 360°F enters the top bed of the reactor as a slurry in light aromatic oil, which is produced by the process. Solids concentration in the slurry is about 33 weight percent. The top bed is called the light oil vaporizer (LOV), and operates at 600°F and 1200 psig. Residence time for the solids is 10 minutes. The slurry sprays into the fluidized bed of dried char and the slurry oil is stripped out of the bed by the hot fluidizing gases rising from below. Raw gas exits from the LOV at a temperature of 600°F.

Dried char flows into the low-temperature reactor (LTR), where it contacts hot gas from the high-temperature reactor (HTR). In the LTR, the char is lifted by the hot gas through a transport zone where the average temperature is 1200°F. The char rapidly devolatilizes and transient active carbon sites react directly with hydrogen in the gas to form additional methane. About one-half of the total methane formed in the hydrogasifier is produced by this rapid-rate methanation reaction. Total solids time in the LTR is 10 seconds, and about 24% of the coal is gasified. Gas and solids are separated by cyclones; the gas flows to the LOV and the solids flow splits, part returning to the low-temperature reactor and part flowing on to the high-temperature reactor. The gas-solids contacting for all three hydrogen-process combinations with HYGAS are shown in Figure 8-15.

The reacted solids returning to the LTR as recycle serve two purposes: 1) the increased solids flow through the LTR helps to modify temperature fluctuations and assists in the rapid transfer of heat to the cool char entering from the LOV, and 2) by diluting the LOV char with partially reacted char, any residual tendency for LOV char to agglomerate while being heated is prevented. If this concept is effective, perhaps external coal pretreatment can be avoided. The ratio of recycled char to fresh char feed in the LTR is about 2.

Char from the LTR flows into the HTR, where it reacts with fluidizing gas from the electrothermal gasifier at 1750°F. About 19% of the coal is gasified during the solids residence time of 85 minutes in the fluid bed. Some additional direct methane formation occurs in the HTR, where steam reacts with char to form hydrogen and carbon monoxide.

Most of the hydrogen consumed in the LTR and HTR stages of hydrogasification is generated in the electrothermal gasifier (EG). The EG is a fluidized-bed reactor in which char from the HTR reacts with steam at 1900°F About 25% of the coal is gasified in the EG during a solids residence time of 20 minutes. Although a small amount of direct methane formation occurs, the greatest reaction is steam with carbon. A large amount of heat is required for the net endothermic reaction. This heat is supplied to the EG by the resistance of the fluidized bed of char to the passage of a direct-current electricity.

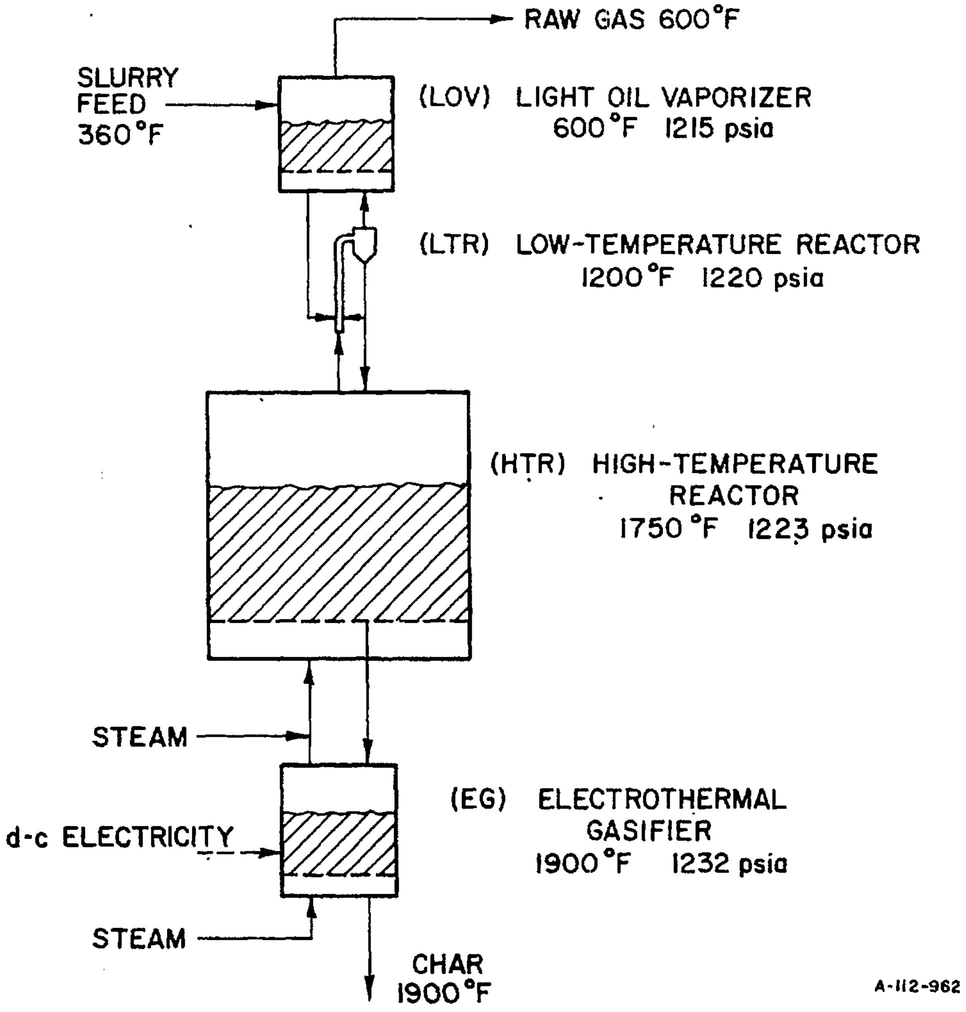
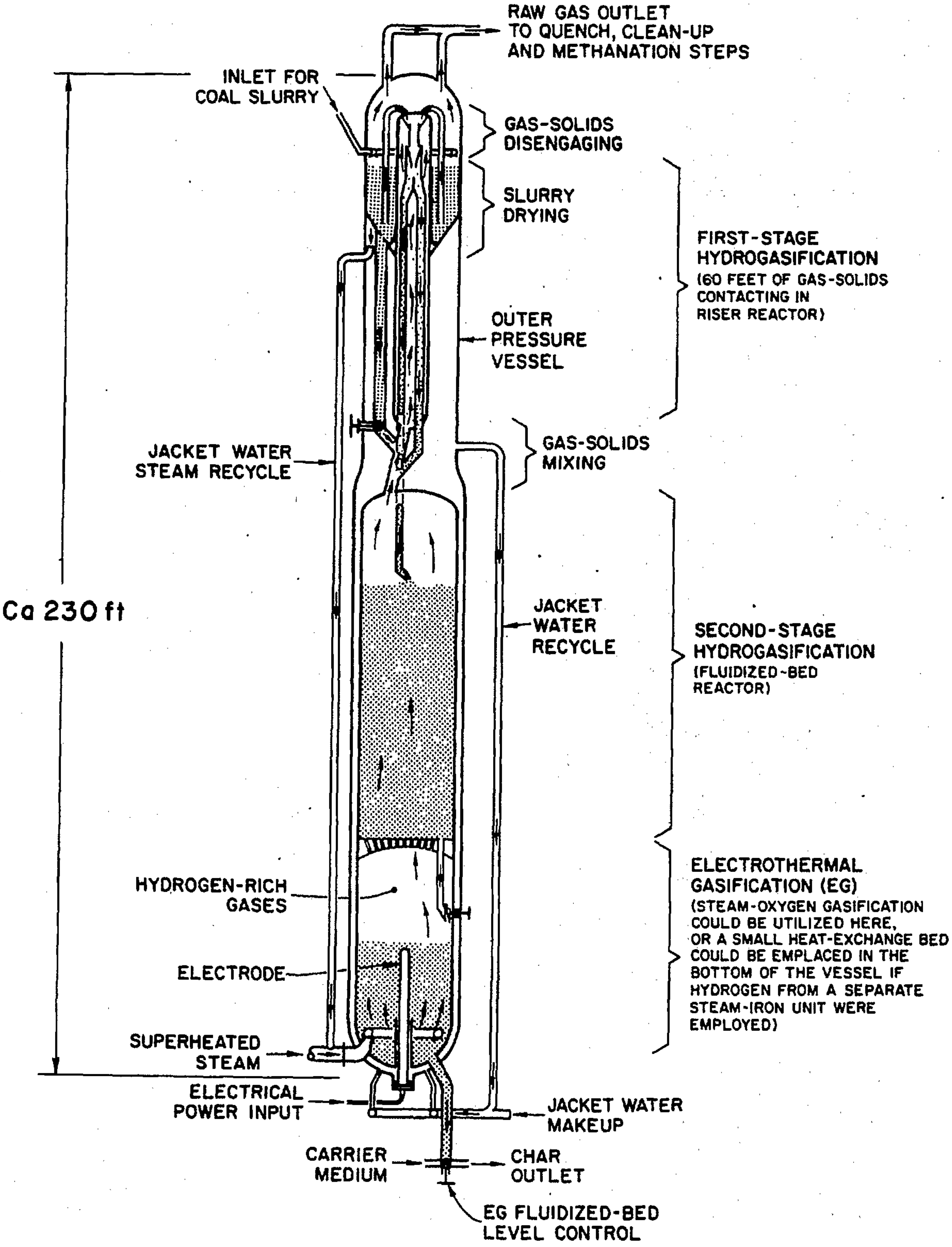


Figure 8-14a. HYGAS REACTOR WITH ELECTROTHERMAL GASIFIER



D-64-940

Figure 8-14b. SIMPLIFIED DRAWING OF CONCEPTUAL REACTOR DESIGN FOR THE HYGAS DEMONSTRATION PLANT

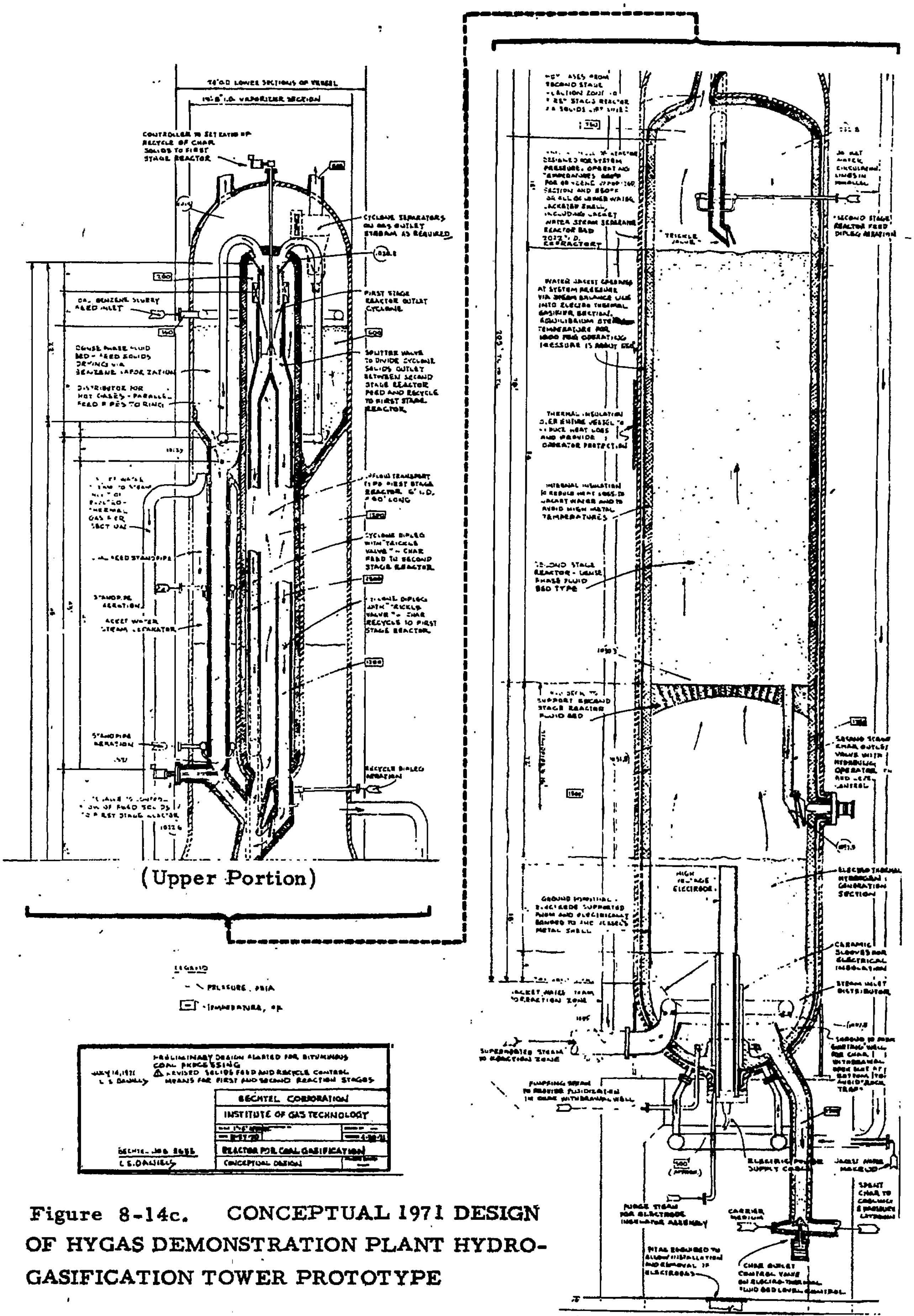


Figure 8-14c. CONCEPTUAL 1971 DESIGN OF HYGAS DEMONSTRATION PLANT HYDROGASIFICATION TOWER PROTOTYPE

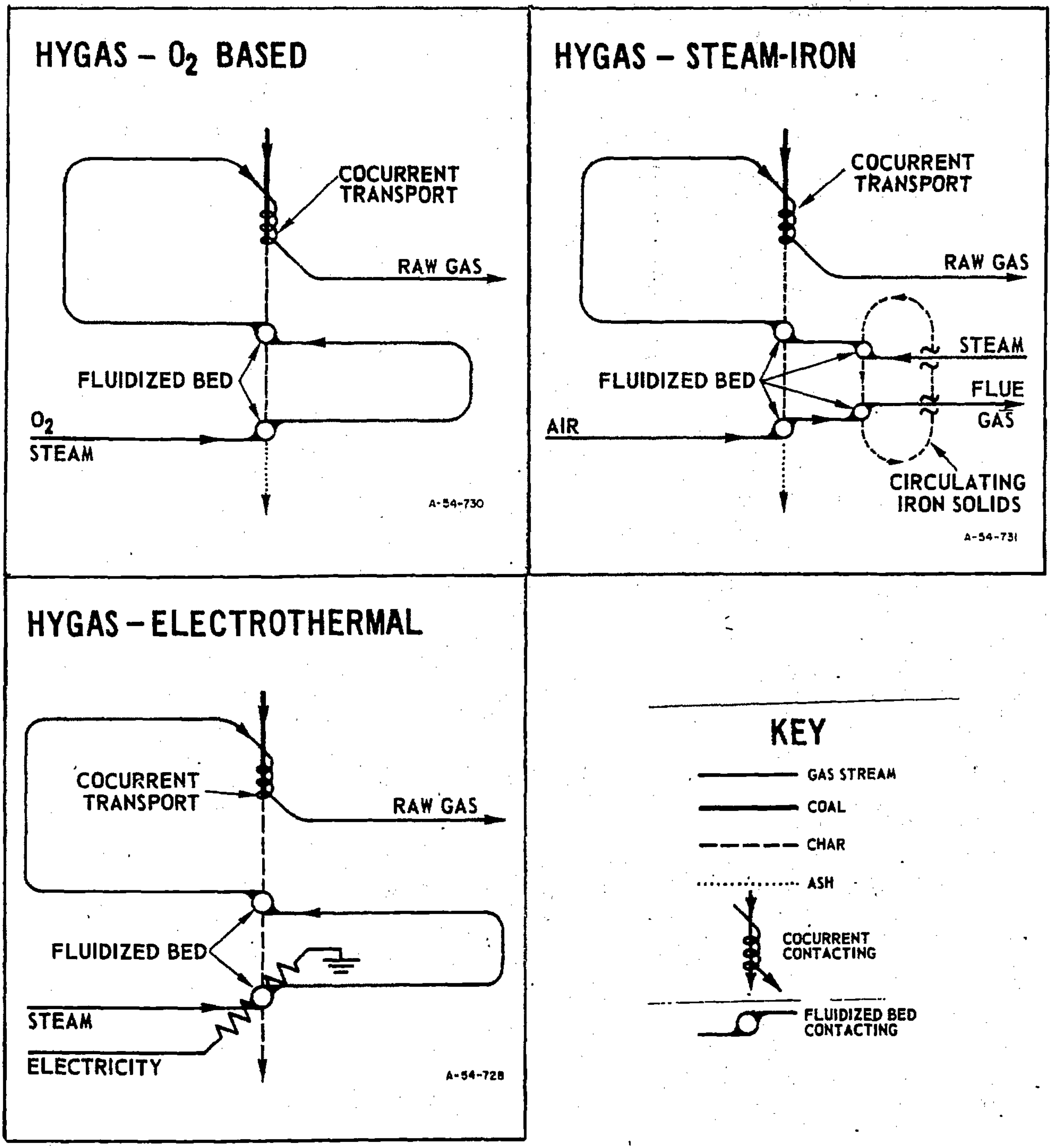


Figure 8-15. GAS SOLIDS CONTACTING FOR THREE HYDROGEN-RICH GAS PRODUCING PROCESSES INTEGRATED WITH THE HYGAS PROCESS

About 33% of the fresh char feed to the LOV leaves the EG as a net product. The carbon content of this char is the most difficult portion of the coal to gasify. After cooling, depressurizing, and recovery, the EG char is used as fuel to generate power for the EG unit and for process steam. The EG char contains no volatile matter and has the following composition·

	wt %
C	57.76
H	0.67
O	--
N	0.99
S	2.75
Ash	37.83
Total	100.00

Electrically, the EG bed is a nonlinear resistive load. Because of bed fluctuations, the load is unstable, varying in resistance over wide ratios causing the current to vary in the same manner. The ratio of maximum to minimum values of current can be as high as 4, and the character of the current is nonsinusoidal. Because of these characteristics, a d-c system is preferred. Voltage is supplied at about 15 kV above ground, which is large enough to require careful design of all power take-offs as well as of the electrode design configuration.

Of several designs that were considered for the gasification reactor, the preferred design consisted of four fluidized-bed zones, mounted one above the other, with the LOV uppermost in the vessel and the electrothermal gasifier (EG) at the bottom. The "hot" zones, i.e., the HTR and the EG, are internally protected with 1-ft-thick insulation. The insulation is installed on the inner alloy shell of a balanced-pressure, boiling water jacket. The water jacket's temperature will be about 570°F. The external vessel shell contains the pressure. With this design, a) fluidized-bed heat losses generate high-pressure steam, which is used in the process; b) the alloy inner shell is relatively thin because of the balanced-pressure design; c) the outer pressure-containing shell can be made of carbon steel; d) condensation is avoided in the refractory lining; and e) the water jacket protects the vessel's integrity if internal refractory is lost during operation.

An initial step in preparing the demonstration plant design was to establish the plant size, that is, "the largest commercially feasible single-train plant." A detailed investigation was made to determine the size and construction of a reactor for this plant. The study resulted in the selection of a 24-ft-outside-diameter reactor because the technology was available to construct pressure vessels of this size in the field. The reactor capacity, which sets the size and capacity of the other plant equipment, is 80 million SCF/day of pipeline-quality gas.

The reactor has an outside diameter of 24 feet and is 240 feet tall, including skirt height. The weight of steel in the reactor is about 2750 tons.

Steam enters the reactor process area at 1250 psig and 1000°F. About 25% of the steam passes into the reactor above the EG fluidized bed and mixes with the hydrogen-rich gas flowing to the HTR. The rest of the steam is used to transport excess EG char to a cyclone separator. The char temperature is reduced from 1900° to 1230°F as the transport steam is heated. Steam from the cyclone separator enters the EG bed as the fluidizing and reacting gas to produce synthesis gas. The cooled char flows on to the char recovery section.

8.4.3 Main Stream Gas Treatment

The raw gas effluent from the gasifier must be cleaned, its composition adjusted, and the gas methanated and dried to produce pipeline-quality gas. The block flow diagram for this part of the plant is shown in Figure 8-16, and stream data have been presented in Table 8-1.

8.4.3.1 Raw Gas Quench

Gas flows from the reactor at a temperature of 600°F and a 1215-psia pressure. It has the following composition:

	mol %
CO	17.80
CO_2	12.01
H_2	20.07
H_2O	14.25
CH_4	16.62
C_2H_6	0.62
C_6H_6	0.21
NH_3	0.53
H_2S	1.05
Light Oil	16.84
Total	100.00

Small amounts of the following components are also present: HCN, COS, CS_2, and, in the light oil, toluene, xylene, cumene, naphthalene, phenol, cresol, and a heavy oil (667°F).

The purpose of the raw gas quench is to condense and recover most of the slurry oil present in the gas. The oil vapor content of the cooled gas should be less than 2% before entering the shift conversion section of the plant. As much useful heat as possible is recovered from the gas during cooling and condensation. The process flow diagram for the section appears in Figure 8-17 and stream data are in Table 8-7.

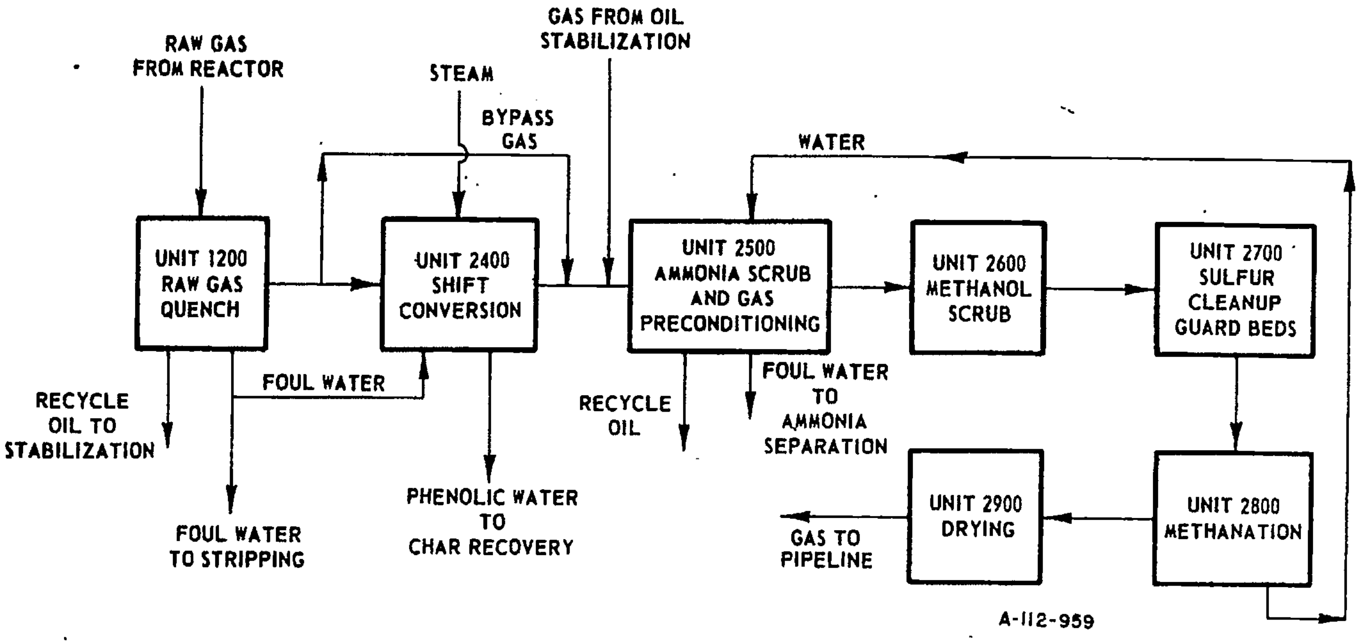


Figure 8-16. HYGAS DEMONSTRATION PLANT MAIN STREAM GAS-TREATMENT UNITS

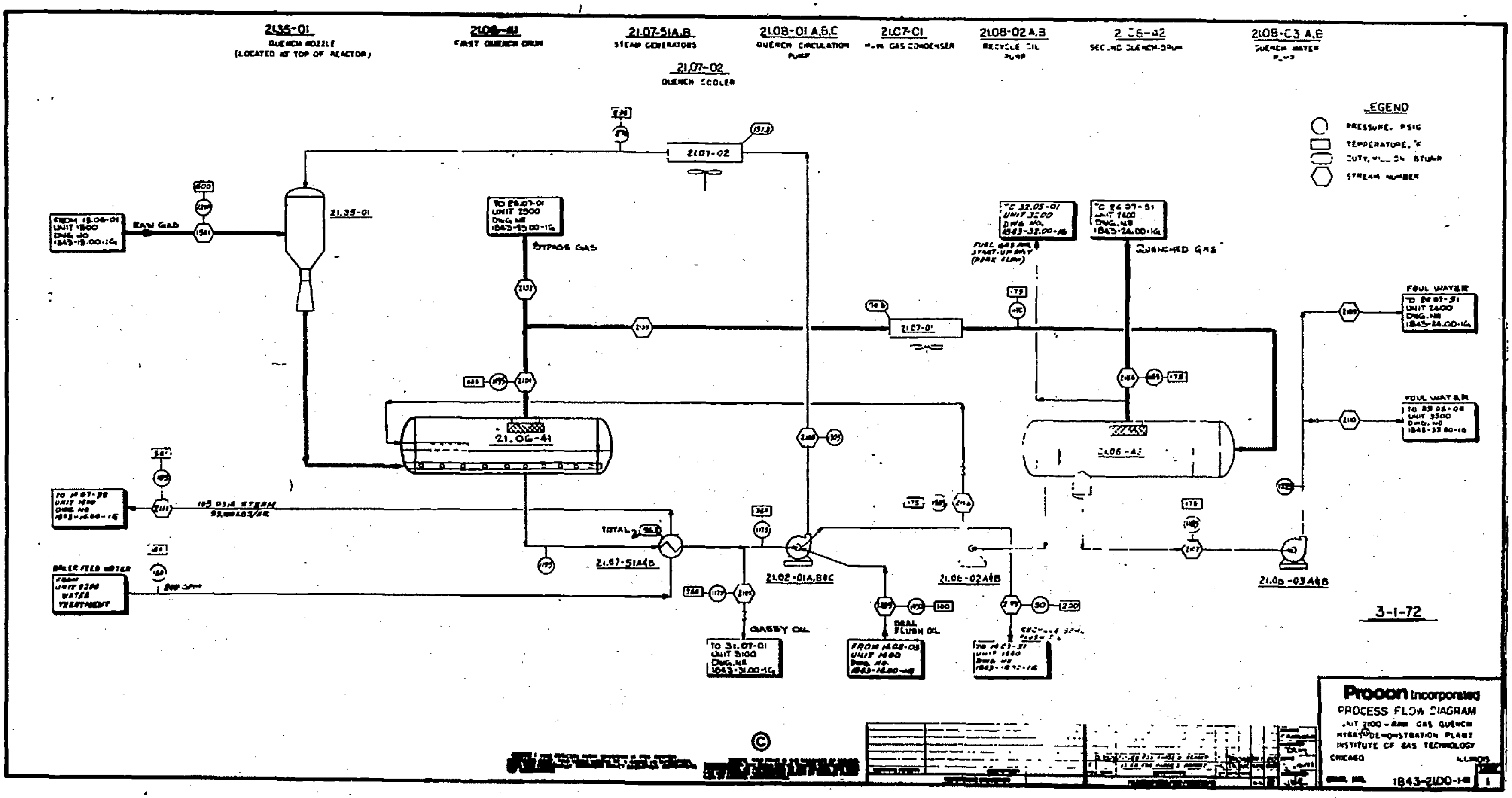


Figure 8-17. RAW GAS QUENCH SECTION PROCESS FLOW DIAGRAM FOR THE HYGAS DEMONSTRATION PLANT

Table 8-7. RAW GAS QUENCH SECTION STREAM CONDITIONS, COMPOSITIONS AND MATERIALS BALANCES FOR THE HYGAS DEMONSTRATION PLANT

Stream Description	Molecular Weight	Raw Gas From Generator	Quench to Nozzle	First Quench Flash Drum Gas	By-Pass Gas	Flash Gas to Second Quench Drum	Second Quench Drum Flash Gas	Gassy Oil to Unit 3100	Recycle Oil to First Quench Drum	Second Quench Drum Water	Water to Unit 2400	Water to Unit 3500	100 psig Steam to Unit 1400
Stream Number		1501	2108	2101	2102	2103	2104	2105	2106	2107	2109	2110	2111
Weight Flow, M lbs/hr		1355.89	4543.6	563.82	162.16	401.66	297.71	832.06	41.27	62.68	36.93	25.63	91.46
MM SCFD, dry basis		287.3	--	198.61	58.31	140.30	133.5	--	--	--	--	--	--
Density at Op. Conditions		4.572	--	2.37	2.37	2.37	3.15	--	--	--	--	--	--
Temperature, °F		600	400	400	400	400	175	350	175	175	175	175	341
Pressure, psia		1215	--	1133.7	1133.7	1209.7	1197.7	1189.7	1239.7	1239.7	1239.7	1239.7	119.7
Molecular Weight		42.8	134.6	21.80	21.80	21.80	20.2	134.6	89.21	19.42	19.42	19.42	18.0
Fuel Equivalent		--			--			204.8	--	--	--		
Component, mols/hr													
CO_2	44.01	3792.6	1073.4	3652.0	1095.6	2556.4	2349.4	197.6	57.0	150.0	88.7	61.3	
H_2S	34.08	332.7	190.1	326.0	97.8	228.2	173.0	24.9	18.2	37.0	21.9	15.1	
HCN	27.03	8.4	13.6	8.3	2.5	5.8	2.75	2.15	2.05	1.0	0.6	0.4	
COS	60.07	3.8	1.6	6.7	2.0	4.7	1.5	0.3	3.2	0.01	0.006	0.004	
CS_2	76.13	1.9	8.1	0.4	0.12	0.28	0.19	1.5	0.083	0.007	0.004	0.003	
NH_3	17.03	165.8	66.9	154.5	40.75	113.75	6.45	12.3	1.0	106.3	62.8	43.5	
H_2O	18.02	4500.0	2571.4	4030.0	1026.6	3003.4	90.1	473.4	3.4	2909.9	1720.1	1189.8	5081.1
H_2	2.02	6340.2	488.9	6251.3	1814.8	4436.5	4434.4	90.0	1.1	1.0	0.6	0.4	
CO	28.01	5620.5	706.4	5492.1	1604.9	3887.2	3883.6	130.0	1.6	2.0	1.2	0.8	
CH_4	16.04	5249.5	727.8	5117.2	1507.3	3609.9	3604.2	134.0	1.7	4.0	2.4	1.6	
C_2H_6	30.08	196.9	48.9	191.0	57.3	133.7	130.3	9.0	3.1	0.3	0.2	0.1	
C_6H_6 Benzene	78.12	1411.6	6654.1	466.7	140.0	326.7	59.4	1210.0	265.1	2.2	1.3	0.9	
C_7H_8 Toluene	92.13	331.5	1748.0	23.7	7.1	16.6	3.1	321.1	13.3	0.2	0.12	0.08	
C_8H_{10} O-Xylene	106.16	160.7	776.7	56.3	16.9	39.4	2.84	141.0	36.56	--	--	--	
C_9H_{12} Cumene	120.19	35.2	170.0	11.0	3.3	7.7	0.95	30.95	6.75	--	--	--	
$C_{10}H_8$ Naphthalene	128.16	8.1	43.5	0.2	0.06	0.14	0.047	8.0	0.1	--	--	--	
C_6H_5OH Phenol	94.1	70.1	348.0	10.45	3.15	7.3	0.0028	64.15	4.5	2.8	1.65	1.15	
C_7H_7OH Cresol	108.1	7.0	34.7	1.055	0.315	0.74	0.00028	6.4	0.5	0.24	0.18	0.06	
Light Oil (429°F)	144	2523.7	13599.2	65.7	19.7	46.0	2.84	2501.2	43.2	--	--	--	
Heavy Oil (677°F)	220	823.8	4474.2	0.3	0.09	0.21	0.019	823.7	0.2	--	--	--	
Total		31584.0	33745.5	25864.9	7440.3	18424.6	14745.1	6181.7	462.6	3217.0	1901.8	1315.2	5081.1

The raw gas exits from the gasifier at a flow rate of 287 million SCF/day and enters an ejector-type quench nozzle, where it is contacted by a 12,800 gpm flow of quench oil at 290°F. The quench nozzle is mounted high in the air near the reactor outlet so that any tars and dust present in the gas are rapidly cooled and flushed away, thus avoiding possible solids and tar buildup in the transfer line. The gas and liquid mixture enters the first separation vessel at 400°F. This temperature is above the water dew point and is high enough that some of the heat can be economically recovered. Quench oil withdrawn from the first separator is cooled by generating 105-psig steam. It then is cooled by an air cooler to 290°F before returning to the quench nozzle. Excess oil flows to the oil stabilization section of the plant.

One-quarter of the gas from this first separator is routed to bypass the downstream shift converter. The rest of the gas goes to the second-stage cooling system through an air cooler which lowers the temperature to 175°F. Because water condenses at temperatures below about 390°F, corrosion may be troublesome in this part of the plant. The cooled mixture is separated into oil, water, and gas phases. The oil is pumped into the first-stage separator; the gas and part of the water flow to shift conversion. The rest of the water is stripped to remove dissolved gas before flowing to wastewater treatment.

8.4.3.2 Shift Conversion

Gas entering the shift conversion section at 175°F and 1185 psig has the following composition:

	mol. %
CO	26.3
CO_2	16.0
H_2	30.1
H_2O	0.6
CH_4	24.5
C_2H_6	0.9
C_6H_6	0.4
NH_3	--
H_2S	1.2
Total	100.00

The purpose of shift conversion is to adjust the ratio of H_2/CO to about 3 for the catalytic methanation of the gas. The shift conversion reaction takes place over a catalyst in the temperature range of 600° to 800°F. The reaction

$$H_2O + CO \rightarrow H_2 + CO_2$$

proceeds to the right to form additional hydrogen from steam and carbon monoxide. The H_2/CO ratio at the shift converter outlet is about 7. To obtain the desired ratio of 3, shifted gas is blended with gas that bypasses the catalyst bed. A process flow diagram is presented in Figure 8-18, and stream data appear in Table 8-8.

The shift conversion catalyst chosen for this design is the BASF K8-11. It is a cobalt molybdenum catalyst on a spinel base especially developed for high-pressure systems. Because the catalyst is most active in the sulfided form, some sulfur must be present in the feed gas. The catalyst is used commercially with gas feeds containing up to 2% aromatic vapors without a loss of activity. The catalyst resists size degradation in case liquid water inadvertently contacts it at operating temperatures.

The shift conversion feed gas is saturated with moisture at 348 °F by bubbling it through part of the foul water withdrawn from the raw gas quench. Heat is supplied by exchange with the shift reactor effluent. The gas is further heated to 650 °F by additional exchange with reactor effluent, and its steam content is adjusted to about 0.7 moles of steam/mole of dry gas.

The shift conversion reactions take place over the catalyst contained in two fixed beds. The forebed is sacrificial and is used only to protect the main catalyst bed from fouling by gum formers or by solids carryover. If the forebed plugs, it can be bypassed. The afterbed contains enough catalyst to perform the desired shifting. The temperature of the afterbed gas effluent is 825 °F. After heat exchange and generation of 50-psi steam, condensed water is separated from the gas and is directed to the char recovery section. The gas flows on to gas preconditioning.

8.4.3.3 Ammonia Scrub and Gas Preconditioning

A process flow diagram is presented in Figure 8-19 and stream data appear in Table 8-9.

Gas from shift conversion is combined with bypass gas and with gas stripped from the recycled slurry oil. The combined gas is air cooled to 180 °F, the liquids are separated, and then the gas is further cooled to 120 °F by cooling water before a second separation step. Process water from the methanator is used to scrub ammonia from the gas at this point in the process. Then, the scrubbed gas is refrigerated to 70 °F for the final liquid separation. Chilled gas from this separator flows on to the acid-gas removal unit. Condensed oil is recycled to slurry preparation and condensed water is directed to the ammonia separator.

8.4.3.4 Acid-Gas Removal (Rectisol)

Feed gas flow rate to this unit is 220 million SCF/day. By scrubbing the gas with methanol at low temperature in three towers, the Rectisol process cleans the gas of all moisture and benzene and reduces the hydrogen sulfide content to 0.1 ppm and the carbon dioxide content to 0.15%. The feed gas has the following composition:

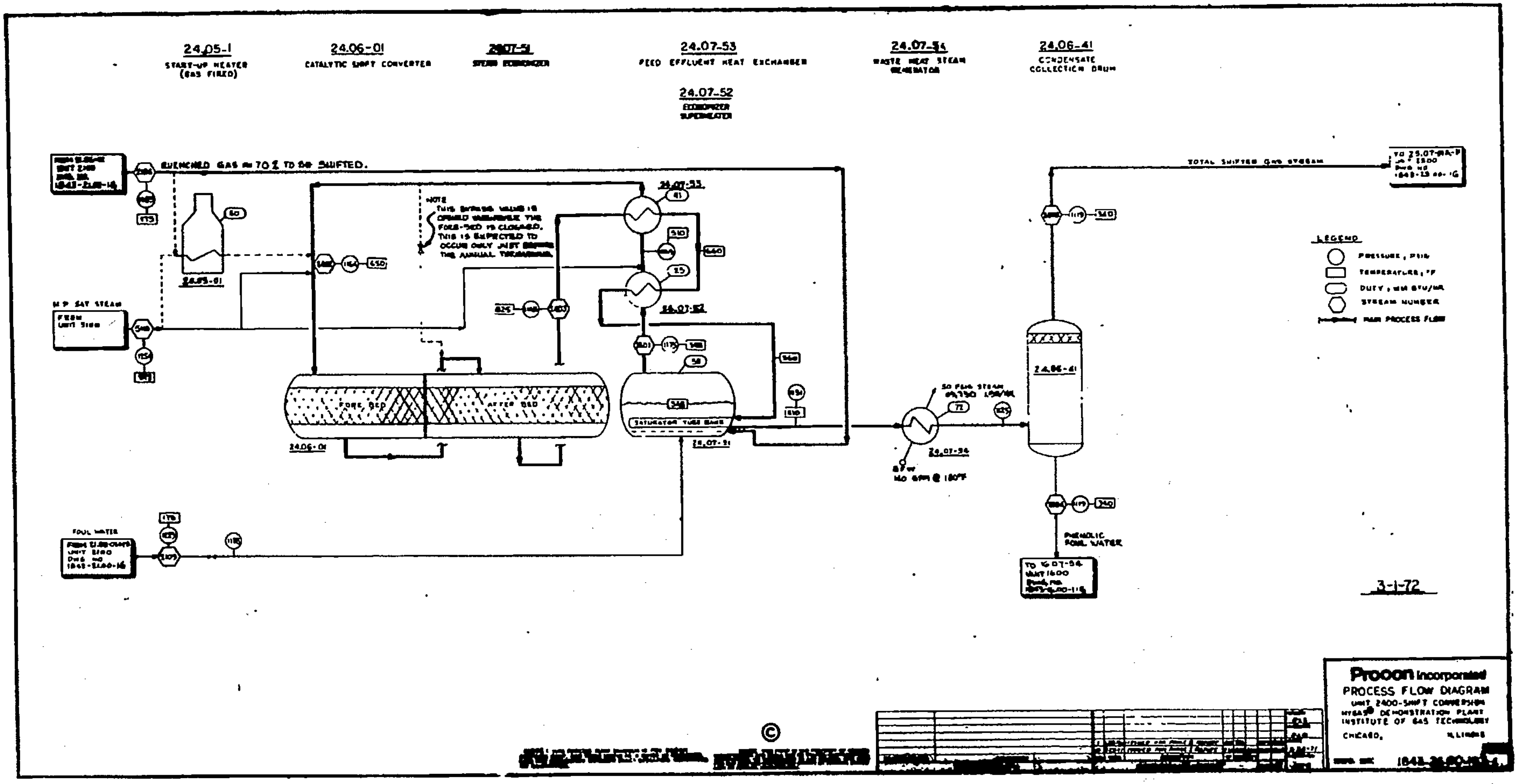


Figure 8-18. SHIFT CONVERSION SECTION PROCESS FLOW DIAGRAM FOR THE HYGAS DEMONSTRATION PLANT

Table 8-8. SHIFT CONVERSION SECTION STREAM CONDITIONS, COMPOSITIONS AND MATERIALS BALANCES FOR THE HYGAS DEMONSTRATION PLANT

Stream Description	Molecular Weight	Feed Gas From Unit 2100	Foul Water From Unit 2100	Economizer Preheated Effluent	Feed to Shift Converter	Product From Shift Converter	Foul Water to Waste Disposal	Shifted Gas to Ammonia Scrub	Steam Added
Stream Number		2104	2109	2401	2402	2403	2404	2405	5110
Weight Flow, M lbs/hr		297.7	36.93	334.6	481.2	481.2	92.9	388.3	146.5
MM SCFD, dry basis		133.5	--	135.1	135.1	160.9	--	160.8	--
Density at Op. Conditions		3.15	60.5	2.45	1.70	1.45	59.3	2.31	1.82
Temperature, °F		175	175	345	650	825	340	340	570
Pressure, psia	--	1197.7	1197.7	1189.7	1178.7	1162.7	1133.7	1133.7	1264.7
Molecular Weight		20.20	19.42	20.11	19.42	19.42	18.26	19.72	18.00
Fuel Equivalent									
Component, mols/hr									
CO_2	44.01	2349.4	88.7	2438.1	2438.1	5278.1	1.1	5277.0	--
H_2S	34.08	173.0	21.9	194.9	194.9	194.9	2.2	192.7	--
HCN	27.03	2.75	0.6	3.35	3.35	--	--	--	--
COS	60.07	1.5	0.006	1.5	1.5	1.5	0.5	1.0	--
CS_2	76.13	0.19	0.004	0.19	0.19	0.19	0.1	0.09	--
NH_3	17.03	6.45	62.8	69.25	69.25	72.6	3.6	69.0	--
H_2O	18.02	90.1	1720.1	1810.2	9951.4	7109.1	5076.1	2033.0	8141.2
H_2	2.02	4434.4	0.6	4435.0	4435.0	7273.9	0.01	7273.9	--
CO	28.01	3883.6	1.2	3884.8	3884.8	1048.2	0.2	1048.0	--
CH_4	16.04	3604.2	2.4	3606.6	3606.6	3606.6	0.1	3606.5	--
C_2H_6	30.08	130.3	0.2	130.5	130.5	130.5	1.1	129.4	--
C_6H_6 Benzene	78.12	59.4	1.3	60.7	60.7	62.35	0.8	61.0	--
C_7H_8 Toluene	92.13	3.1	0.12	3.32	3.32	3.32	0.5	2.82	--
C_8H_{10} O-Xylene	106.16	2.84	--	2.84	2.84	2.84	--	2.84	--
C_9H_{12} Cumene	120.19	0.95	--	0.95	0.95	0.95	--	0.95	--
$C_{10}H_9$ Naphthalene	128.16	0.047	--	0.047	0.047	0.047	0.02	0.027	--
C_6H_5OH Phenol	94.1	0.0028	1.65	1.55	1.65	0.58	0.58	--	--
C_7H_7OH Cresol	108.1	0.00028	0.18	0.18	0.18	0.18	0.05	0.13	--
Light Oil (429°F)	144	2.84	--	2.84	2.84	2.84	0.04	2.8	--
Heavy Oil (677°F)	200	0.019	--	0.019	0.019	0.019	--	0.019	--
Total		14745.1	1901.8	16646.9	24788.1	24788.8	5087.0	19701.2	8141.2

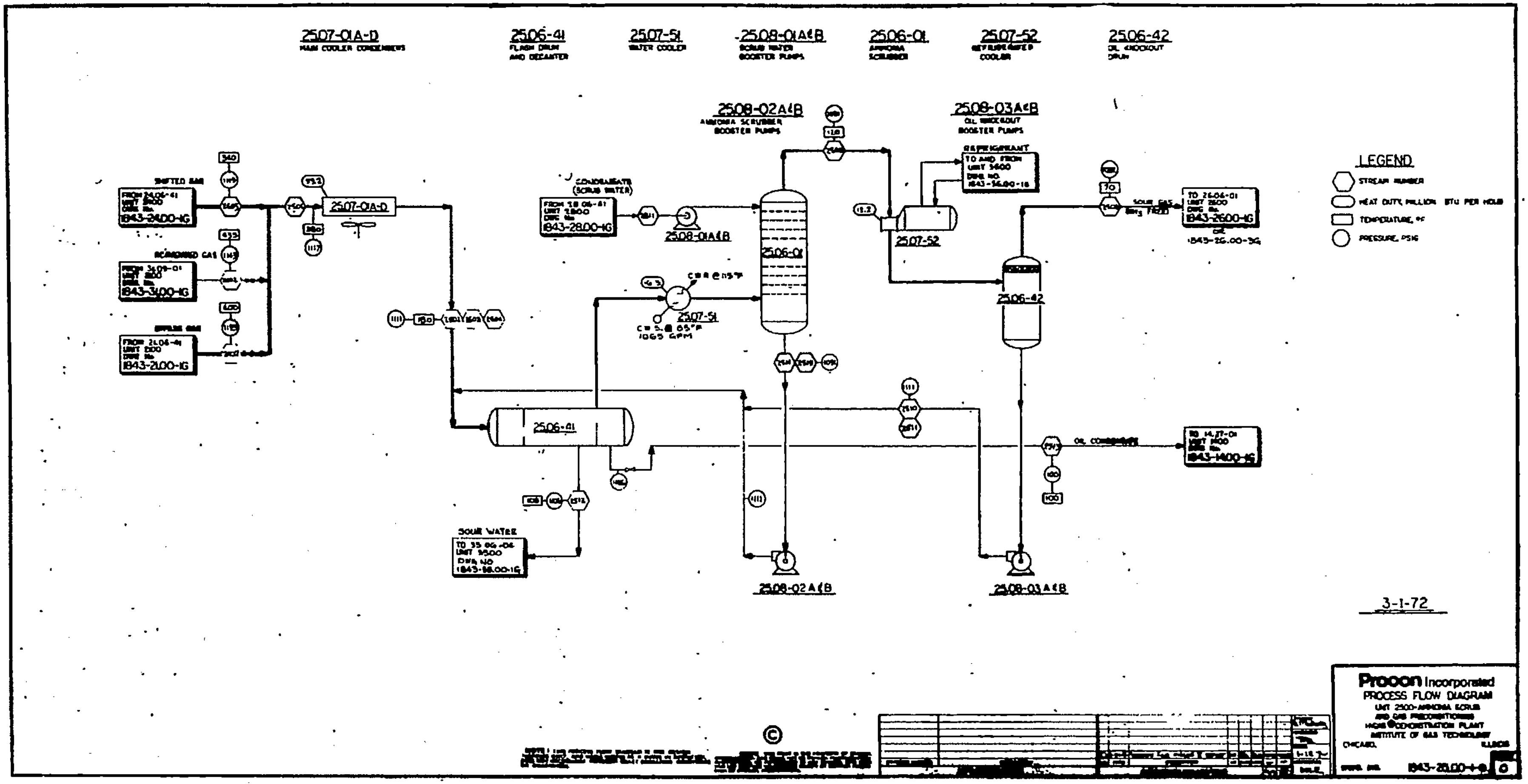


Figure 8-19. AMMONIA SCRUB AND GAS PRECONDITIONING SECTION PROCESS FLOW DIAGRAM FOR THE HYGAS DEMONSTRATION PLANT

Table 8-9. AMMONIA SCRUB AND GAS PRECONDITIONING SECTION STREAM CONDITIONS° COMPOSITIONS AND MATERIALS BALANCES FOR THE HYGAS DEMONSTRATION PLANT

Stream Description	Shifted Gas From Unit 2400	Bypass Gas From Unit 2100	Scavanged Gas From Unit 3100	Main Fred Stream to Unit	Vapor From Main Coolers	Aqueous Liquid From Main Coolers	Oily Liquid From Main Coolers	Scrubbed Hot Gas	Product Gas (NH_3 Free)	Aqueous Liquid From Oil Knockout Drum	Oily Liquid From Oil Knockout Drum	Sour Water to Unit 3500	Oil Condensate to Unit 1400	Aqueous Liquid From NH_3 Scrubber Bottoms	Oil Liquid From NH_3 Scrubber Bottoms	Scrub Water From Unit 2800
Stream Number	2405	2102	3102	2501	2502	2503	2504	2505	2506	2510	1511	2512	2513	2514	2515	2811
Weight Flow, M lbs/hr	388.3	162.16	16.1	566	506	55.5	3.98	492	484	1.24	6.99	113	18	56.6	7.26	49.44
MM SCFD, dry basis	161	58.3	5.4	244	224	--	--	221	220	--	--	--	--	--	--	--
Density at Op. Conditions	2.30	2.37	2.84	2.33	2.98	60.5	50.0	3.19	3.42	58.7	51.0	60.5	51.0	59.5	48.5	61.1
Temperature, °F	340	400	430	360	180	180	180	120	70	70	70	100	100	120	120	150
Pressure, psia	1133.7	1133.7	1133.7	1133.7	1125.7	1125.7	1125.7	1105.7	1096.7	1096.7	1096.7	1120.7	1120.7	1105.7	1105.7	1105.7
Molecular Weight	19.68	21.80	27.05	20.41	20.55	18.20	98.48	20.24	20.03	21.3	71.58	18.66	74.32	19.07	67.91	18.0
Fuel Equivalent																
Component, mols/hr																
CO_2	5277.0	1095.6	192.4	6565.0	6549.5	13.3	2.2	6445.0	6440.0	1.0	4.0	118.7	6.3	104.4	0.1	--
H_2S	192.7	97.8	22.0	312.5	302.6	9.0	0.9	284.4	280.3	2.6	1.5	27.2	5.0	15.6	2.6	--
HCN	--		0.1	2.6	0.5	2.1	--	--	--	--	--	2.6	--	0.5	--	--
COS	1.0	2.0	0.05	3.05	1.35	0.8	0.9	0.45	--	0.2	0.25	1.9	1.15	0.9	--	--
CS_2	0.9	0.12	0.18	0.23	0.12	0.08	0.03	0.02	--	0.2	--	0.2	0.03	0.1	--	--
NH_3	69.0	40.75	7.8	117.55	102.7	14.85	--	--	--	--	--	117.55	--	102.7	--	--
H_2O	2033.0	1026.6	3.2	3062.8	58.6	3033.3	0.9	58.5	5.6	52.9	--	5830.1	1.0	2773.9	0.1	2773.9
H_2	7273.9	1814.8	90.0	9178.7	9178.6	0.1	--	9177.4	9176.2	0.1	1.1	0.4	2.1	0.2	1.0	--
CO	1048.0	1604.9	130.0	2782.9	2781.9	0.5	0.5	2781.3	2781.0	--	0.3	0.7	1.2	0.2	0.4	--
CH_4	3606.5	1507.3	133.5	5247.3	5245.75	0.9	0.65	5220.7	5211.0	--	9.7	1.0	38.3	0.1	27.95	3.0
C_2H_6	129.4	57.3	8.9	195.6	194.87	0.1	0.63	192.1	190.0	0.05	2.05	0.2	5.4	0.05	2.72	--
C_6H_6 Benzene	61.0	140.0	4.6	205.6	190.9	0.5	14.2	136.0	66.0	0.1	69.9	1.0	138.6	0.4	54.5	--
C_7H_8 Toluene	2.82	7.1	0.3	10.22	7.44	0.3	2.48	5.2	4.0	0.1	1.1	0.6	5.62	0.2	2.04	--
C_8H_{10} Xylene	2.84	16.9	0.7	19.81	16.3	--	3.51	7.6	0.1	--	7.5	--	19.71	--	8.7	--
C_9H_{12} Cumene	0.95	3.3	0.014	4.26	3.43	--	0.83	1.4	0.1	--	1.3	--	4.16	--	2.03	--
$C_{10}H_8$ Napthalene	0.027	0.06	0.006	0.093	0.043	0.01	0.04	0.003	--	--	0.003	0.01	0.083	--	0.04	--
C_6H_5OH Phenol	--	3.15	0.002	3.15	2.68	0.27	--	0.75	0.05	0.7	--	2.15	0.95	1.58	0.35	--
C_7H_7OH Cresol	0.13	0.315	0.0002	0.447	0.247	0.15	0.05	0.05	--	--	0.01	0.347	0.05	0.217	0.02	--
Light Oil (429°F)	2.8	19.7	0.06	22.56	9.1	0.1	13.36	0.9	0.06	0.4	0.44	1.0	21.5	0.5	7.7	--
Heavy Oil (677 °F)	0.019	0.09	0.0001	0.11	--	--	0.11	--	--	--	--	--	0.11	--	--	--
Total	19701.2	7440.3	593.6	27734.5	24646.6	3046.4	41.5	24311.8	24154.5	58.2	99.1	6105.7	251.3	3001.1	109.7	2776.9

A7506 1665

	mol %
CO	11.5
CO_2	26.6
H_2	37.9
H_2O	Trace
CH_4	21.7
C_2H_6	0.8
C_6H_6	0.3
H_2S	1.2
Total	100.0

A process flow diagram is presented in Figure 8-20 and stream data appear in Table 8-10.

The gas first enters the benzene wash tower in which both oil vapor and moisture are removed by scrubbing with methanol. The gas is cooled by heat exchange and flows to the hydrogen sulfide wash tower where it is scrubbed by methanol containing CO_2. The final scrubbing takes place in the CO_2 wash tower. Two solvents are used in this tower. The greatest solvent flow is flashed methanol, which contains some CO_2. Methanol free of CO_2 is used at the top of the tower for the final wash. The CO_2 heat of absorption is rejected by intermediate stream refrigeration in the absorber. Part of the CO_2-rich methanol is used as solvent in the H_2S wash tower; the remainder flows to the stripping section. Cold, clean, product gas is heat exchanged with feed gas to reduce the plant's refrigeration requirements.

The Rectisol process is proprietary; therefore, only limited details are presented here. Discussion of the second part of this two-stage process appears in Section 8.4.4.2.

8.4.3.5 Sulfur Clean-Up Guard Beds

Gas from the Rectisol scrubbers is clean enough to feed directly to the methanation unit; however, as additional protection for the methanation catalyst, sulfur guard beds are used. These fixed beds are contained in two parallel vessels. The entering gas flows through one or the other of the beds. Sulfur compounds contained in the feed gas are removed to very low levels by chemical reaction with the bed material. A process flow diagram is presented in Figure 8-21 and stream data appear in Table 8-11.

Zinc oxide was chosen as the reagent for this design. It is effective in removing only H_2S at the 117°F temperature used. If applied from 500° to 700°F, its capacity for sulfur is increased from 3 to 17 weight percent and organic sulfur compounds as well as H_2S can be removed. When the zinc oxide is spent, it will be returned to the supplier for recovery.

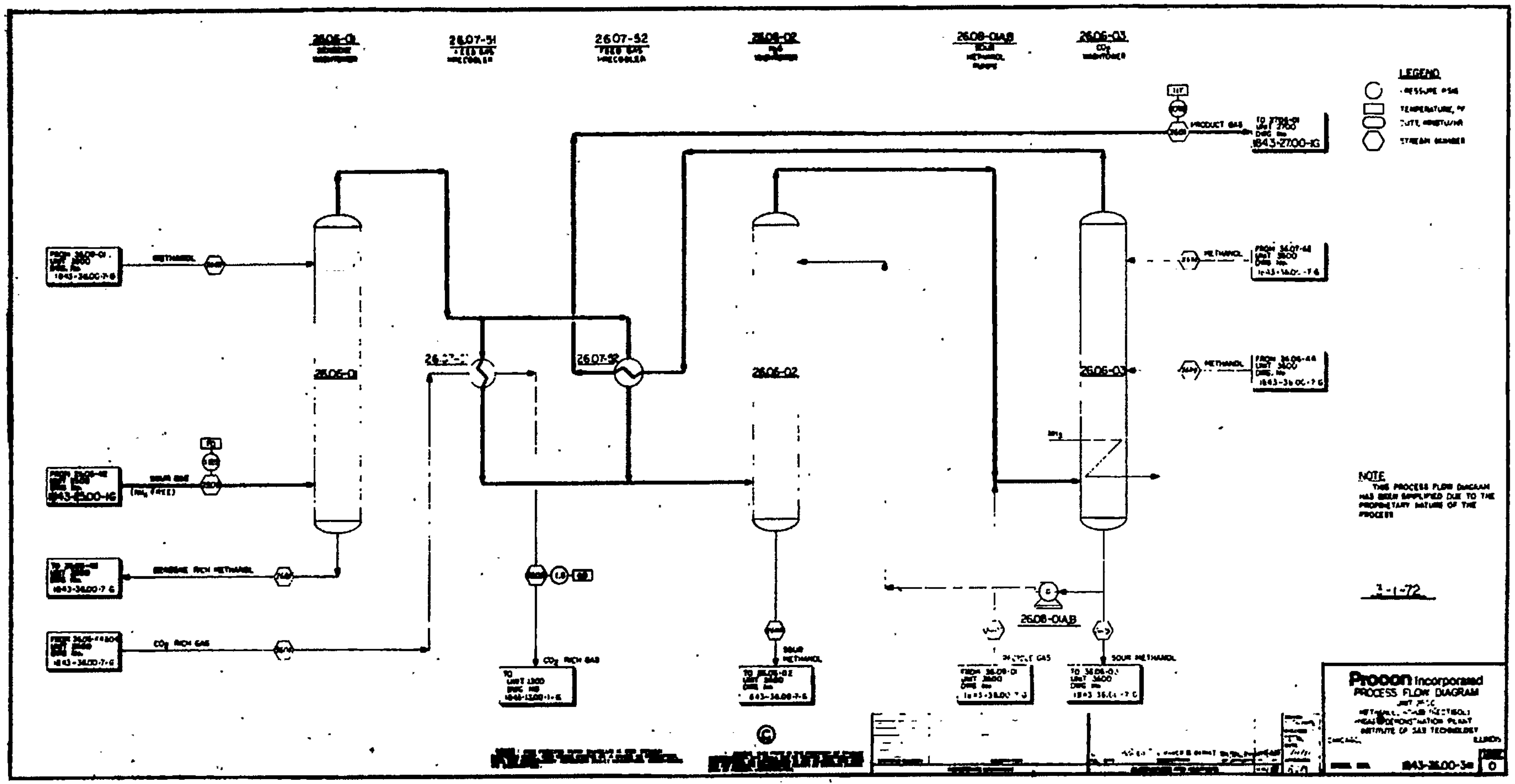


Figure 8-20. ACID-GAS REMOVAL SECTION PROCESS FLOW DIAGRAM FOR THE HYGAS DEMONSTRATION PLANT

Table 8-10. ACID-GAS REMOVAL STREAM DATA FOR THE HYGAS DEMONSTRATION PLANT

Stream Description	Molecular Weight	Sour Gas	Sweetened Product Gas	CO_2-Rich Gas
Stream Number		2506	2601	2602
Weight Flow, M lbs/hr		484	182	261
MM SCFD, dry basis		220	156	55
Density at Op. Conditions		3.42	1.82	0.13
Temperature, °F		70	117	60
Pressure, psia		1096	1061	16.5
Molecular Weight		20.0	10.6	43.2
Component, mols/hr				
CO_2	44.01	6440.0	25.6	5810.0
H_2S	34.08	280.3	0.1 ppm	5 ppm
H_2O	18.02	5.6	--	--
H_2	2.02	9176.2	9123.4	52.5
CO	28.01	2781.0	2776.8	4.0
CH_4	16.04	5211.0	5195.4	13.3
C_2H_6	30.08	190.0	32.4	154.2
C_6H_6	78.12	66.0	--	--
C_7H_8+	94.06	4.4	--	--
CH_3OH	32.04	--	0.6	1.7
Total		24154.5	17154.2	6035.7

A7506 1666

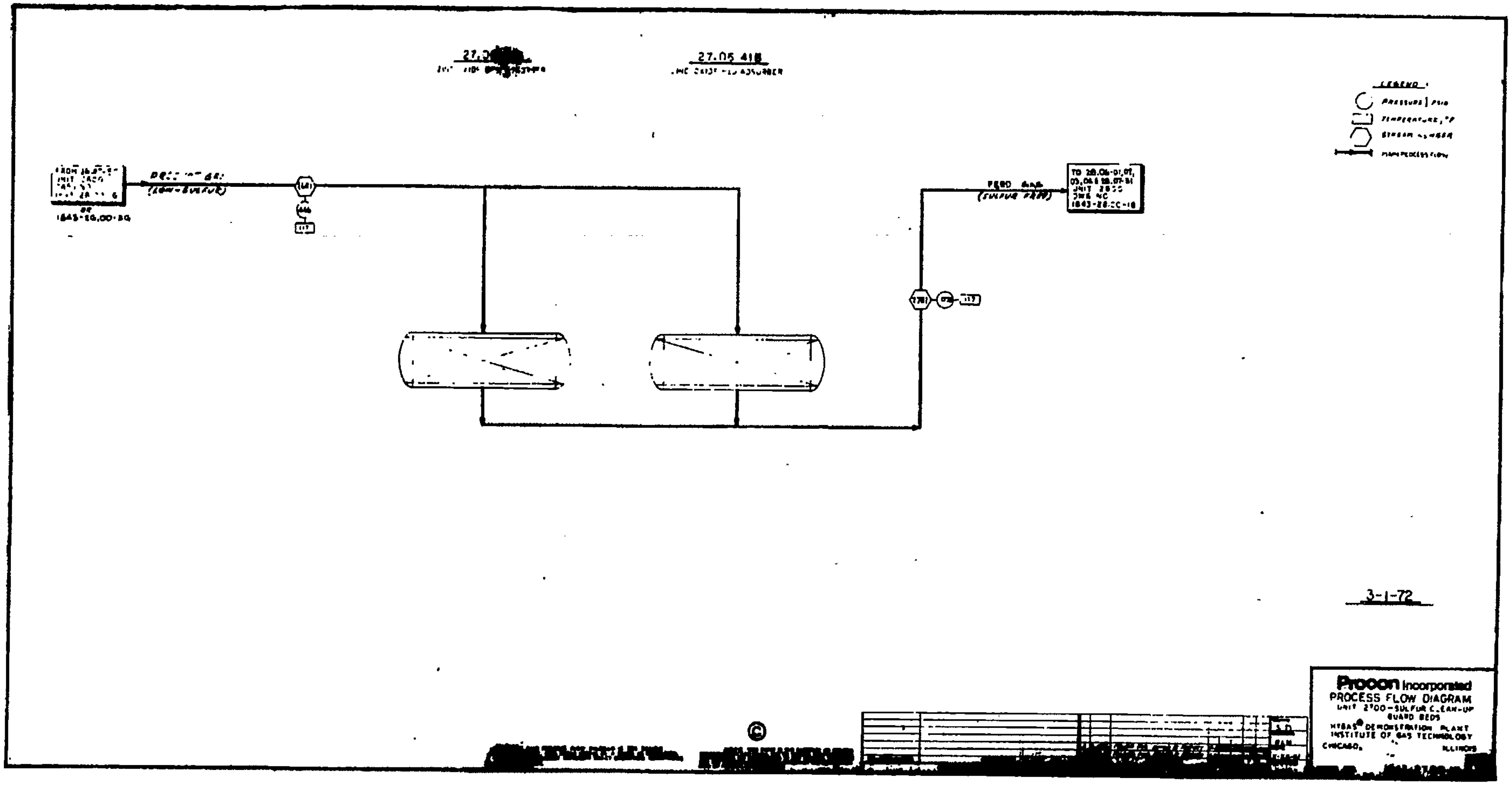


Figure 8-21. SULFUR CLEAN-UP GUARD BEDS SECTION PROCESS FLOW DIAGRAM FOR THE HYGAS DEMONSTRATION PLANT

Table 8-11. SULFUR CLEAN-UP GUARD BEDS SECTION STREAM DATA FOR THE HYGAS DEMONSTRATION PLANT

Stream Description	Feed Stream	Product Stream
Stream Number	2601	2701
Weight Flow, M lbs/hr	182.0	182.0
MM SCFD, dry basis	156	156
Density at Op. Conditions	1.82	1.81
Temperature, °F	117	117
Pressure, psia	1059.7	1053
Molecular Weight	10.6	10.6
Component, mols/hr		
COS	25.6	25.6
H_2S	0.0017	0.00034
H_2O	0	0
CO	2776.8	2776.8
H_2	9123.4	9123.4
CH_4	5195.4	5195.4
C_2H_6	32.4	32.4
CH_3OH	0.6	0.6
Total	17154.2	17154.2

A7506 1667

Metal-oxide-promoted activated carbon is a practical alternative reagent. It is effective n H_2S removal only and may not be satisfactory when much COS is present. When the promoted carbon is spent, it can be regenerated in place using air and steam. During regeneration, the sulfur is converted to the elemental form and is absorbed by the carbon. After several cycles, elemental sulfur present interferes with the adsorption of sulfur compounds from the gas, and the carbon must be replaced.

8.4.3.6 Methanation

Cleaned gas enters the methanation unit at a rate of 156 million SCF/day with the following composition:

	mol %
CO	16.2
H_2	53.2
CH_4	30.4
C_2H_6	0.2
Total	100.0

The methanation unit converts CO and H_2 to methane by the following reaction:

$$CO + 3H_2 \rightarrow CH_4 + H_2O$$

Ethane in the feed gas is hydrogenated to methane. The 530 Btu/SCF heating value of the feed gas is increased to 954 Btu/SCF in the dry gas leaving the methanator. A process flow diagram is presented in Figure 8-22 and stream data appear in Table 8-12.

The methanation reaction takes place over a nickel catalyst contained in four fixed beds. The reaction temperatures range between 550° and 900°F. The reaction is highly exothermic, generating 94,860 Btu/mole of CO reacted. The main problem in the design of this unit is to perform methanation to the desired extent and remove heat usefully from the process so that the reaction temperatures will not exceed 900°F. Higher temperatures may reduce catalyst activity and may also result in carbon formation from the gas.

In this design, the heat of reaction is removed as sensible heat in the reactor effluent gas. By limiting the CO content of the reactor inlet gas to about 4%, the outlet temperature will not exceed 900°F when the inlet temperature is 550°F. The CO content of the feed is adjusted by product gas recycle. The amount of recycle is reduced by increasing the number of methanation reactors. The volume of catalyst used corresponds to a space velocity of 8000 hr^{-1}. The catalyst is contained in four reactors in the ratio, from first to last of 1·2·4:8. The inlet temperature to each bed is 550°F and the outlet temperature is 900°F.

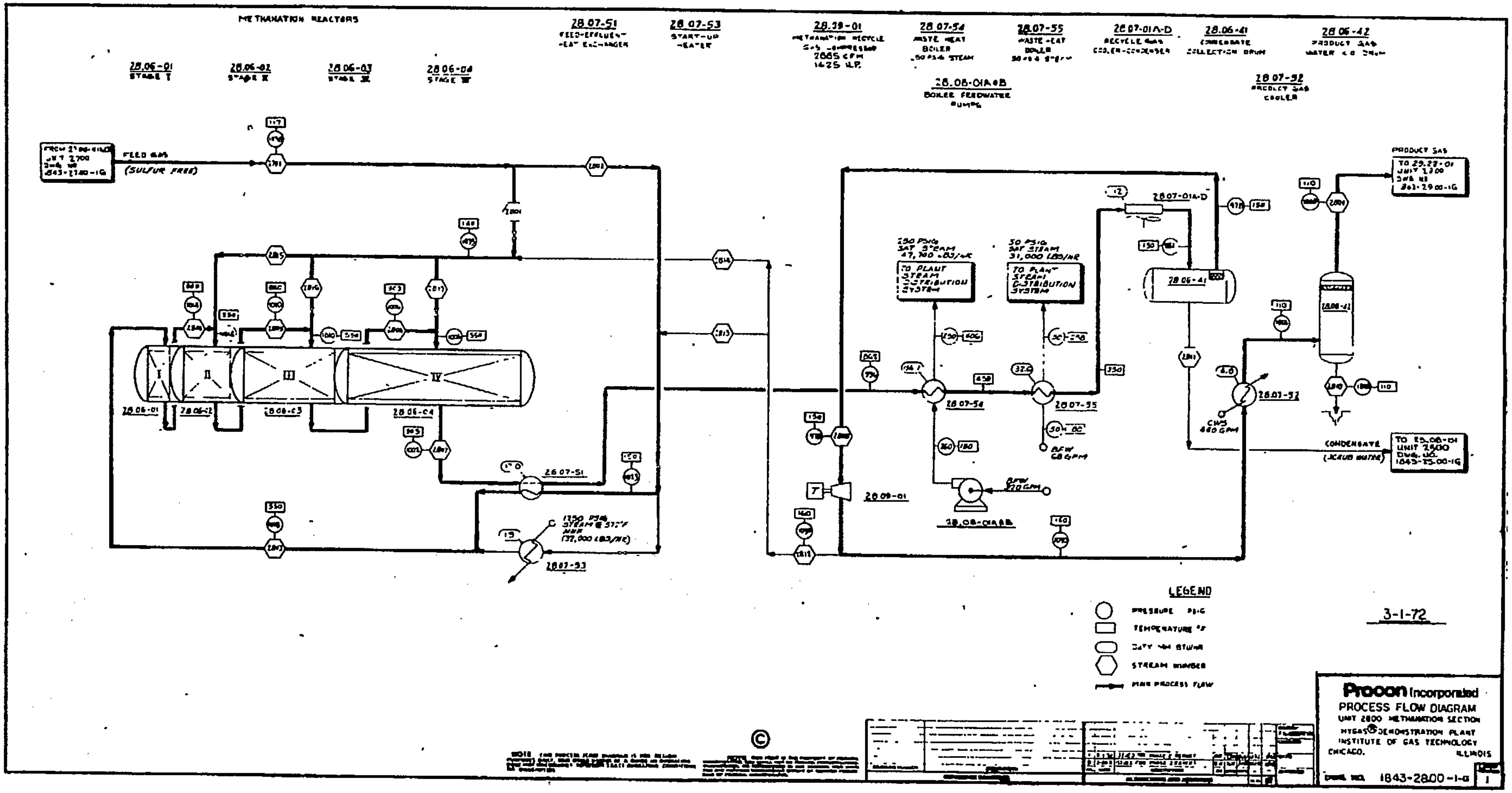


Figure 8-22. METHANATION SECTION PROCESS FLOW DIAGRAM FOR THE HYGAS DEMONSTRATION PLANT

Table 8-12. METHANATION SECTION STREAM CONDITIONS, COMPOSITIONS AND MATERIALS BALANCES FOR THE HYGAS DEMONSTRATION PLANT

Stream Description	Total Fresh Feed Gas From Unit 2700	Fresh Feed Gas to Stages II, III, and IV	Fresh Feed Gas to Stage I	Total Feed Gas to Stage I	Stage I Effluent Gas	Stage II Effluent Gas	Stage III Effluent Gas	Stage IV Effluent Gas	Recycle Compressor Feed Gas	Product Gas to Dryer (Unit 2900)	Water Effluent	Water Effluent	Total Recycle Gas
Stream Number	2701	2801	2802	2803	2804	2805	2806	2807	2808	2809	2810	2811	2812
Weight Flow, M lbs/hr	182	169.5	12.1	61.2	61.2	121.6	242.4	484.1	434.0	131.2	0.4	50.0	302.4
MM SCFD, dry basis	156.0	145.6	10.4	40.1	34.9	67.5	132.6	262.9	262.9	79.7	0.8 gpm	102 gpm	183.2
Density at Op. Conditions	1.81	1.78	1.76	2.18	1.06	1.06	1.05	1.05	2.27	2.57	61.86	61.19	2.38
Temperature, °F	117	117	117	150	899	901	900	900	150	110	110	150	160
Pressure, psia	1053	1050	1038	1033	1029	1025	1021	1017	993	1055	1055	993	1065
Molecular Weight	10.58	10.58	10.58	13.83	15.12	15.18	15.21	15.22	14.96	14.95	18.02	18.02	14.96
Fuel Equivalent	8227.1									8224.7			
Component, mols/hr													
C_2H_6	32.4	30.2	2.2	2.2	0.0	0.0	0.0	0.0	0.0	0.0			0.0
CH_3OH	0.6	0.6	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0			0.0
CO_2	25.6	23.9	1.7	5.4	1.3	5.4	14.4	32.5	32.5	9.5			23.0
CO	2776.8	2591.7	185.1	185.7	0.4	1.0	2.3	4.8	4.8	1.4			3.4
H_2	9123.4	8515.2	608.2	869.7	295.2	582.0	1159.2	2313.3	2313.3	699.8			1613.5
H_2O	0.0	0.0	0.0	12.1	205.6	588.9	1353.6	2883.0	109.1	10.5	23.8	2773.9	74.8
CH_4	5195.4	4849.0	346.4	3346.4	3540.2	6828.0	13402.7	26552.3	26549.3	8049.3		3.0	18500.0
Total	17154.2	16010.6	1143.6	4421.5	4042.7	8005.3	15932.2	31785.9	29009.0	8770.5	23.8	2776.9	20214.7

A7506 1668

Table 8-12, Cont. METHANATION SECTION STREAM CONDITIONS, COMPOSITIONS AND MATERIALS BALANCES FOR THE HYGAS DEMONSTRATION PLANT

Stream Description	Recycle Gas to Stage I	Recycle Gas to Stages II, III, and IV	Recycle Gas Plus Fresh Feed to Stage II	Recycle Gas Plus Fresh Feed to Stage III	Recycle Gas Plus Frech Feed to Stage IV
Stream Number	2813	2814	2815	2816	2817
Weight Flow, M lbs/hr	49.0	253.4	60.4	120.8	241.7
MM SCFD, dry basis	29.7	153.5	42.7	85.5	170.9
Density at Op. Conditions	2.35	2.35	2.04	2.04	2.04
Temperature, °F	160	160	141	141	141
Pressure, psia	1050	1050	1035	1035	1035
Molecular Weight	14.96	14.96	12.83	12.83	12.83
Fuel Equivalent					
Component, mols/hr					
C_2H_6	0.0	0.0	4.3	8.6	17.3
CH_3OH	0.0	0.0	0.1	0.2	0.3
CO_2	3.7	19.3	6.2	12.3	24.7
CO	0.6	2.8	370.6	741.3	1482.6
H_2	261.5	1352.0	1409.6	2819.2	5638.4
H_2O	12.1	62.7	9.0	17.9	35.8
CH_4	3000.0	15500.0	2907.0	5814.0	11628.0
Total	3277.9	16936.8	4706.8	9413.5	18827.1

Feed gas entering the first bed represents 6.7% of the total. This is blended with recycle gas to 4.2% CO content, preheated to 550°F, and fed to the first reactor. The rest of the feed gas is mixed with recycle to obtain a 7.9% CO content. This mixture is then divided in the ratio 1:2:4 and fed to reactors 2, 3, and 4. Before entering each of these reactors, 900°F product gas from the preceding reactor joins the feed, preheating it to 550°F and reducing the CO content to 4.2%. Effluent from the fourth reactor heats feed gas flowing to the first reactor, and then generates 250 psig and 50 psig steam before being cooled to 150°F. Separated gas is compressed for recycle to the methanation reactors and to boost the final product gas pressure. Separated water is used to scrub ammonia from gas in the ammonia scrub section.

8.4.3.7 Product Gas Drying

The product gas is dried to a moisture content of less than 7 lb/million SCF by scrubbing with 13 gpm of triethylene glycol. The dried gas dew point is −40°F. A conventional glycol absorption-stripping plant is used. The 80 million SCF/day of product gas has a temperature of 115°F and a 1035-psig pressure, and the following composition:

CH_4	91.9 mol %
H_2	8.0 mol %
CO	Trace (160 ppm)
H_2O	Trace (120 ppm)
CO_2	0.1 mol %
Total	100.0

Its heating value is 954 Btu/SCF, and it is completely interchangeable with natural gas. A process flow diagram is presented in Figure 8-23 and detailed stream data appear in Table 8-13.

8.4.4 Process Section Environmental Control Units

These units handle and dispose of side streams from the main process flow. They are shown in Figure 8-24. Potential pollutants that are recovered or treated in these units include dissolved gases, oil, sulfur compounds, ammonia, phenol, and hexavalent chrome salts.

8.4.4.1 Oil Stabilization

The purpose of this unit is to stabilize the gassy oil recovered in the raw gas quench before the oil is recycled to char feed preparation. Oil feed is flashed in three depressurizing stages. The atmospheric-pressure separating vessel is maintained at 150°F by steam heating. The stabilized oil is pumped to storage and from there it is recycled to char feed preparation. The net oil make is withdrawn at a rate of about 1000 bbl/day. The flashed gases from each stage are compressed to the preceding higher pressure stage. Gas compressed from the highest pressure flash drum reenters the main

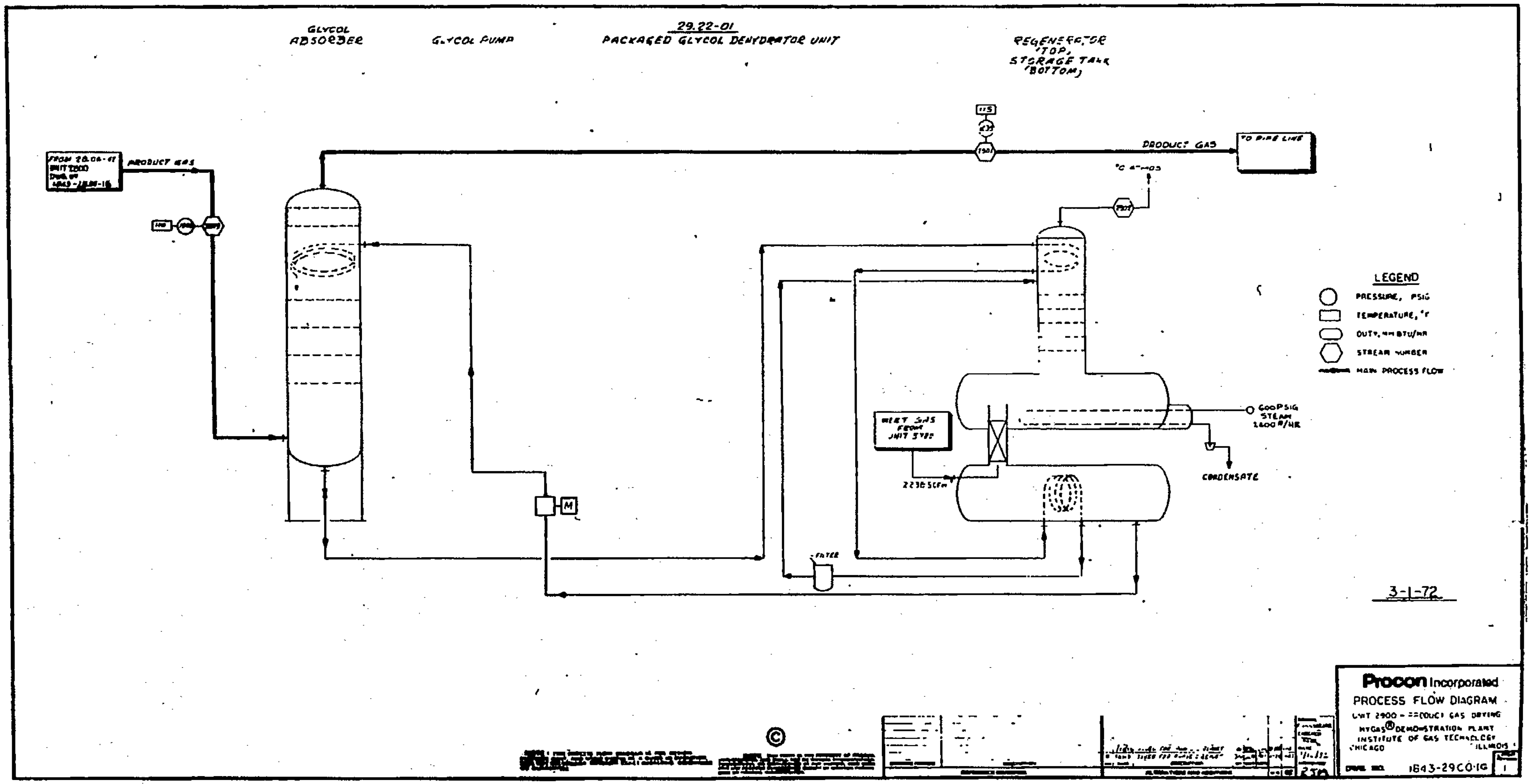


Figure 8-23. PRODUCT GAS DRYING SECTION PROCESS FLOW DIAGRAM FOR THE HYGAS DEMONSTRATION PLANT

Table 8-13. PRODUCT GAS DRYING STREAM DATA FOR THE HYGAS DEMONSTRATION PLANT

Stream Description	Molecular Weight	Wet Product Gas	Dry Product Gas	Water
Stream Number		2809	2901	2902
Weight Flow, M lbs/hr		131.2	131.2	--
MM SCFD, dry basis		79.7	79.7	--
Density at Op. Conditions		2.57	2.57	--
Temperature, °F		110	115	--
Pressure, psia		1055	1050	--
Molecular Weight		14.95	14.95	--
Fuel Equivalent		8225	8225	--
Component, mols/hr				
CO_2	44.01	9.5	9.5	--
H_2O	18.02	10.5	1.0	9.5
H_2	2.02	699.8	699.8	--
CO	28.01	1.4	1.4	--
CH_4	16.04	8049.3	8049.3	--
Total		8770.5	8761.0	9.5

A7506 1669

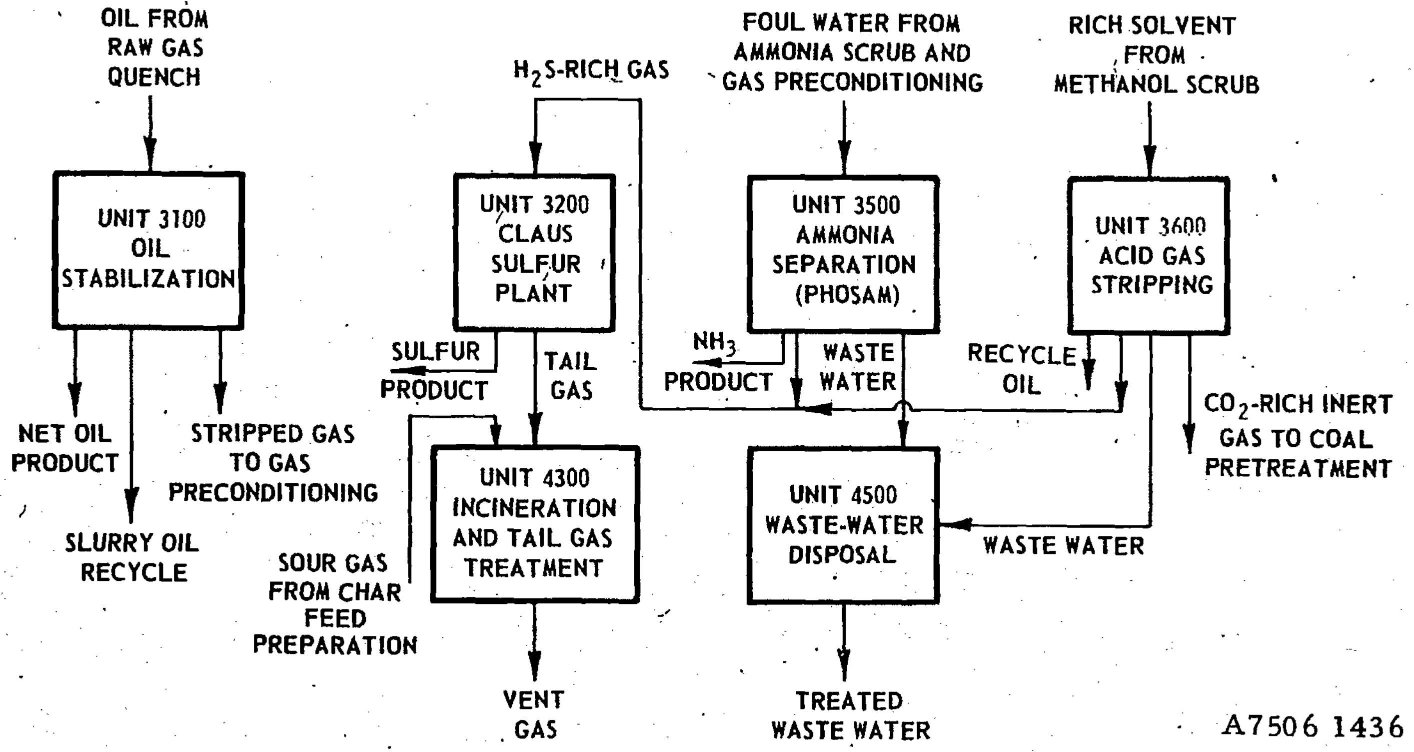


Figure 8-24. HYGAS DEMONSTRATION PLANT PROCESS SECTION ENVIRONMENTAL CONTROL UNITS

flow stream after shift conversion and before the ammonia scrub section. Some dissolved water separates from the oil in the first flash separator. It is sent to the ammonia separation process unit. A process flow diagram is presented in Figure 8-25 and stream data appear in Table 8-14.

8.4.4.2 Acid-Gas Stripping (Rectisol)

This is the stripping section of the Rectisol unit. Three methanol feed streams enter this part of the plant: One contains benzene and water as the main impurities, another contains dissolved H_2S and CO_2, and the third contains CO_2. Net products from the stripping unit include a CO_2-rich gas containing less than 5 ppm H_2S, an H_2S-rich stream containing more than 25% H_2S, and a liquid benzene stream. This unit contains six separation columns, a recycle compressor, plus auxiliary equipment. The flow scheme is complex and a detailed description will not be given; however, a process flow diagram is presented in Figure 8-26 and stream data appear in Table 8-15.

Methanol-containing benzene and water is diluted with additional water to separate the dissolved benzene. Then the methanol-water mixture is distilled for recovery of the methanol. Methanol from the H_2S and the CO_2 absorbers is stripped in separate towers to remove dissolved methane, which is recompressed to the main gas flow system. The CO_2 content of the H_2S-rich methanol is reduced by stripping at lower pressure, and then H_2S-rich gas is recovered by heating the solvent. The CO_2-rich methanol is stripped by flashing.

The Rectisol process is effective in cleaning coal gasification gases, and the quality of the recovered acid-gas stream is good. Undesirable process features include the complex flow scheme, the need for refrigeration, and the loss of about one-half of the ethane in the feed gas which is finally produced in the CO_2-rich gas. The 2.5% ethane content of this gas is high enough that the stream cannot be vented directly to the atmosphere. The CO_2-rich gas is used as inert gas in the coal pretreatment step before it is finally burned in the boiler. The H_2S-rich gas flows to the Claus process, and the recovered benzene to char feed preparation. A very small net water flow is produced which passes to waste water disposal.

8.4.4.3 Ammonia Separation (Phosam)

The ammonia separation unit receives process water from the ammonia scrub section, the oil stabilization section, and the raw gas quench section of the plant. The water is rich in dissolved ammonia, hydrogen sulfide, and carbon dioxide. This plant strips the dissolved gas from the water and recovers an ammonia product. A process flow diagram is presented in Figure 8-27 and stream data appear in Table 8-16.

Steam-stripping is used to remove the dissolved gases from the process water feed. Major components of the stripped gas are NH_3, H_2S, CO_2, and H_2O with a small amount of light hydrocarbons. Ammonia is selectively removed from this gas by the Phosam process, a development of the United States Steel Engineers and Consultants, Inc., a subsidiary of U.S. Steel Corp.

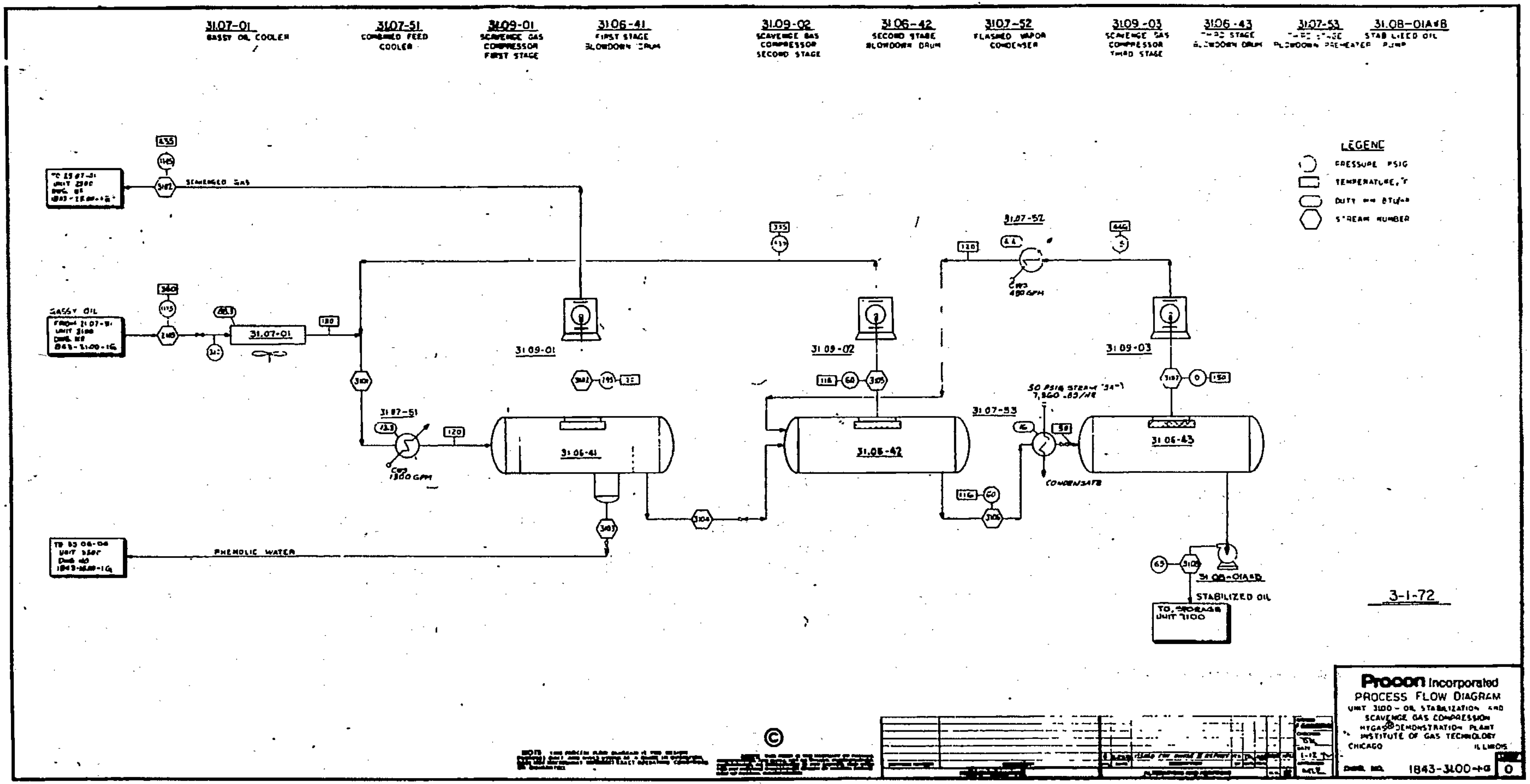


Figure 8-25. OIL STABILIZATION SECTION PROCESS FLOW DIAGRAM FOR THE HYGAS DEMONSTRATION PLANT

Table 8-14. OIL STABILIZATION SECTION STREAM CONDITIONS, COMPOSITIONS AND MATERIALS BALANCES FOR THE HYGAS DEMONSTRATION PLANT

Stream Description	Molecular Weight	Gassy Oil	Combined Feed to First Stage Blowdown Drum	1st Stage Blowdown Drum Flash Gas	1st Stage Blowdown Drum Water	1st Stage Blowdown Drum Liquid	2nd Stage Blowdown Drum Flash Gas	2nd Stage Blowdown Drum Liquid	3rd Stage Blowdown Drum Flash Gas	Stabilized Oil
Stream Number		2105	3101	3102	3103	3104	3105	3106	3107	3108
Weight Flow, M lbs/hr		836	859	16.1	7.25	836	22.7	826	12.8	813
MM SCFD, dry basis		--	--	5.4	--	--	5.08	--	1.82	--
Density at Op. Conditions		50.0	50.0	3.27	61.5	50.0	0.47	50.0	0.1	50.0
Temperature, °F		350	160	435	120	120	116	116	150	150
Pressure, psia		1189.7	334.7	1159.7	309.7	309.7	74.7	74.7	14.7	79.7
Molecular Weight		134.6	126.4	27.1	18.4	143.9	38.8	149.9	44.4	155.7
Fuel Equivalent		204.8	--	204.1	--	--	--	--	--	0.7
Component, mols/hr										
CO_2	44.01	197.6	495.1	192.4	0.8	301.9	297.5	51.3	46.9	4.4
H_2S	34.08	35.0	108.6	22.0	0.7	85.9	73.6	51.7	39.4	12.3
HCN	27.03	2.5	2.6	0.1	0.3	2.5	0.1	2.17	0.07	2.1
COS	60.07	0.3	0.35	0.05	0.0001	0.3	0.05	0.31	0.07	0.25
CS_2	76.13	1.5	1.518	0.018	0.02	1.5	0.018	1.49	0.03	1.46
NH_3	17.03	12.3	33.0	7.8	2.5	22.7	20.7	9.9	7.9	2.0
H_2O	18.02	473.4	500.5	3.2	389.3	108.0	27.1	167.6	86.7	80.9
H_2	2.02	90.0	97.1	90.0	0.007	7.1	7.1	0.191	0.19	0.001
CO	28.01	130.0	152.9	130.0	0.003	22.9	22.9	1.21	1.2	0.01
CH_4	16.04	134.0	203.8	133.5	0.002	70.3	69.8	13.1	12.6	0.5
C_2H_6	30.08	9.0	25.7	8.9	0.002	16.8	16.7	1.1	1.0	0.1
C_6H_6 Benzene	78.12	1225.0	1265.5	4.6	0.5	1260.4	40.5	1293.7	73.8	1219.9
C_7H_8 Toluene	92.13	321.8	330.8	0.9	0.2	329.7	9.0	336.7	16.0	320.7
C_8H_{10} O-Xylene	106.16	143.0	143.07	0.07	0.001	143.0	0.07	144.2	1.2	143.0
C_9H_{12} Cumene	120.19	31.3	31.3	0.014	0.001	31.3	0.014	31.5	0.2	31.3
$C_{10}H_8$ Naphthalene	128.16	8.0	8.0	0.006	0.01	8.0	0.0006	8.01	0.01	8.0
C_6H_5OH Phenol	94.1	64.05	64.05	0.002	0.7	63.3	0.002	63.3	0.003	63.3
C_7H_8OH 'Cresol	108.1	6.4	6.4	0.0002	0.07	6.3	0.0002	6.3	0.0003	6.3
Light Oil (429°F)	144	2503.6	2503.6	0.06	0.0001	2503.6	0.06	2503.6	0.06	2503.6
Heavy Oil (677°F)	226	823.7	823.7	0.0001	--	823.7	0.00001	823.7	0.002	823.7
Total		6212.5	6797.7	593.6	395.1	5809.2	585.2	5511.1	287.3	5223.8

A7506 1670

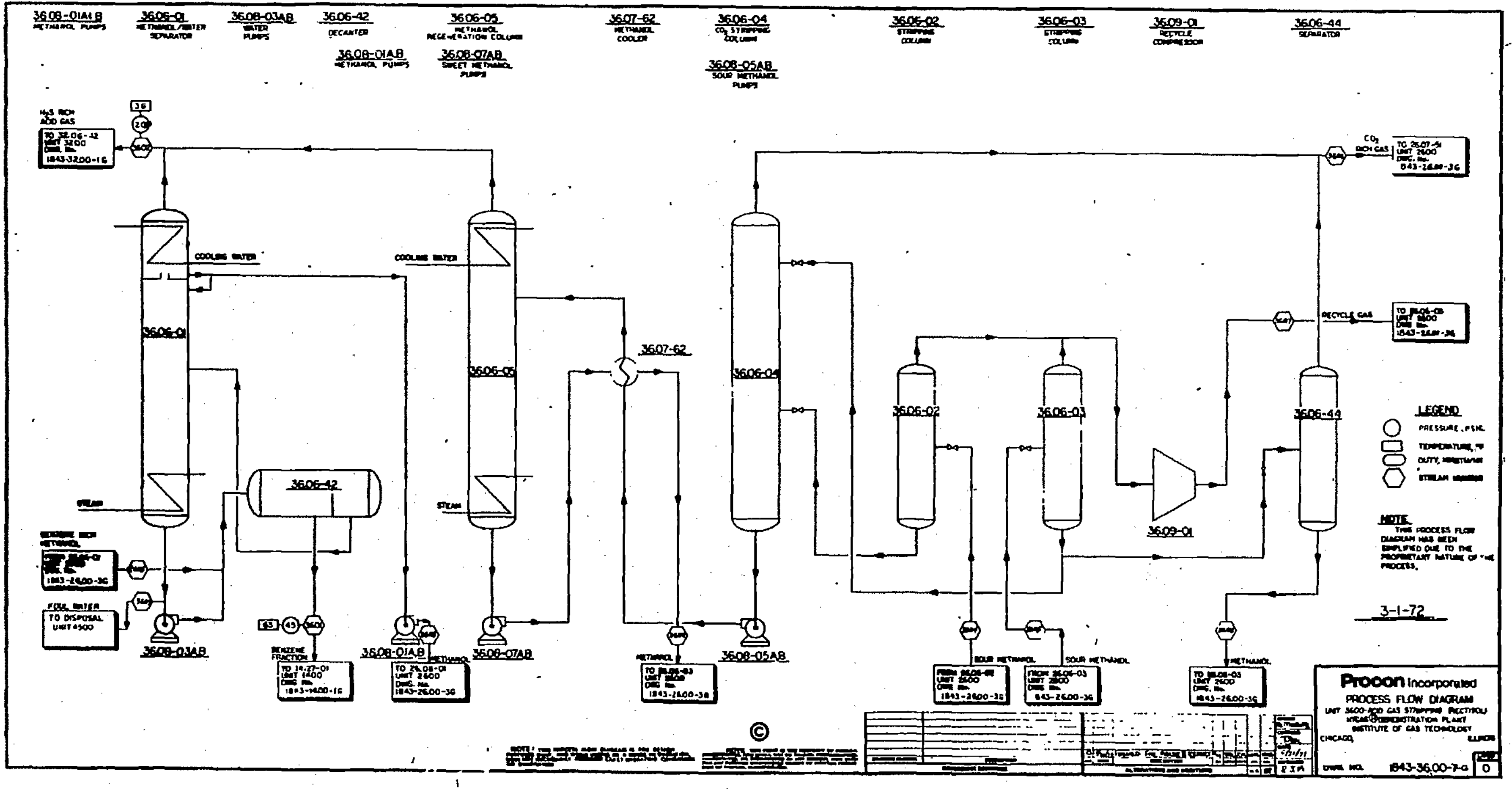


Figure 8-26. ACID-GAS STRIPPING SECTION PROCESS FLOW DIAGRAM FOR THE HYGAS DEMONSTRATION PLANT

Table 8-15. ACID-GAS STRIPPING SECTION STREAM DATA FOR THE HYGAS DEMONSTRATION PLANT

Stream Description	Molecular Weight	Benzene Fraction	H_2S-Rich Acid Gas	Foul Water
Stream Number		3601	3602	3605
Weight Flow, M lbs/hr		5.7	36.3	0.1
MM SCFD, dry basis		--	8.1	--
Density at Op. Conditions		51.2	0.27	57.3
Temperature, °F		64	36	300
Pressure, psia		60	35	59
Molecular Weight		76.4	40.7	18
Component, mols/hr				
CO_2	44.01	0.5	603.9	--
H_2S	34.08	0.4	279.9	--
H_2O	18.02	--	--	5.6
H_2	2.02	--	0.3	--
CO	28.01	--	0.2	--
CH_4	16.04	--	2.3	--
C_2H_6	30.08	0.1	3.3	--
C_6H_6	78.12	65.8	0.2	--
C_7H_8+	94.06	4.4	--	--
CH_3OH	32.04	3.4	0.8	--
Total		74.6	890.9	5.6

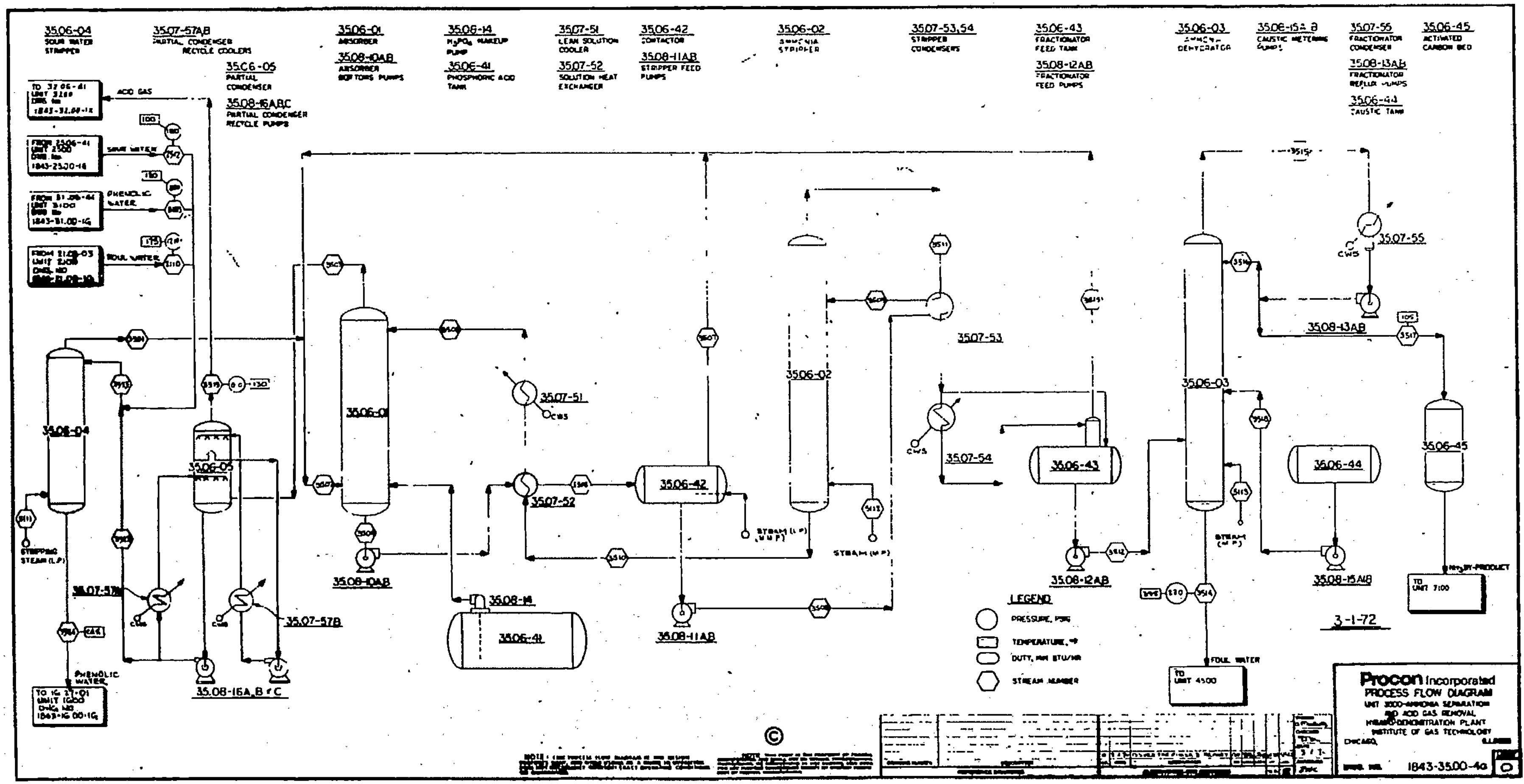


Figure 8-27. AMMONIA SEPARATION SECTION PROCESS FLOW DIAGRAM FOR THE HYGAS DEMONSTRATION PLANT

Table 8-16. AMMONIA SEPARATION SECTION STREAM CONDITIONS, COMPOSITIONS AND MATERIALS BALANCES FOR THE HYGAS DEMONSTRATION PLANT

Stream Description	Component Molecular Weight	Water to Sour Water Stripper	Water to Sour Water Stripper	Water to Sour Water Stripper	Condensate to Sour Water Stripper	Combined Feed to Sour Water Stripper	Sour Gases to Absorber	Stripper Bottoms to Unit 1600	Dry Acid Gas to Sulfur Recovery	Fractionator Bottoms	Product Ammonia
Stream Number		2512	3103	2110	3522	3523	3501	3524	3519	3514	3517
Weight Flow, M lbs/hr		113.6	7.2	25.4	22.1	169.2	24.8	174.0	10.2	16.1	2.33
MM SCFD, dry basis		--	--	--	--	--	3.40	--	2.10	--	--
Density at Op. Conditions		61.3	61.5	60.5	59.3	61.5	0.078	59.0	0.144	58.0	37.5
Temperature, °F		100	120	175	230	130	215	245	130	395	105
Pressure, psia		1100	295	1225	30	25	24.7	28	22.7	235.0	--
Molecular Weight		18.54	18.00	19.31	18.06	18.56	23.15	19.35	39.90	18.0	17.03
Fuel Equivalent											
Component, mols/hr											
NH_3	17.03	117.55	2.5	43.5	2.1	165.65	140.0	25.65	0.05	0.9	136.5
H_2O	18.02	5830.1	389.3	1189.8	1218.1	8627.3	700.0	9627.3	24.2	889.2	0
CO_2	44.01	118.7	0.8	61.3	1.8	182.60	180.0	2.6	178.2	nil	0
H_2S	34.08	27.2	0.7	15.1	0.7	46.0	45.0	1.0	41.7	nil	0
Other Gases		12.1	1.8	5.1	0.5	17.9	8.9	9.0	8.4	nil	0
NaOH	40.00	--	--	--	--	--	--	--	--	0.6	--
(gpm) Liquid		215	15	52	52	343	--	368		37.2	8.0
Total		6105.7	395.12	1315.2	1223.1	9041.5	1073.4	9040.1	252.6	890.7	136.5

A7506 1672

In the Phosam process, the stripped gases are scrubbed in an absorber by an aqueous solution of ammonium phosphate, which selectively removes ammonia from the gas. Gas from the absorber contains less than 200 ppm of ammonia and it is cooled before flowing on to the Claus plant. The rich solution from the absorber is flashed to remove dissolved acid gases and then is stripped of excess ammonia and water. The regenerated solution is cooled and recirculated to the absorber. The ammonia-water overhead of the stripper is fed to a dehydrator tower from which an overhead product of 30 tons/day of anhydrous ammonia is produced. Water from the dehydrator flows to waste water treatment.

8.4.4.4 Sulfur Recovery (Claus Plant)

The Claus plant converts H_2S to elemental sulfur by these chemical reactions:

$$H_2S + 1.5\ O_2 \rightarrow H_2O + SO_2$$

$$2H_2S + SO_2 \rightarrow 2H_2O + 3S$$

Two reactor stages are used, which results in greater than 93% sulfur recovery. A process flow diagram is presented in Figure 8-28 and stream data appear in Table 8-17.

Two sources of feed gas enter the plant. The largest flow is from the acid-gas stripping section and the other from the ammonia separation plant. One-third of the total acid gas is burned in the noncatalytic furnace and its heat of combustion generates 250 psig steam. Combustion gases are mixed with bypass gas to yield the desired H_2S/SO_2 ratio of 2. In the first stage reactor, H_2S and SO_2 react over a bauxite catalyst in the 460° to 660°F temperature range. Effluent gas is cooled, and condensed sulfur is separated.

Reheated gas flows to the second-stage reactor, where additional sulfur is formed. After condensing and separating the sulfur product, the tail gas flows on to incineration. It contains slightly more than 1% of sulfur compounds, which are mostly unreacted H_2S and SO_2. Other sulfur compounds in the tail gas are COS, CS_2, sulfur vapor, and liquid sulfur mist. Liquid sulfur is collected in a pit, from which it is pumped to solids sulfur storage. About 100 long tons of sulfur/day are produced.

8.4.4 5 Incineration and Tail Gas Treatment

The tail gas from the Claus plant is too polluting to vent directly to the atmosphere. The tail gas is combined with sour vent gas from the char feed preparation section and burned to convert all of the sulfur to SO_2. The combustion gases are then scrubbed by a limestone slurry to remove SO_2 before release to the stack. The limestone scrubbing tower is located near the Claus unit. It receives a slurry slipstream from a central limestone slurry supply system that also supplies slurry to the power plant stack scrubbers (see Figures 8-29 and 8-30).

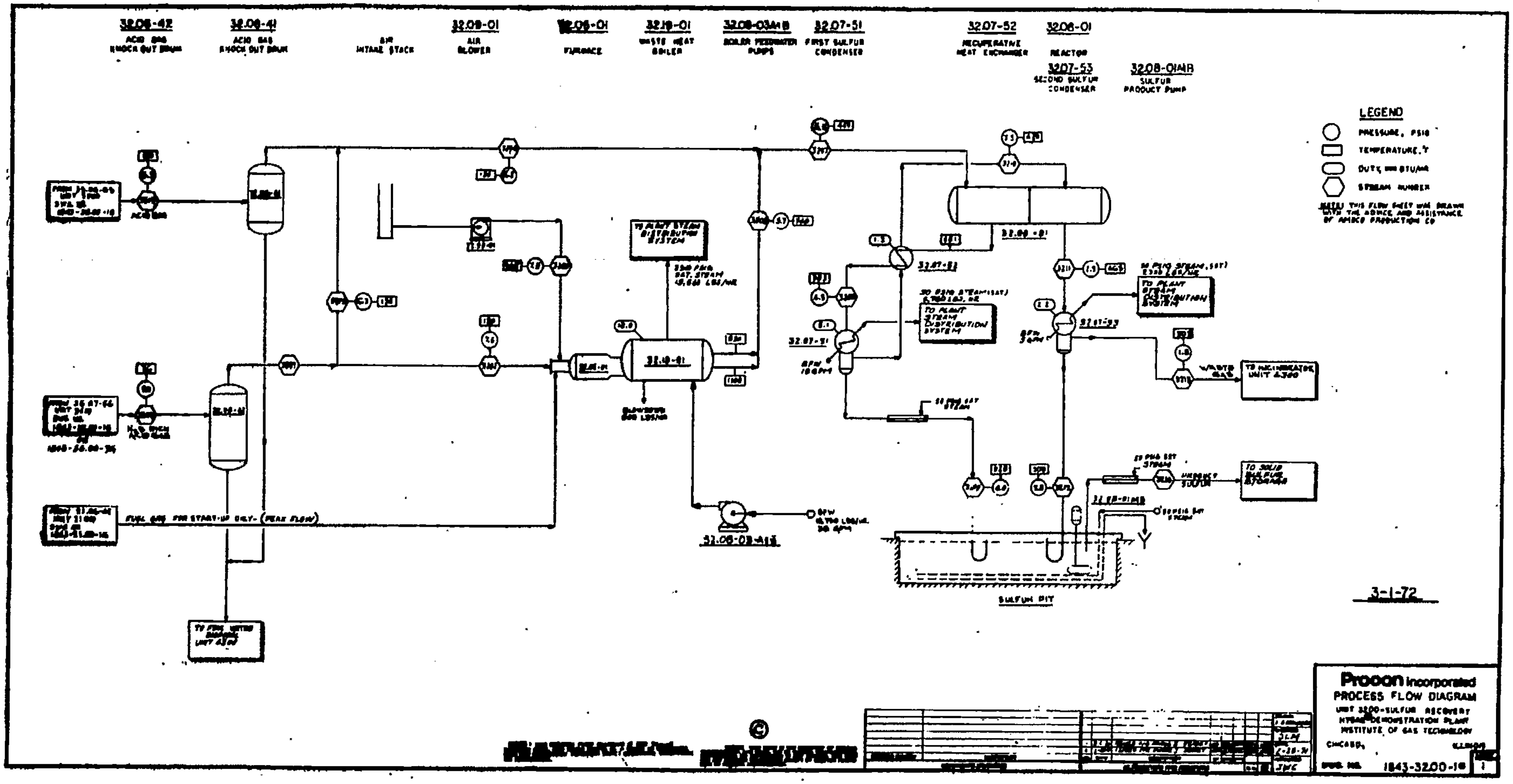


Figure 8-28. CLAUS PLANT SULFUR RECOVERY PROCESS FLOW DIAGRAM FOR T" HYGAS DEMONSTRATION PLANT

Table 8-17. CLAUS PLANT SULFUR RECOVERY STREAM CONDITIONS, COMPOSITIONS AND MATERIALS BALANCES FOR THE HYGAS DEMONSTRATION PLANT

Stream Description	Component Molecular Weight	Acid Gas From Acid Gas Stripping Unit	Acid Gas Feed to Furnace	Furnace Bypass Acid Gas	Acid Gas From Phosam Unit	Combined Furnace Bypass Acid Gas	Furnace Air	Liquid Sulfur Product	Liquid Sulfur Sulfur	Waste Gas to Incinerator	Total Sulfur Product
Stream Number		3602	3202	3203	3519	3204	3205	3209	3212	3213	3214
Weight Flow, M lbs/hr		36.250	16.340	19.910	10.180	30.090	20.125	7.640	1.510	57.400	9.150
MM SCFD, dry basis		8.10	3.65	4.45	2.10	6.64	6.21	--	--	15.20	--
Density at Op. Conditions											
Temperature, °F		36	130	130	130	130	202	325	305	305	325
Pressure, psia		34	22.7	22.7	20.6	20.6	22.1	18.3	16.3	16.1	--
Molecular Weight		40.7	40.7	40.7	39.9	41.2	28.7	32.07	32.07	34.3	32.07
Fuel Equivalent											
Component, mols/hr											
H_2S	34.08	279.0	125.97	153.93	41.7	180.03	--	--	--	10.76	--
SO_2	64.07	--	--	--	--	--	--	--	--	5.32	--
S_2	64.13	--	--	--	--	--	--	--	--	--	--
S_4	128.26	--	--	--	--	--	--	--	--	--	--
S_6	192.40	--	--	--	--	--	--	--	--	0.12	--
S_8	256.53	--	--	--	--	--	--	--	--	0.35	--
Si (Liquid)	32.07	--	--	--	--	--	--	238.78	47.14	2.35	285.42
COS	60.08	0.11	0.05	0.06	--	0.06	--	--	--	0.02	--
CS_2	76.14	--	---	--	--	--	--	--	--	0.02	--
Subtotal Atoms S		280	125	154	42	180	--	238	47	37	285
CO_2	44.01	603.9	272.10	331.80	178.2	510.0	--	--	--	781.6	--
NH_3	17.03	0.13	0.06	0.08	0.05	0.13	--	--	--	0.13	--
CO	28.01	--	--	--	1.53	--	--	--	--	6.15	--
H_2		--	--	--	--	--	--	--	--	1.42	--
CH_4		4.60	2.07	2.53	4.00	6.53	--	--	--	4.00	--
C_2H_6		2.02	0.91	1.12	2.87	3.99	--	--	--	2.87	--
H_2O	--	--	--	----	24.2	24.2	18.95	--	--	316.29	--
O_2		--	--	--	--	--	143.03	--	--	--	--
N_2		--	--	--	--	--	539.69	--	--	540.72	--
Total		890.67	401.46	489.52	252.6	729.94	701.67	238.28	47.14	1672.12	285.42

A7506 1673

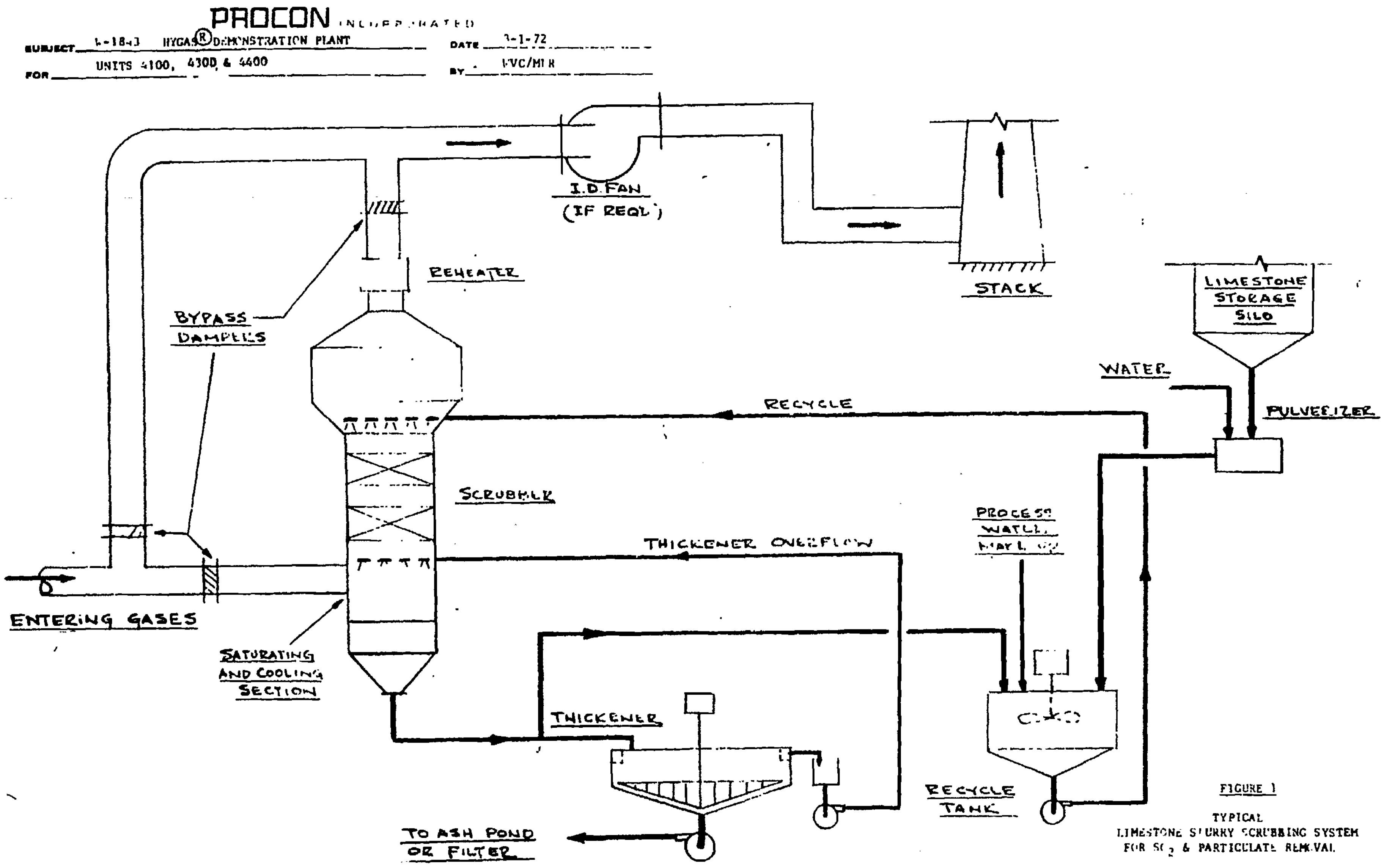


Figure 8-29. TYPICAL LIMESTONE SLURRY SCRUBBING SYSTEM FOR SO_2 AND PARTICULATE REMOVAL

PROCON INCORPORATED

SUBJECT V-1843 HYGAS® DEMONSTRATION PLANT DATE 3-1-72

FOR UNITS 4100, 4300 & 4400 BY WVC/MJR

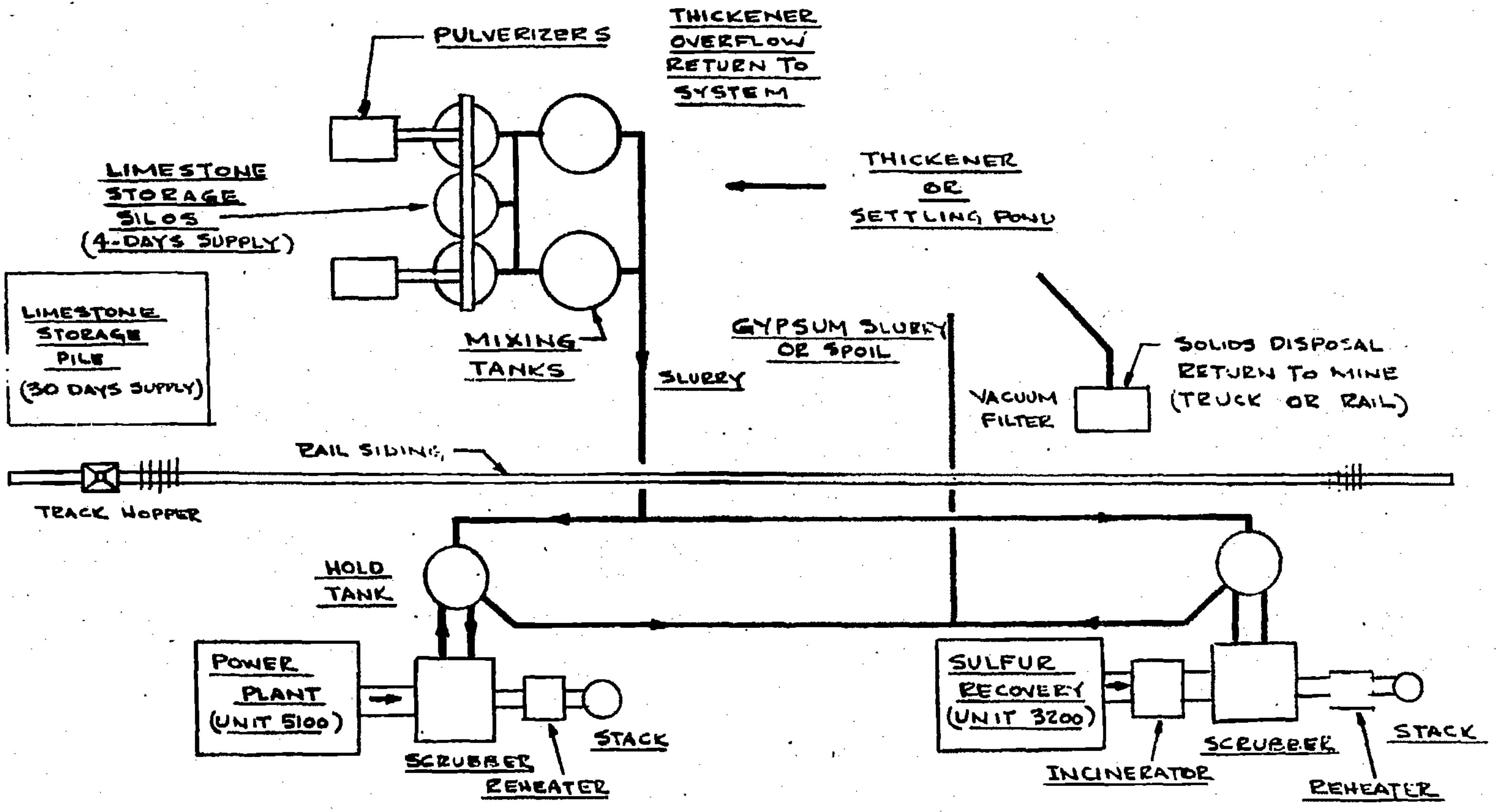


Figure 8-30. PROCON-DESIGNED SCRUBBING SYSTEM FOR DEMONSTRATION PLANT

8.4.4.6 Waste Water Treatment

The waste-water treatment system was designed to meet the standards of the State of Illinois. Several different sources of waste water enter this system. The flow rates of these streams, their contaminants, and the methods of treatment used are discussed (see Figures 8-31 and 8-32).

The process and power plant cooling towers blow down about 800 gpm of water to control dissolved solids buildup to 1000 ppm. A chromate corrosion inhibitor is used in conditioning cooling tower water. The 20 milligrams per liter (mg/ℓ) of hexavalent chrome in the blowdown water must be reduced to trivalent chrome so that chromic hydroxide can be precipitated by lime addition. The reduction is performed using sodium bisulphite or SO_2. After precipitation in the hydroxide form, less than 1 mg/ℓ of trivalent chrome is contained in the treated water (see Figure 8-33).

Part of the treated cooling-tower blowdown water joins the plant effluent and the remainder is used as makeup in the limestone slurry section. Coal-pretreatment quench-section water supplies the rest of the water makeup to the limestone slurry system. About two-thirds of the water entering the limestone slurry system is lost by evaporation; the rest is associated with the gypsum cake discharged from the system.

Excess water flows from the char recovery unit at 495 gpm. This water is contaminated by 1600 mg/ℓ of NH_3, 370 mg/ℓ of H_2S, 420 mg/ℓ of oil, and 10 mg/ℓ of phenolic material. Water from the ammonia separation section containing ammonia and caustic (1500 mg/ℓ) joins the char recovery water at a rate of 38 gpm. The combined stream is stripped to reduce its NH_3 and H_2S concentrations to below 100 ppm. The stripped water is joined by regeneration water from the power plant water-treatment facilities and water-softener backwash water. These two streams contain about 1000 ppm of total dissolved solids, and their average flow rate is 570 gpm. An additional 0.2 gpm of water containing 100 mg/ℓ of methanol from the acid-gas stripping section joins the above water streams as they flow into the aerated lagoon.

The aerated lagoon is designed to treat 1.6 million gal/day with a holding time of 2 days. The lagoon has a depth of 10 feet and a surface area of about 1 acre. Three floating aerators cool the water and raise the dissolved oxygen content for biological treatment.

From the aeration lagoon, the water enters a two-stage, activated sludge system which uses an enriched oxygen atmosphere. The first stage removes H_2S and organic material. The second stage removes NH_3. The H_2S is converted to sulfites or sulfates, NH_3 to nitrates, and the organic material to water, CO_2, and a biological cell mass. This system differs from a conventional sludge system in its use of oxygen instead of air, the covering of the activated sludge tanks, and the internal partitioning of the sludge tanks. Wasted sludge is oxidized in aerobic digestion tanks using waste gas from the activated sludge tanks.

Blowdown streams are collected from process steam generators, from the high-pressure process boiler, and from the electrothermal gasifier power-plant boiler. The total flow rate is about 206 gpm and the dissolved

PROCON INCORPORATED

SUBJECT -1843 - HYGAS® DEMONSTRATION PLANT DATE 7-1-72

FOR UNIT -500 - FOUL WATER DISPOSAL BY BPS/MFR

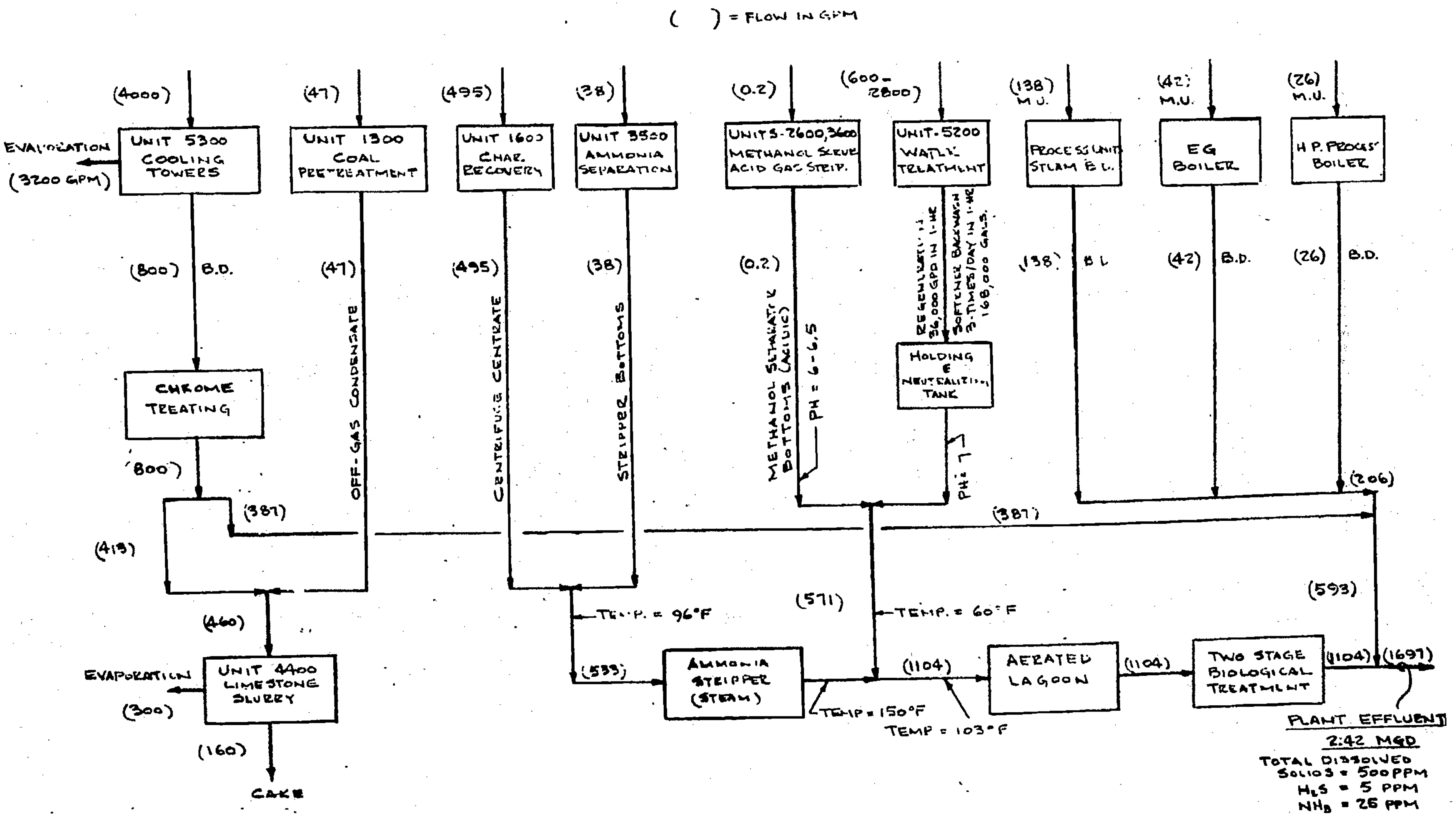


Figure 8-31. FOUL WATER DISPOSAL FLOW DIAGRAM

PROCON INCORPORATED

SUBJECT 6-1843R HYGAS® DEMONSTRATION PLANT DATE 1-1-77

FOR UNIT 4500 - FOUL WATER DISPOSAL BY BPS/HER

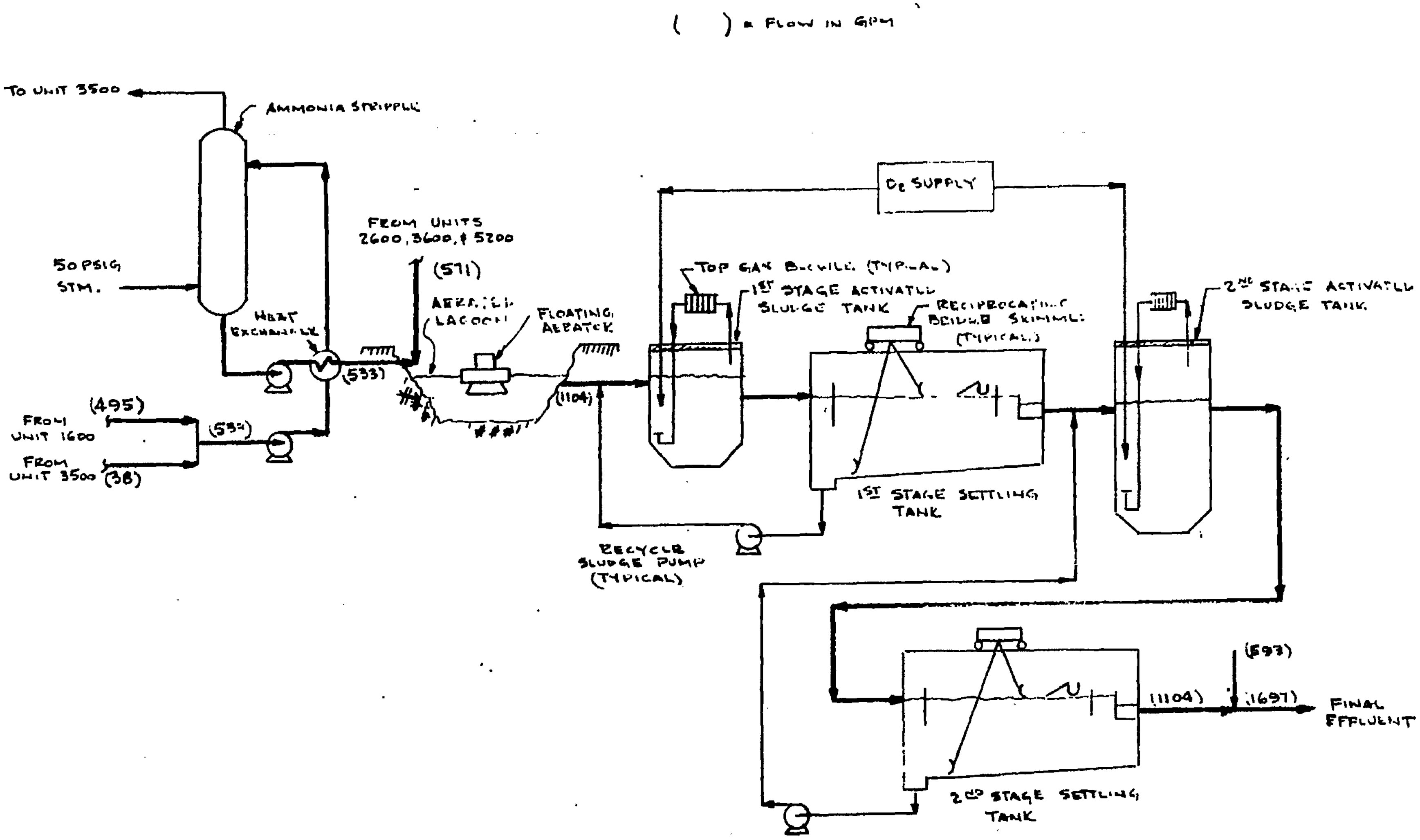


Figure 8-32. FOUL WATER DISPOSAL PROCESS FLOW DIAGRAM FOR THE HYGAS DEMONSTRATION PLANT

PROCON INCORPORATED

SUBJECT V-1847 HYGAS® DEMONSTRATION PLANT DATE 3-1-77
FOR UNIT 4500 - FOUL WATER DISPOSAL BY BPS/MIR

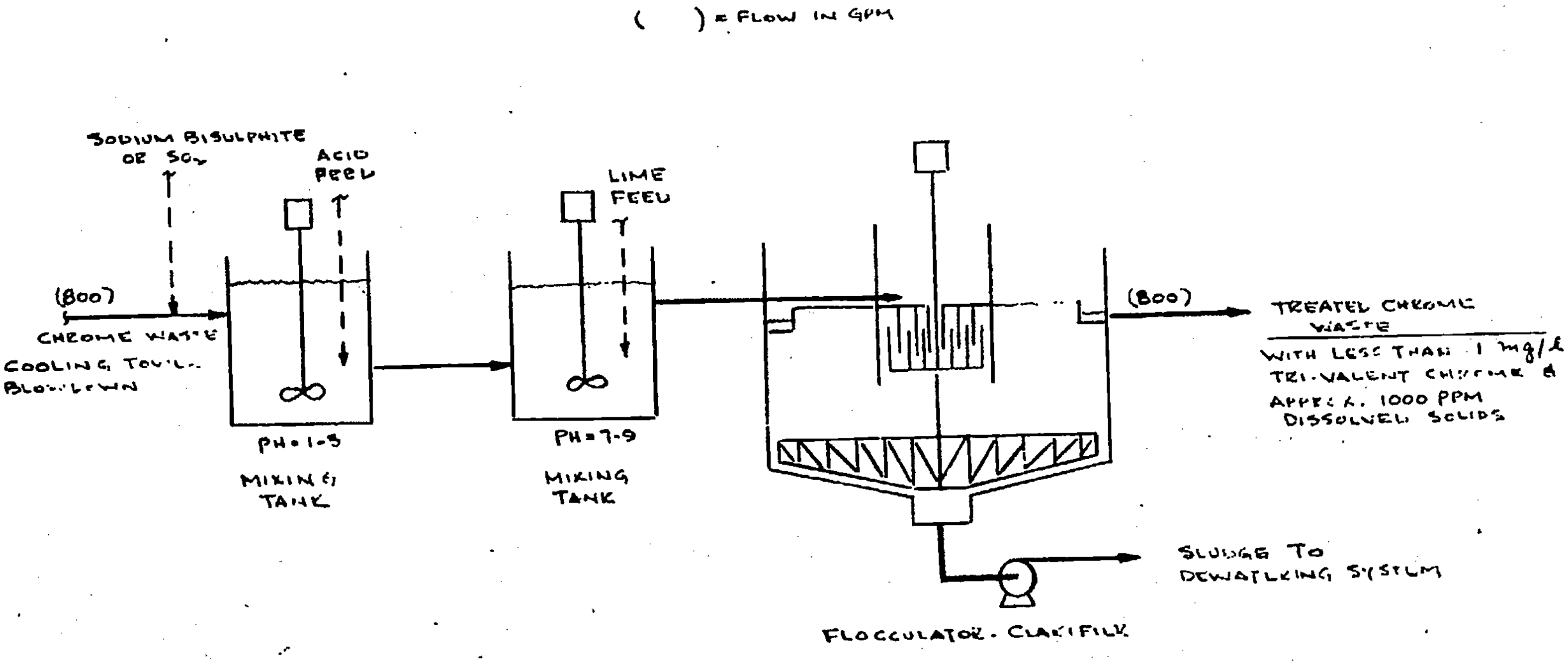


Figure 8-33. COOLING TOWER BLOWDOWN AND CHROME TREATMENT FOR THE HYGAS DEMONSTRATION PLANT

solids content averages about 2000 ppm. These streams are mixed with part of the treated cooling-tower blowdown and with all of the water from the two-stage biological treatment. The combined stream represents the treated plant effluent water and flows at a rate of 2.4 million gal/day. The water contains 500 ppm of total dissolved solids, 5 ppm of H_2S, 2.5 ppm of ammonia, and less than 0.2 ppm of phenols.

8.4.5 Utilities and Miscellaneous Support

Figure 8-34 shows the interrelationship of the utilities and other miscellaneous supporting facilities. Figure 8-35 is a process flow diagram of the steam and water utilities for the HYGAS demonstration plant. A proposed site plan is shown in Figure 8-36, and a proposed plot plan is presented in Figure 8-37.

8.4.5.1 Water Treatment

The quality of the local water supply and its specific treatment were not known because the plant site has not been selected; however, clean pre-treated water was assumed to be available at the plant battery limits. Feedwater to the high-pressure boilers is demineralized and deionized: the estimated requirement is 1130 gpm. Softened water at 4100 gpm is supplied to low-pressure steam generators and for cooling-tower makeup. Allowing for demineralizer and softener regeneration, backwash requirements, utility and fire protection needs, and water for drinking and sanitary facilities, the total water requirement is about 6500 gpm.

In the gasification process, 520 gpm of water is chemically consumed. Boiler blowdown requires 206 gpm of water and cooling tower evaporation and blowdown amounts to 3600 gpm. The wet limestone scrubbing system requires 460 gpm, which is evaporated or lost as moisture in the solids product. The remaining quantity of water leaves the plant as treated waste water.

8.4.5.2 Boilers and Power Plant

To supplement the saturated 1250-psig steam generated in the main reactor water jacket, a 400,000 lb/hr high-pressure process boiler is used which generates both saturated and superheated 1250-psig steam. At several locations in the process 250-, 100-, and 50-psig steam is generated. The lower pressure steam is used mainly for turbine drives, for heating, and for deaeration. The high-pressure steam is used in the gasifier reactor and in the shift conversion unit.

The electrothermal gasifier requires 110,000 kW of electricity. This power is generated by an in-plant boiler turbine-generator set, which is fueled by char produced from the EG unit and by pretreater off-gas. Including associated auxiliaries and controls, a total power plant capacity of 116 MW is needed. The superheat-reheat steam boiler operates at 1800 psig, 1000°F steam conditions, and 3.5-in. Hg of turbine exhaust pressure. The system is designed to utilize available heat from process units for feedwater heating.

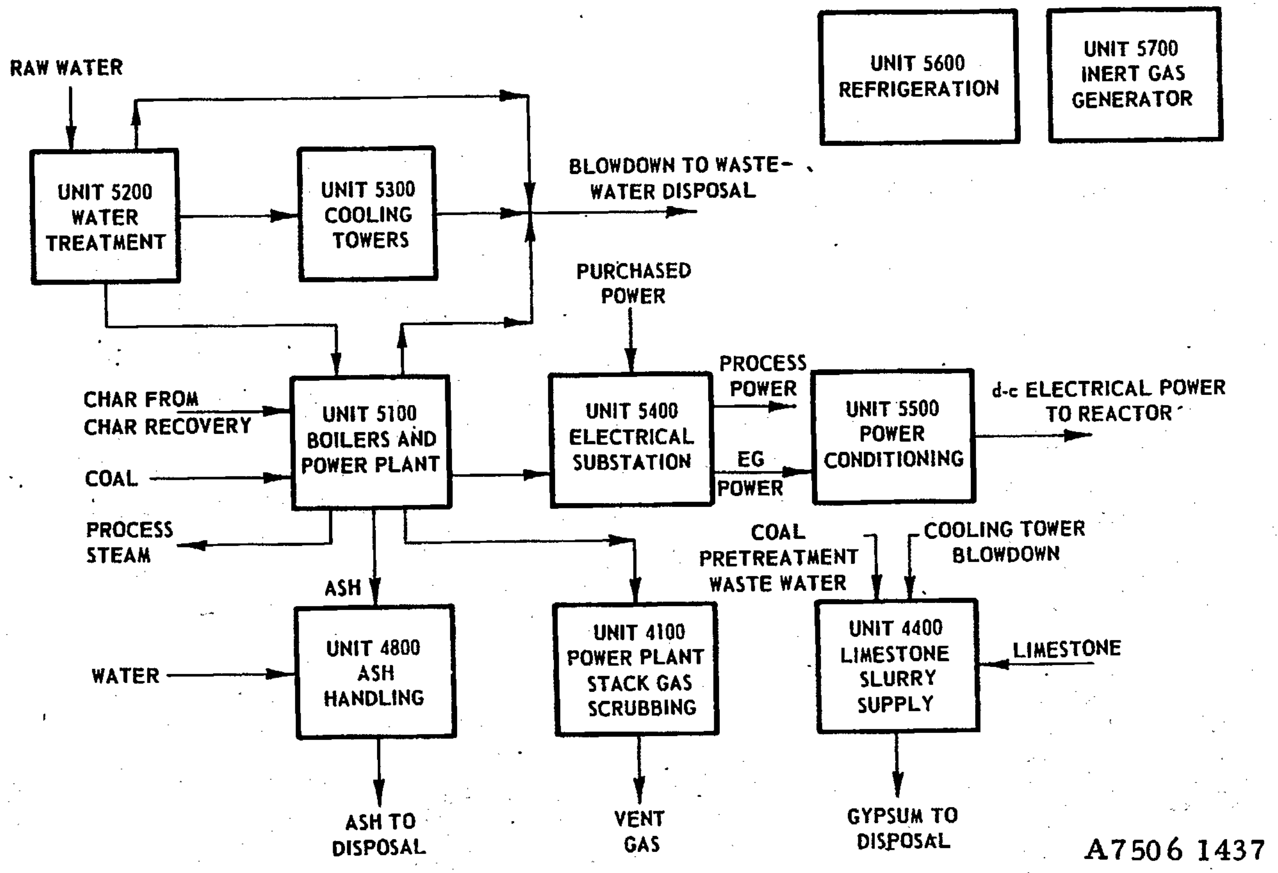


Figure 8-34. HYGAS DEMONSTRATION PLANT UTILITIES AND MISCELLANEOUS SUPPORT FACILITIES

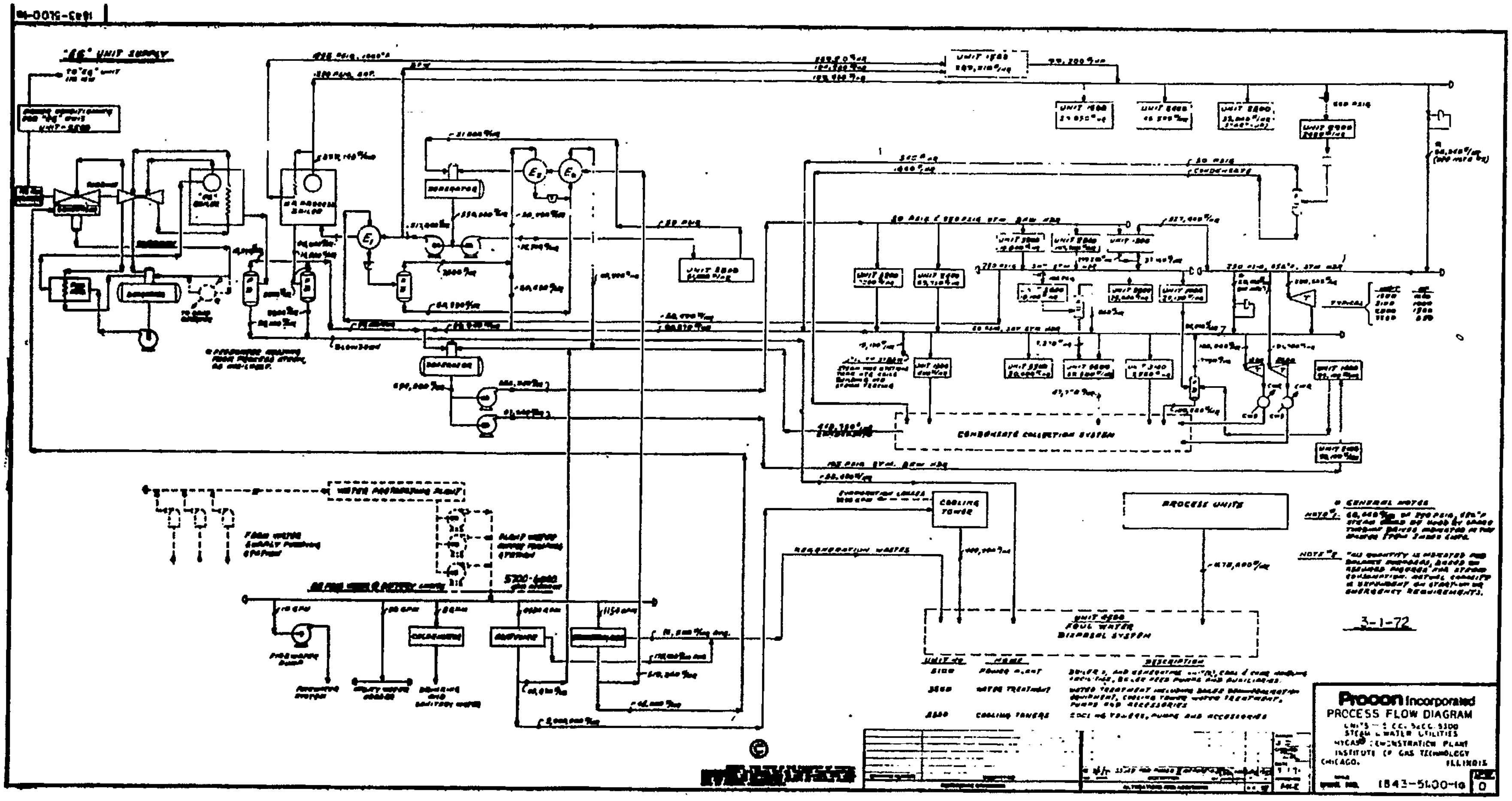


Figure 8-35. PROCESS FLOW DIAGRAM FOR STEAM AND WATER UTILITIES

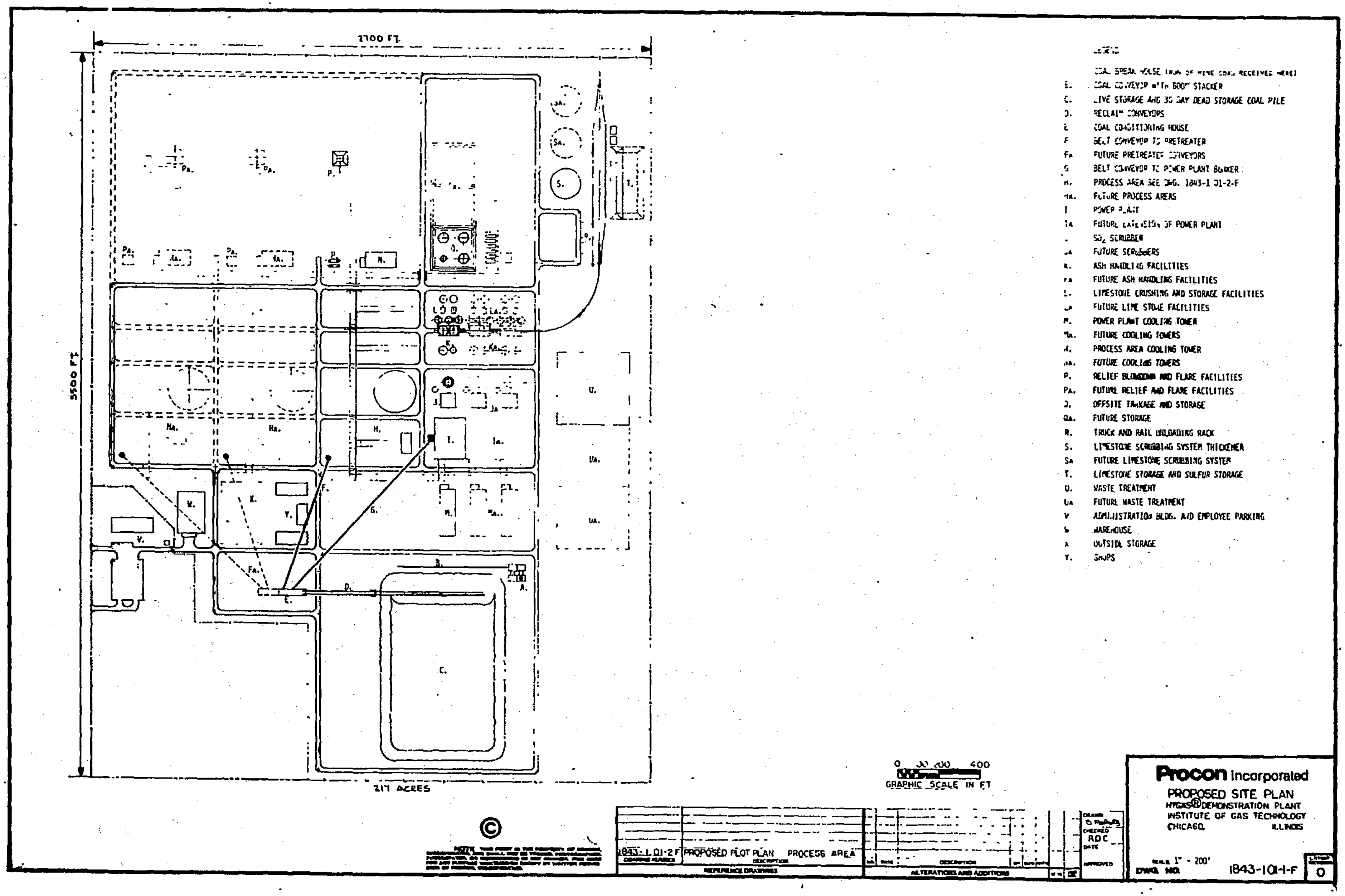


Figure 8-36. PROPOSED DEMONSTRATION PLANT SITE PLAN

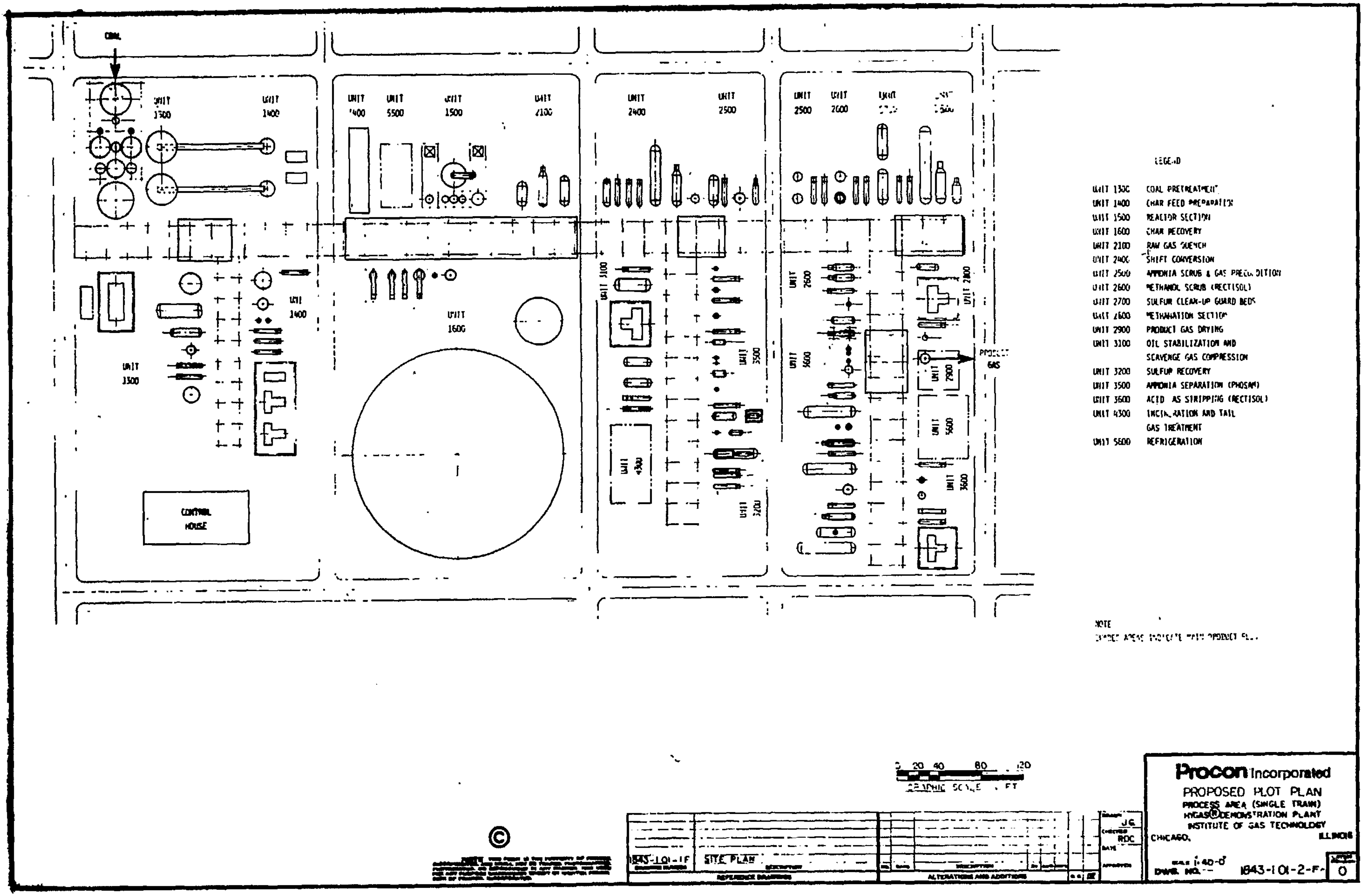


Figure 8-37. PROPOSED SINGLE-TRAIN PROCESS AREA PLOT PLAN

Plant power requirements, aside from the EG unit, amount to 40,000 kW. Purchased power fulfills this requirement.

8.4.5.3 Power Plant Stack-Gas Scrubbing and Limestone Slurry Supply

The electrothermal gasifier char which is used to fuel the boilers and power plant contains 2.75% sulfur and 37.8% ash. Sulfur content of the fuel is high enough that equipment must be provided to scrub the power plant stack gas for SO_2 removal. About 60 tons/hr of electrothermal gasifier char is the primary boiler fuel; the secondary fuel is off-gas from coal pretreatment which has a heating value of 39 Btu/SCF and contains 0.45% SO_2. The power plant flue gas flow will be about 844,000 actual CF/min at 300°F and 14.2 psia. It will contain 0.29% SO_2. This represents about 45% of the sulfur contained in the feed coal.

A wet limestone scrubbing system was selected to clean the power plant stack gas primarily because it appears to be the stack-scrubbing system which will be most widely used by the utility industry. Wet limestone scrubbing is used not only to clean the boiler stack gases, but also to scrub a relatively small incinerated stream of Claus plant tail gas.

The limestone makeup rate to the slurry supply system is 22 tons/hr. This represents a stoichiometric excess of 50% limestone. Moist spent stone for disposal amounts to 53 tons/hr. The gas to be cleaned enters the cooling section of the scrubber where it is saturated with moisture at a temperature ranging from 115° to 135°F. Fly ash is also removed during the cooling process. The saturated gas then enters the scrubber absorption section where the SO_2 content of the gas is reduced by reaction with limestone slurry to form calcium slufite and some calcium sulfate. Formation of calcium sulfate may cause scaling problems. To reduce this problem, the slurry concentration in the scrubber recycle stream is held from 5 to 10%. Effluent slurry from the scrubber flows to a thickener and from there to an ash pond or filter. Ultimate disposal of the spent stone will be in the coal mine. The remaining support facilities are described briefly.

8.4.5.4 Ash Handling

The total ash production rate from the boiler is estimated to be about 26 tons/hr. Molten ash from the slagging boiler discharges from the furnace bottom in a continuous stream into water-filled slag tanks. The solidified ash in water slurry is periodically transferred to dewatering bins from which conventional dry ash is transferred into the coal mine for disposal. Ash handling is outlined in Figure 8-38.

8.4.5.5 Refrigeration Unit

The refrigeration unit is designed to deliver 1500 tons of refrigeration at −50°F to the acid-gas scrubbing section and 900 tons of refrigeration at 60°F to the ammonia scrub and gas-preconditioning section. A conventional ammonia refrigeration system is used.

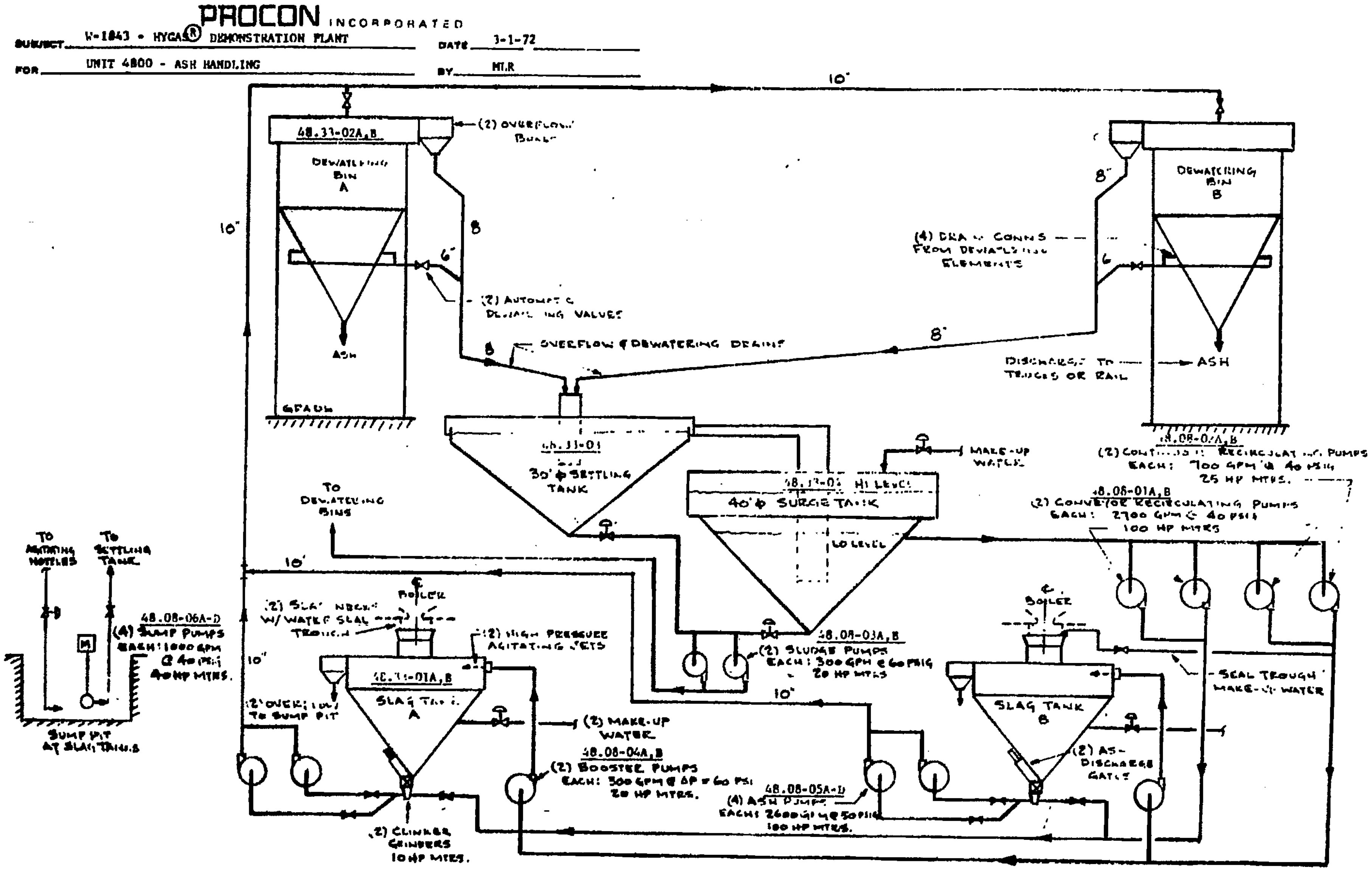


Figure 8-38. ASH HANDLING SYSTEM FOR THE HYGAS DEMONSTRATION PLANT

8.4.5.6 Inert Gas Generator

The inert gas generator is designed to supply a maximum flow of 2100 SCF/min by combustion of pipeline gas. Gas from the generator is cooled and compressed to 100 psig and 1250 psig and stored in receivers of 6000-SCF capacity each. The inert gas is used primarily during startup and shutdown of the plant, although during operation it is used for some instrument purges.

8.4.5.7 Storage Tanks

The storage tanks planned are as follows: 3 slurry oil tanks of 9000-bbl capacity each, 6 ammonia tanks having a capacity of 34,000 gal each, 1 methanol tank of 28,000 gal capacity, a 2000-gal storage tank for glycol, a 4000-gal tank for phosphoric acid, and a 10,000-gal tank for caustic storage.

8.4.5.8 Mobile Equipment

The following mobile equipment is needed. Twenty heavy-duty, off-road dump trucks will be required for ash and gypsum disposal; 1600 to 1700 tons/day must be hauled 7 days/week at 1 shift/day. One clam-shell bucket, rail or caterpillar mounted, is needed as well as one heavy-duty bulldozer and two Huff Payloaders to unload and transfer a maximum of 550 tons/day of limestone to and from the dead storage pile and to load an average of 100 tons/day of sulfur into railroad hopper cars. The additional equipment needed includes two pickup trucks, one fire engine, one A-frame flat-bed truck, one 1-1/2 ton flat-bed truck, one Dempster Dumpster with four bins, one ambulance, and 3 sedan automobiles.

8.5 Manpower Estimate

The total number of personnel needed to operate the HYGAS demonstration plant is estimated to include 60 salaried and 281 hourly people. This estimate includes the plant manager; an office manager with accounting, purchasing, and warehousing responsibilities; a safety and inspection manager; a technical manager who handles technical service, engineering, and laboratory; an operating superintendent; and a maintenance superintendent. In operations, a total of 24 operators will be needed with 96 helpers, 12 boardmen, and 39 shift breakers and extra coverage men. The maintenance manpower includes 20 pipefitter-welders, 25 mechanics, 20 electricians and instrument men, 20 general utility men, and 25 laborers. A plant manning organization chart appears in Figure 8-39.

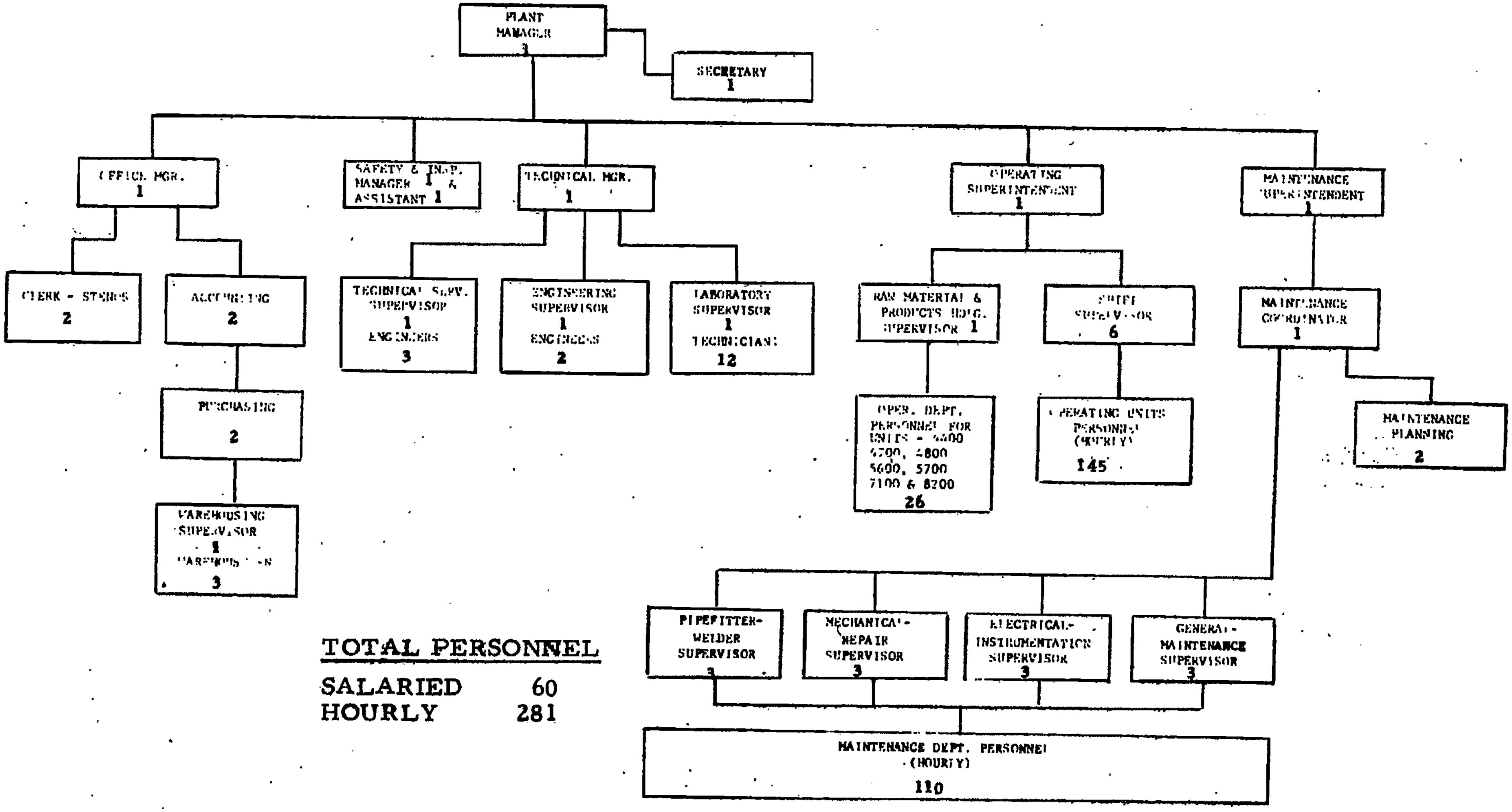
8.6 Project Schedule

Field construction could begin 18 months after engineering is started. At this time about 50% of the engineering should be completed. The construction time will cover a period of up to five years, dependent upon the procurement schedule, which is controlled by design and field construction of the hydrogasification reactor. The boilers and turbine generator in the power plant would require about 21 months for delivery and field erection.

PROCON INCORPORATED

SUBJECT 6-1843 - HYGAS® DEMONSTRATION PLANT DATE [illegible]-1-72

FOR PLANT MANNING ORGANIZATION CHART BY PAB/MIR



TOTAL PERSONNEL

SALARIED	60
HOURLY	281

Figure 8-39. PLANT MANNING ORGANIZATION CHART FOR THE HYGAS DEMONSTRATION PLANT

8.7 Suggested Areas for Design Improvement

The HYGAS demonstration plant design described uses commercial processes or techniques throughout except for the coal gasifcation reactor and the methanation system. This preliminary design shows the feasibility of constructing large coal gasification plants as soon as the gasification and methanation steps are proved at the pilot plant level. In studying this design, it is apparent that improvements in several areas could contribute to increasing plant reliability, decreasing costs, and reducing process complexity. Most of the improvements require new designs, new equipment, or new processes; discussion follows.

8.7.1 Process Design Improvements

8.7.1.1 Generation of Hydrogen

Electrothermal gasification as used in this design for the generation of reducing gas has several advantages. A minimum amount of CO_2 is contained in the gases produced from the electrothermal gasifier unit. The gasifier bed is of simple design, and the gasification rate is readily controlled. By withdrawing difficult-to-gasify char for power plant fuel, only the more reactive portions of the coal feed are gasified.

In a practical plant design, however, the disadvantages of electrothermal gasification become apparent. The large conventional power plant required is not only expensive in capital cost, but it is also costly in reducing the plant's thermal efficiency for converting coal to pipeline gas. Because of the char fuel's sulfur content, stack-gas scrubbing is needed for the power plant. The total cost of the power plant, its stack-gas scrubbing system, and the power-conditioning system amounts to more than half the cost of all the process units. Electricity is supplied to the electrothermal gasifier as direct current at about 13,000 volts and 8400 amperes. Very real design problems are encountered in safely introducing this electricity into a process vessel that operates at 1200 psi and 1900°F. These questions must be answered: What kind of pressure seal can be used on the electrical cable and how will the cable be electrically insulated as it enters the vessel? The current flows through the fluidized bed of the EG reactor, radially, from a central electrode to a peripheral concentric electrode. With this electrode configuration, the heat release is far greater per unit of bed volume near the center electrode than in an equal volume near the outer radius of the bed. Unless heat release is carefully modified by electrode design or by control of flows within the fluid bed, hot spots may occur causing localized fusion of coal ash and of the central electrode surface.

An alternative to electrothermal gasification is steam-oxygen gasification (OG). The function of OG is the same as EG, that is, to generate reducing gas by the steam carbon reaction with char from the high-temperature reactor. A major difference is in how the heat of reaction is supplied. Heat is supplied to the OG fluid bed for the steam-carbon reaction by burning part of the carbon with oxygen.

The OG operates from 1850° to 1900°F. Instead of producing a large volume of char as does the EG, the OG's solid product is a low-carbon-content ash. More CO_2 is generated in the OG gasifier which increases