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FEASIBILITY STUDY

HYPERSORPTION PROCESS FOR SEPARATION OF COMPONENTS OF A MEDIUM -BTU GAS

FINAL REPORT
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PREPARED
BY

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1. SUMMARY

This study has been performed to determine the technical and economic feasibility of employing hypersorption process technology to separate and purify a medium - BTU gas, derived from oxygen-blown gasifiers, to obtain a H_2/CO ratio of 2:1 as a feed to a Fischer-Tropsch type plant. Technical feasibility is a measure of the ability to design a hypersorption separation and purification process from available data. Economic feasibility can be made through comparisons with commercially available process technology.

Three gasification processes have been used as a basis for this study. These processes are based upon EPRI Report AF-244 for the Lurgi moving bed, oxygen-blown, dry bottom gasifier, and EPRI Report AF-642 for the Texaco, slurry fed, oxygen-blown, entrained bed gasifier and for the Foster-Wheeler, oxygen-blown, entrained bed gasifier. Process designs for the hypersorption separation and purification for each case have been made using engineering judgments based on the available adsorption isotherms, which have been obtained from the Calgon Corporation. No attempt has been made to optimize these designs although some additional studies have been done where it has been deemed desirable. Cryogenic separation and acid gas purification for each case have been supplied by Lotepro as a packaged unit.

In addition to evaluating the technical and economic feasibility, a research plan has been presented to aid in alleviating the risks associated with designing a hypersorption process to separate and purify medium -BTU gases obtained from oxygen-blown gasifiers.

The moving bed adsorption process was patented by Soddy in 1922 to separate coal gases. Several other processes were patented between 1922 and 1946. In 1946 Berg working for the Union Oil Company of California published results of separation obtained with a moving charcoal bed. The process was renamed hypersorption. Berg's work excited world wide interest. In the United States, six large-scale hypersorption units were constructed by Foster-Wheeler. These units had an early demise and only sparse data is available on their operation. No indication has been given about the reasons for their early shut down and no new units have been built. Very little data has been found in the literature particularly for multicomponent systems. Due to the lack of available data, proprietary information from the Calgon Corporation has been used for the design of the hypersorption units.

The hypersorptive processing scheme is based upon single isotherm data. Pressure and temperature requirements for the adsorption are based on the gases to be separated. Two conditions prevailed, the possible azeotropic formation of CH_4 and CO and the design requirement that there was to be no interstage compression. In both the Lurgi and the Foster-Wheeler cases three hypersorption towers are required. Hypersorber 1 separates H_2 and CO from all the other gases, hypersorber 2 separates H_2 from CO and hypersorber 3 separates the CH_4 from the acid gases. Texaco gasification products are lean in CH_4. Therefore only two hypersorbers are required. Sufficient trains are supplied to handle the volume of gases to be processed.

Direct heating and cooling of the activated carbon eliminates the need for heat exchangers within the hypersorption towers, lowering the initial investment cost. In one case the operating charges were adversely effected by using a direct heating design. A dense phase transport system has been chosen to minimize attrition losses. CO shift conversion is applied where excess CO is available. Sulfur clean up has also been addressed.

Economic evaluations are ambiguous. All plant investments are within the ± 30% accuracy of this study. No clear cut choice between cryogenic separation - acid gas purification and hypersorptive separation - purification can be made based on plant investment. Operating costs are within the ± 30% accuracy. However, in the Foster-Wheeler case the operating costs are 28% greater for the hypersorption process.

Part of the ambiguity results from the method of obtaining plant investment costs. The cost of equipment not required for the separation and purification are the same as those in the EPRI Reports. That is, in dollars for the year the report was published, with minor modifications as were necessary to design a complete system. The cost of equipment required for the hypersorption and cryogenic processes are in 1981 dollars. Escalation of the previous costs introduces uncertainty in the results. Capital charges have not been evaluated because they are outside the scope of the work.

Detailed design of a hypersorption unit required knowledge of: maximum loading on the adsorbent of various components of a multicomponent mixture; dynamics of adsorption, mass-transfer coefficients and adsorption mechanism; thermal consideration, and attrition expected. Direction is given for methods of obtaining multicomponent adsorption data and adsorption dynamics. The main thrust of the research recommendation is a plan to build a 6" diameter tower which can be used in a pilot plant to obtain adequate data to minimize the risk involved in design of hypersorption processes.

2. INTRODUCTION AND SCOPE

Preparation of a gaseous feed to a Fischer-Tropsch type plant, from gas produced by an oxygen blown gasifier, requires gas cleaning and gas separation to attain a feed gas with a H_2/CO molal ratio of 2:1. Presently, gas separation would be accomplished using cryogenic distillation. Moving bed adsorption is another technique for both gas cleaning and separation. Soddy (1) in the early 1920's patented a moving bed adsorber to purify coal gases. This technology was revived by Clyde Berg (2) after the second world war. Several patents were obtained and a flurry of activity was soon evident. The process technology was renamed hypersorption. Altogether six large scale hypersorption units were built in the late 1940's, only sparse data are available on their performance and operation.

The intent of the present program is to evaluate and determine the feasibility of using hypersorption process technology for the separation of certain components of a medium - BTU gas, derived from oxygen blown gasification of coal, into a high purity synthesis gas with a H_2/CO mole ratio of 2:1. Economic comparisons are to be made for hypersorption treatment of gases from oxygen-blown gasifiers versus conventional acid gas purification and cryogenic separation to produce a high purity synthesis gas under comparable conditions. Comparisons are also made in total plant investment, unit operating charges and cost of conventional gas purification and separation as opposed to hypersorptive separation and purification.

2.1 TECHNICAL EVALUATION

Several factors are to be considered in the technical evaluation.

- evaluate the adequacy of the adsorptivity data for separation of gaseous components in the applicable pressure range and desorption temperatures

- develop a processing scheme that can separate and purify the gaseous components

- assess available data on attrition losses of activated carbon adsorbents

- specify a small pilot plant scale unit for proving the process feasibility of separating gases by hypersorption

- recommend areas of research required to reduce the risk of applying hypersorption technology to gasifier effluent as a separation technique

2.2 ECONOMIC EVALUATION

The scope of the economic assessment is to consider producing a purified synthesis gas with a H_2/CO ratio of 2:1 and a minimum pressure of 330 psi, for direct feed to a Fischer-Tropsch type plant for the production of liquid fuels. The following are to be accomplished:

- compare plant investment of the oxygen-blown gasifier/oxygen plant/shift/hypersorption system versus oxygen-blown gasifier/oxygen plant/shift/conventional acid and gas purification/cryogenic separation

- compare unit operating costs for the production of the Fischer-Tropsch feed gas from the hypersorption plant versus the conventional purification and cryogenic separation plant

- assess the cost difference of eliminating the conventional acid gas purification and cryogenic separation and replacement with a hypersorption system of purification and separation.

2.3 GENERAL CONSIDERATIONS

Economic evaluations and assessments are based upon three different gasification processes: Lurgi, moving bed, oxygen-blown gasifier reported in EPRI AF-244; Texaco, slurry fed, entrained bed, oxygen-blown gasifier and Foster-Wheeler entrained bed, oxygen-blown gasifier both reported in EPRI AF-642.

The published costs studies are to be used insofar as possible for all systems assessed, with the exception of the hypersorption system, CO shift, acid gas purification and cryogenic separation. Published costs were not to be updated to a common base year. However, operating costs could not be determined on that basis. Thus escalation of previous investments were prepared. Plant capacities and gasifier effluent are to be identical to the published EPRI reports.

Plant designs are to be limited to schematic block flow diagrams except for the hypersorption system, in which the number and size of the units are to be indicated, overall heat and material balances are to be included. A process flow diagram for the hypersorption system is to be provided.

3. TECHNICAL CRITERIA

The basic technical feasibility of adsorption processes for separating gas has been shown by the various processes in operation at the present time. Moving bed adsorption using activated carbon was attempted on an industrial scale in the late 1940's. Technical feasibility had been undoubtably considered when these facilities were built. Whether these separation units were economically viable is in doubt, because they did not have an extended life. Minimal information has been published concerning their demise.

A general literature survey has been made for the present study, and a means of determining the hypersorption design are to be discussed. Various assumptions in the design and problems that had been anticipated in carrying out these designs are addressed. Required data and assumptions in the design criteria are elucidated. Research recommendations aimed at minimizing the risk of applying this technology to the separation of medium - BTU gas process streams are presented in another section.

3.1 LITERATURE SURVEY

The first continuous moving bed adsorbent was invented by F.D. Soddy in 1922 (1). The operating principles are identical to those of the modern version, but in technical details the differences are real. Soddy's patent was soon followed by one from the Metallbank Gesellschaft in 1924 (3). A notable difference between the two is the means by which desorption was performed. The next patent by the Carbo Norit Union (4) in 1931 did not contain anything new when compared with Soddy's and Metallbank Gesellschaft's patents. It is also worthwhile to mention an interesting process invented by Vollmer (Lurgi Geselschaft) (5). In general, it consists of four treating chambers and two or more adsorbent cycles, which are independent of each other. This process, for which a high degree of separtion was reported, is claimed to be very versatile, inasmuch as desorption can be achieved by continuous, discontinuous, or extractive methods. It requires, however, a great deal of supervision, and the strong movement of the adsorbent results in excessive attrition.

A renaissance of Soddy's original process came in 1946, when C. Berg, working for the Union Oil Company of California, published results on a moving charcoal bed (2). The process itself was renamed hypersorption. This name has been persistently carried through technical writings, and soon after textbooks referred to hypersorption as a new unit operation. At the time, the article by C. Berg excited worldwide interest, which had languished for years. In the United States it resulted in six large-scale hypersorber units constructed and built by the Foster-Wheeler Corporation, around 1947 and 1949. At the present time, only sparse data (6) are available on their operating conditions. There is no information in the literature of any new units having been built since the late 1940's.

Despite the general interest in the process with its capability of performing sharp separation, and at a time when which chromatography was making its entry into laboratories, the process has been overlooked after the initial flurry of activity. The possibility of matching the sharpness of chromatography, led

Szirmay (7) to achieve the separtion of acetylene from carbon dioxide in a pilot plant. Szirmay, obtained C_2H_2 in its purest state as the bottom product, with CO_2 as the top product, using activated carbon as the adsorbent.

Very little data are available in the open literature pertaining to adsorption isotherms of gases contained in a medium BTU gas obtained from an oxygen blown gasifier. Othmer and Sawyer (8) correlated adsorption data for C_2H_5Cl, n-C_5H_{12}, C_6H_6, CH_3OH, CH_3OCH_3, CO, CS_2 and NH_3. These data are for individual gases. Adsorption isotherms for single gases can also be obtained from the literature supplied by various activated carbon manufacturers. Multicomponent data is virtually impossible to find in the literature, with the exception of the Lewis' et. al (9) publication. In commercial practice, multicomponent mixture adsorption data will be required, particularly if the system exhibits non-ideality, such as azeotropic behavior. In the present work, due to the lack of available data in the literature, proprietary information from the Calgon Corporation has been used. These data consisted of adsorption isotherms for single components and N_2-CO bicomponent adsorption distribution and the heats of adsorption. The most comprehensive review of moving bed adsorption processes has been reported by Szirmay (10).

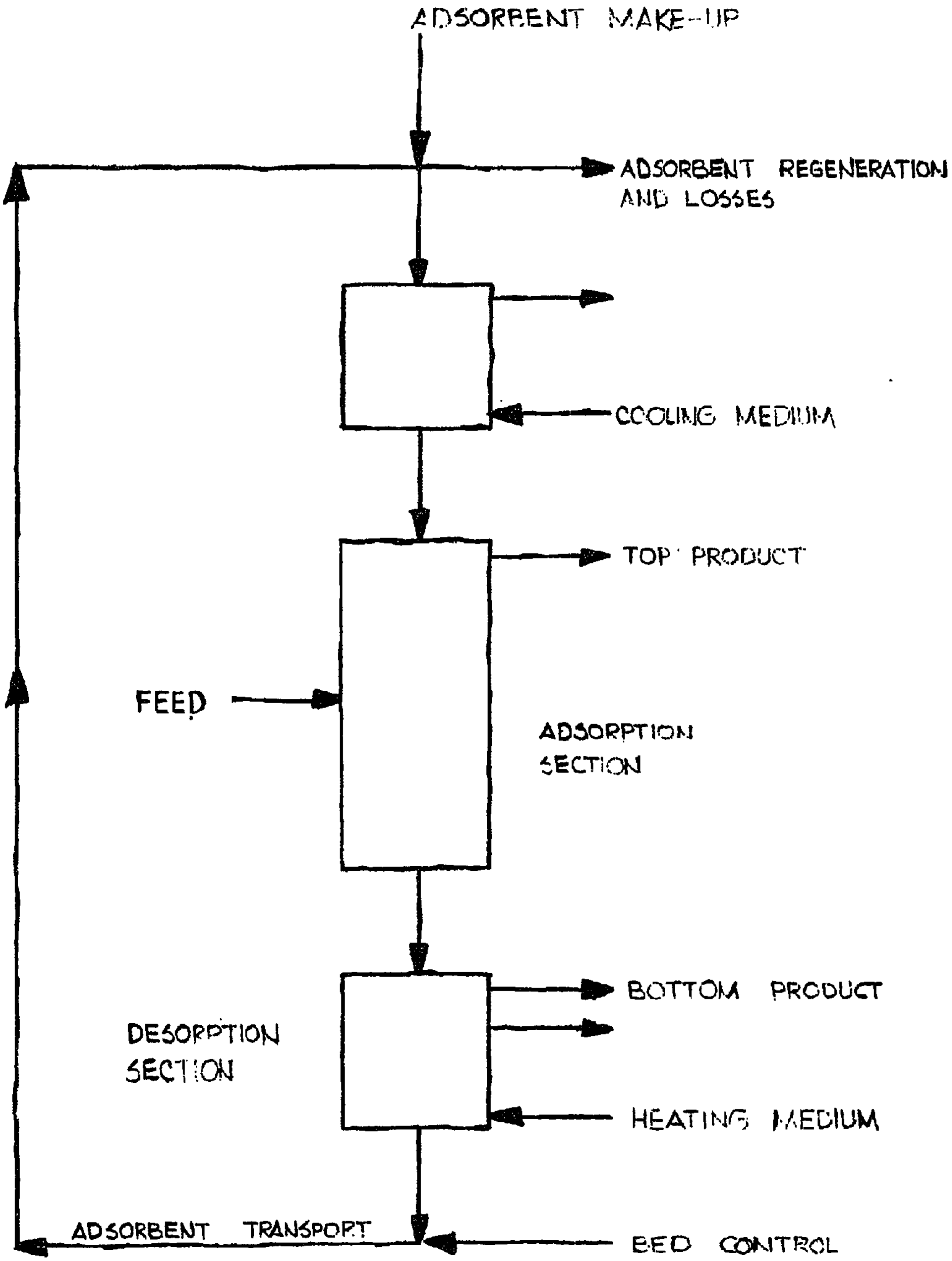
3.2 HYPERSORBER DESIGN CRITERIA

The moving adsorbent bed technology (hypersorption) is a feasible means to separate gases. Its function can be followed with the aid of Sketch, SK-1. The adsorbent moves down the column by gravity, controlled at the top of the column by a feeder mechanism and returned to the top of the column by means of solid transport. Gaseous feed enters the middle of the column. Above the feedline, the less adsorbing components are discharged at the top. Control can be maintained by initially flooding the column with the most adsorbent material. Those components that are not adsorbed as rapidly, enter into a competition with the more highly adsorbed components and therefore exit at the top of the column. The highly adsorbed components move down the column with the adsorbent. At the bottom of the column, desorption is performed by heating the adsorbent until the desorption temperature of the adsorbed components is reached and the adsorbed components are refluxed to the top of the desorption section. Cooling of the adsorbent is usually performed at the top of the column.

The moving adsorbent process has some outstanding features, one of the most important being the case of performing sharp separations, that can be compared with chromatography. Kehde, et.at (6), has reported that only 5 ft. is required to separate ethylene from a feed containing eight other gases. Thus, the height of the adsorption section is short. Unlike cryogenic separation which is limited in operation because of the critical temperature and pressure of the components, hypersorption operating conditions can be quite versatile, restricted in operating pressure primarily by azeotropic formations.

Design of hypersorption units was initially based upon equilibrium stage models. These have proved fallacious. Loading can be obtained from equilibrium data but not rates. A model of the adsorption process from a fluid stream should contain:

- diffusion of the components external to the adsorbent



SCHEMATIC OF HYPERSORPTION TOWER

DRAVO
AFC-3283 SK-1

- diffusion through the pores of the adsorbent internal to the particle
- adsorption phenomena itself

All three steps in the model contribute to the overall rate. Several mathematical solutions are available, the latest by Rusmuson (11).

The design of the hypersorption units for this task is based on available data found in the literature. In addition, single isotherms were provided by the Calgon Corporation for the following gases: Ar, N_2, CO, CO_2, CH_4, C_2H_4, H_2S and COS as well as N_2-CO bicomponent data. These data are proprietary and cannot be disclosed.

The operating temperature was originally set at 100°F to eliminate heating the feed gas to the hypersorption train as the gas leaving the gas cooling sections is approximately 100°F. A lower temperature would also require a refrigerant cooling system for the activated carbon cooldown cycle.

The operating pressure was established by trial and error within the following two constraints: first, to meet the desired separation of groups of components or individual components from each other, secondly to stay at or below the pressure where such separation is feasible. For example: effective and complete separation of CH_4 from CO could not be accomplished above 180 psia. Above this pressure an azeotrope-like situation would be created with unpredictable result. One constraint on the operating pressure was to eliminate the possibility of having to recompress the gas between hypersorbers.

Activated carbon circulation rates are based on the rates and composition of the expected bottom products assuming adsorption equilibrium (adsorption load) of an ideal mixture at said pressure and temperature (8).

The column diameter is based on the rates and composition of the feeds. The design point is at 80 percent of the superficial gas velocity at the onset of fluidization. Here, the Ledoux equation was used with a modified coefficient based on the data of Calgon (12). If sufficient area for heat transfer is not available in the desorption section of the column at the design point, column diameter is then determined by the area required for direct heating of the activated carbon.

The total length of the hypersorption column, in general, for the case of indirect heating would consist of the length of the adsorption and desorption-rectification sections, and the cooling section if included within the tower. In the adsorption section equilibrium conditions are considered prevailing, but the "break through curve" actually has dynamic characteristics. The rectification section design is strictly governed by a dynamic adsorption mechanism. At the present time dynamic adsorption data exists only for the ethane-ethylene separation (13) indicating an extremely short column length of less than one foot.

In the present case, the advantages of direct heating in the rectification section are realized. Therefore, the desorption and rectification are performed in one step by recirculating the heated bottom product. Since direct heating for desorption requires a greater column length than rectification, the design length of this section is based on the heat transfer rate between the activatd carbon and the recirculating bottom products. The adsorbents are cooled in the storage bins of the adsorbent recirculating system. As dynamic data is not available, the length of the adsorption section is based on the ethane-ethylene separation analogy (13) which indicated length of less than one foot. The length used, five feet, includes contingencies for radial and longitudinal diffusion.

In all cases, the temperature of adsorption is based on CO or CO_2 as the pure or the key component in the desired bottom products. The temperature of desorption was determined from the CO_2 and CO adsorption isoteres, such that only 90 percent of the total adsorbed gases will be desorbed. It is not expected that at this rate of desorption, the purity of the top products would be affected.

Heat load for desorption consists of two parts: one is to heat the adsorbent to the desired desorption temperature, the other is to supply the necessary heat of desorption. For the sensible heat, the specific heat of the activated carbon was provided by Calgon Corporation, whereas the heat of desorption was obtained from the literature of Brunauer and Courty (14 and 15) and verified from the respective isoteres through the Clausius-Clapeyron equation.

Activated carbon regeneration and make-up are based on the work of Kehde, et. al. (6), who reported an attrition rate of 0.0002315 lb. of adsorbent lost per pound of adsorbent circulated. At that time they predicted a reduction to one fourth of the above value for harder adsorbents. Considering the present case, with a much lower column height, much harder adsorbent and without the destructive action of the original carbon feeder and high velocity pneumatic adsorbent recirculation system, further reduction maybe expected. A conservative value of 0.00012 lb. lost/lb. circulated is chosen as the attrition rate. During the 7 months of operation, Kehde and coworkers did not observe any loss of activity of the adsorbent used. For conservative design about 20 percent of the adsorbent loss, 0.000024 lb. regenerated/lb. circulated, is used for activity loss.

A dense phase transport system is employed in the design of the activated carbon recirculating system. Low velocity should decrease the attrition losses. However, due to the state of the art, the maximum material that can be transported in one 12" transport line, according to the vendors, in dense phase flow is 150 tons/hr. Four transport lines per hypersorber are considered the maximum allowable, thus 600 ton/hr. (1.2 x 10^6 lbs./hr.) would be the greatest flow rate through any adsorber.

3.3 DESIGN BASIS

Gas composition entering the hypersorption separator or the acid gas removal system of the cryogenic separation processes after gas cooling are based upon EPRI Report AF-244 for the Lurgi Oxygen-Blown Gasifier and EPRI Report AF-642 for the Texaco, Slurry Feed, Oxygen-Blown Gasifier and the Foster-Wheeler Oxygen-Blown Gasifier. The gasifier effluents are shown in Tables 3-1.

Product output is to contain the maximum quantity of gas possible with a H_2/CO molal ratio of 2:1. Excess H_2 is burned in the boiler and excess CO is shifted by the water gas reaction to form H_2. The product is to be fed to a Fischer-Tropsch Type Plant at 340 psig and 460°F. Maximum impurities are listed in Table 3-2.

TABLE 3-1

COMPOSITION OF GASIFIER EFFLUENT TO BE PROCESSED

Gas Composition	Lurgi	Texaco	Foster-Wheeler
		lb. moles/hr.	
CH_4	6,464.0	72.0	5,383.2
C_2H_4	192.0		
C_2H_6	293.6		
H_2	31,590.8	25,974.6	26,306.7
CO	11,315.2	38,236.5	31,498.5
CO_2	23,077.2	7,944.5	11,154.3
H_2S	888.8	907.0	917.1
COS	38.8	54.0	91.5
NH_3	605.6	178.0	570.3
N_2, Ar	277.2	705.5	400.5
H_2O	121,145.6	23,640.0	12,841.5
Total	195,888.8	97,712.1	89,163.6

TABLE 3-2

STREAM PURITY FOR FEED TO FISCHER-TROPSCH TYPE PLANT

H_2	2% INERTS	2mg/m^{3}S
CO	2% INERTS	2mg/m^{3}S
H_2S-COS	MAXIMUM RECOVERY	PURITY NOT CRITICAL
CO_2	10 ppm S	
N_2	10 ppm S	
MIXED FUEL (as CH_4)	400 ppm S	

4. PROCESS DESCRIPTIONS

Technical evaluation of the hypersorption process demands a process scheme be developed and analyzed. The process developed is based upon the best data available, good engineering judgment and is consistent with the budget allotted to this study. The process scheme chosen for the application of hypersorptive technology is one of several configurations possible. No attempt has been made to optimize the process scheme. Some alternative studies were carried out when deemed necessary.

Information on cryogenic processes has been obtained from commercially available sources. Process equipment and methods prior to the hypersorption separation - purification or the cryogenic separation - acid gas removal are based upon EPRI Report AF-244 for the Lurgi cases and EPRI Report AF-642 for the Texaco and Foster-Wheeler cases.

Six separate cases are described consisting of hypersorption and a cryogenic process for each of the three gasification systems. Similar equipment and processing methods are employed. Each major section includes a general description, gasifier effluent gas composition, a process description for the hypersorption process, a list of equipment, (the equipment changes are referred to the corresponding EPRI reports), and an overall energy balance and steam utilization.

4.1 PROCESS DESCRIPTION: (LURGI GASIFICATION CASE)

The coal receiving and storage, coal conveying, gasification and gas cooling areas are the same as described in the EPRI Report AF-244 except minor changes were made as reported elsewhere in this section.

Gasifier effluent from the gas cooling area is the feed for the hypersorption or acid gas removal area of the cryogenic separation and purification area of the plant. The gas composition is shown in Table 4-1.

4.1.1 HYPERSORPTION SEPARATION AND PURIFICATION

A Block Flow Diagram and a material balance for the Lurgi, hypersorption separation and purification case, AFC-3283-100-001, is presented as an aid in understanding the following discussion. A hypersorption Process Flow Diagram AFC-3283-100-002 is also available.

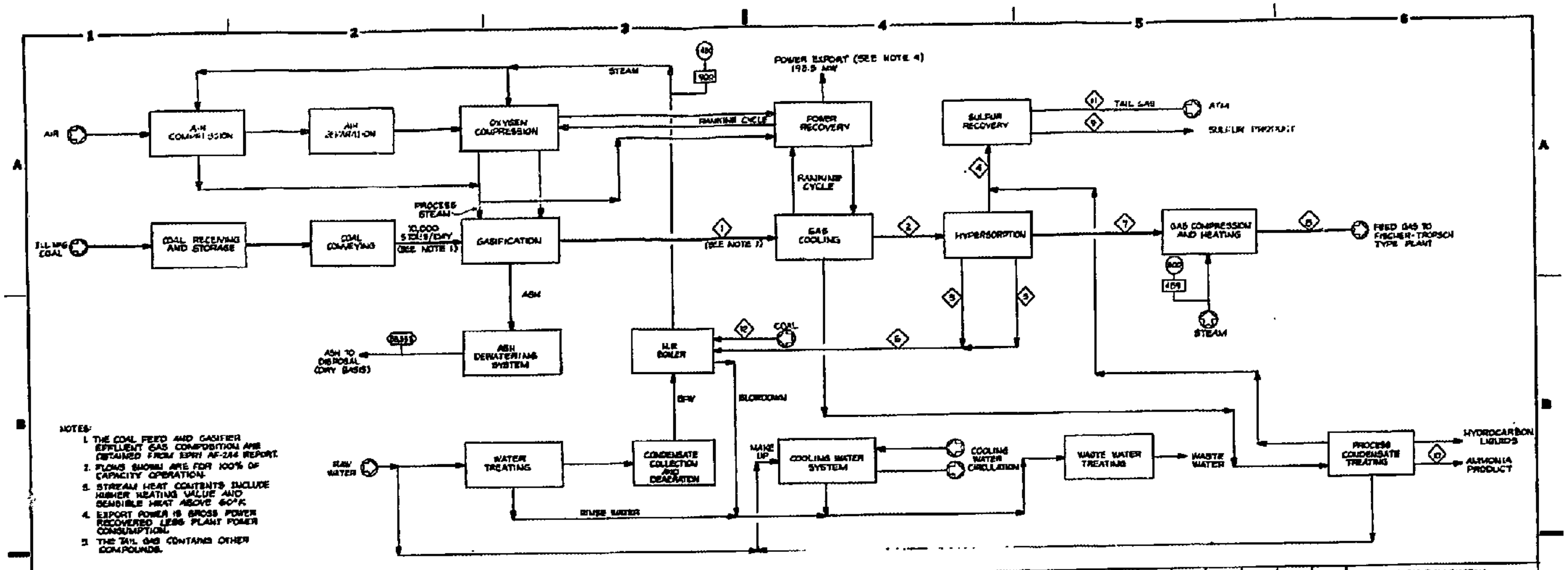
The hypersorption gas separation and purification is based on the single component isotherms provided by the Calgon Corporation for type PCB granular activated carbon. Pressure and temperature required for the adsorption are based on the gas components to be separated. In the present case, a pressure of 165 Psig and a temperature of 100°F have been selected. At these conditions, effective and complete separation of CH_4 from CO and H_2 can be achieved.

Hypersorber 1: Separates H_2 and CO from other gas components

Hypersorber 2: Separates H_2 from CO

Hypersorber 3: Separates CH_4 from acid gas

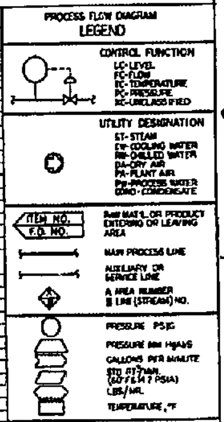
In most cases, very sharp separation of the gas components can be achieved.



NOTES:

1. THE COAL FEED AND GASIFIER EFFLUENT GAS COMPOSITION ARE OBTAINED FROM EPRI AF-244 REPORT.
2. FLOWS SHOWN ARE FOR 100% OF CAPACITY OPERATION.
3. STREAM HEAT CONTENTS INCLUDE HIGHER HEATING VALUE AND SENSIBLE HEAT ABOVE 60°F.
4. EXPORT POWER IS GROSS POWER RECOVERED LESS PLANT POWER CONSUMPTION.
5. THE TAIL GAS CONTAINS OTHER COMPOUNDS.

MATERIAL / STREAM No.	1	2	3	4	5	6	7	8	9	10	11	12
LB MOLES/HR	GASIFIER EFFLUENT	CRUDE GAS	EXCESS H_2	ACID GAS	METHANE	FUEL GAS	CO AND H_2	PRODUCT GAS	SULFUR PROD.	AMMONIA PROD.	TAIL GAS (NOTE 5)	COAL
CH_4	6464.0	6464.0			6464.0	6464.0						
C_2H_4	192.0	192.0		192.0								
C_2H_6	293.6	293.6		293.6								
H_2	51590.8	51590.8	8960.4			8960.4	42630.4	42630.4				
CO	11315.2	11310.2					11319.7	11319.2				
CO_2	23077.2	23077.2		23077.2							4.4	
H_2S	868.8	868.8		868.8								
COS	58.8	58.8		58.8								
SULFUR									832.1			
NH_3	605.6									605.6		
N_2, Ar	271.2	271.2		138.6			138.6	138.6				
H_2O	121145.6	268.0		268.0								
COAL LBS/HR												236044
TOTAL LB MOLES/HR	215388.0	94408.6	8960.4	24891.0	6464.0	15424.4	54084.2	54084.1	851.1	605.6	4.4	
TOTAL LBS/HR	3747912	1591189	18064	1071178	103700	121764	366406	366446	26650	10315	150	236044
AVE. MOL. WT.	19.13	20.96	2.016	43.02	16.043	7.894	10.751	10.751	32.064	17.03	34.08	
SCFM (14.4 PSIA, 60°F)	1364134	460201	51829	160681	40116	97541	187974	187974				
PRESSURE, PSIG	502	242	165	165	165	165	165	540				
TEMPERATURE, °F	380	100	100	105	100	100	100	460				
ENTHALPY, MM BTU/HR	2140.15	28.495	2.906	14.235	2.344	4.846	3.475	65.048				



PRELIMINARY NOT APPROVED FOR CONSTRUCTION

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DRAWN	D.JANTACK	2-15-82
CHECKED	R. ROODMAN	3-26-82
APPROVED	L. SAROFF	

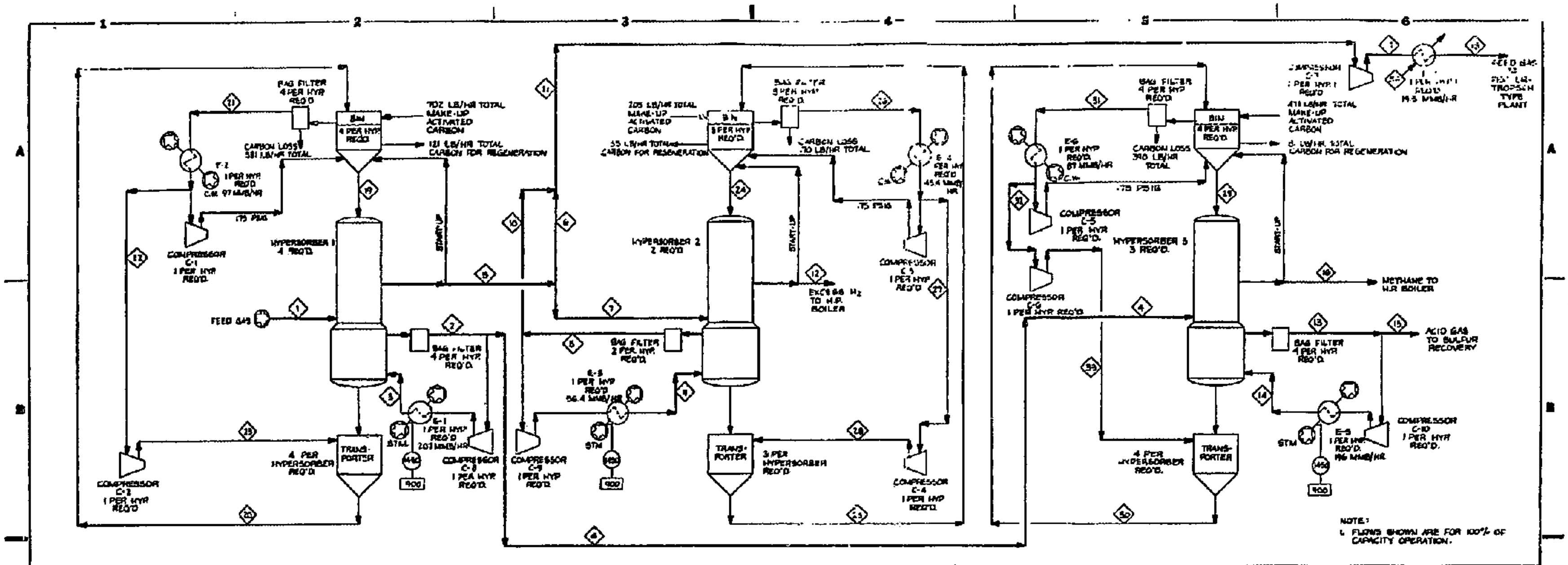
PROCESS BLOCK FLOW DIAGRAM
LURGI MOVING BED COAL GASIFICATION
OXYGEN BLOWN-HYPERSORPTION SEPN.

DE-AC21-81 MC 16447

Dravo Engineers and Constructors

Pittsburgh Pennsylvania

Contract AFC-3283 Drawing No. 100-001 Rev. A



NOTE:
1. FLOWS SHOWN ARE FOR 100% OF CAPACITY OPERATION.

MATERIAL \ STREAM No.	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17
LB MOLES/HR	CRUDE GAS	BOTTOM PRODUCT	BOTTOM PRODUCT RECYCLE	ACID GASES	TOP PRODUCT	PRODUCT GAS	HYP 2 GAS FEED	BOTTOM PRODUCT	BOTTOM PRODUCT RECYCLE	CO PRODUCT	FEED TO COMP	EXCESS H_2	BOTTOM PRODUCT	BOTTOM PRODUCT RECYCLE	ACID GASES	METHANE	FEED TO EXCH
CH_4	6464.0	30179.7	23714.7	6464.0												6464.0	
C_2H_4	192.0	1130.7	938.7	191.0									1026.5	834.5	192.0		
C_2H_6	293.6	1740.1	1446.5	293.6									1672.2	1378.6	293.6		
H_2	31590.8				31590.8	22630.4	8960.4				22630.4	8960.4					22630.4
CO	11315.2				11315.2	8105.8	3209.4	26405.4	23196.0	3209.4	11315.2						11315.2
CO_2	25077.2	156317.9	132403.5	25077.2									125946.1	100868.9	25077.2		
H_2S	888.8	5259.0	4370.2	888.8									4757.3	3868.5	888.8		
COS	58.8	323.4	264.6	58.8									712.2	653.4	58.8		
N_2, Ar	277.2	835.6	677.0	158.6	158.6	99.5	59.5	323.3	264.0	59.3	158.6		745.4	606.8	138.6		138.6
H_2O	268.0	1536.0	1180.0	268.0									1427.5	1159.5	268.0		
ACTIVATED CARBON LBS/HR																	
TOTAL LB MOLES/HR	74405.6	405301.5	[illegible]	35961.0	43044.6	30835.5	12209.1	26118.7	23480.0	3148.7	34084.2	8960.4	[illegible]	[illegible]	24971.0	6464.0	34084.2
TOTAL LBS/HR	1559588	[illegible]	[illegible]	1146117	384510	275448	109062	146572	657475	90998	366446	18064	5731026	4661647	1071178	103700	366446
AVE. MOL. WT	20.96	31.46	31.46	31.46	8.933	8.933	8.933	28.01	28.01	28.01	10.75	2.016	43.02	43.02	43.02	16.043	10.75
SCFM (14.4 PSIA, 60°F)	460701	[illegible]	[illegible]	[illegible]	177004	[illegible]	72796	111505	[illegible]	20967	211714	51829	[illegible]	[illegible]	[illegible]	41718	219974
PRESSURE, PSIG	165	165	170	165	165	165	165	165	170	165	165	165	165	170	165	165	340
TEMPERATURE, °F	100	105	700	105	100	100	160	105	700	105	100	100	105	700	105	100	220
ENTHALPY, MM BTU/HR	26.475	41.141	909.11	16.519	16.925	6.547	3.583	7.261	119.179	0.863	9.425	1.500	76.000	571.775	14.225	2.346	51.796

MATERIAL \ STREAM No.	18	19	20	21	22	23	24	25	26	27	28	29	30	31	32	33	34
LB MOLES/HR	FINAL PRODUCT	CARBON TO HYP 1	CARBON RECYCLE	GAS FROM BIN HYP 1	GAS TO COMP HYP 1	GAS TO TRANS HYP 1	CARBON TO HYP 2	CARBON RECYCLE	GAS FROM BIN HYP 2	GAS TO COMP HYP 2	GAS TO TRANS HYP 2	CARBON TO HYP 3	CARBON RECYCLE	GAS FROM BIN HYP 3	GAS TO COMP HYP 3	GAS TO TRANS HYP 3	STEAM
CH_4														123810.0	36450.0	36450.0	
C_2H_4																	
C_2H_6																	
H_2	22630.4			150145.0	35374.0	35374.0			65773.0	16073.0	16073.0						
CO	11315.2			54000.0	12672.0	12672.0											
CO_2																	
H_2S																	
COS																	
N_2, Ar	138.6			657.0	154.0	154.0											
H_2O																	4360.6
ACTIVATED CARBON LBS/HR		4842640	4842640				1412720	1412720				3247640	3247640				
TOTAL LB MOLES/HR	34084.2			205400.0	48100.0	48200.0			65713.0	16073.0	16073.0			123840.0	36450.0	36450.0	4360.6
TOTAL LBS/HR	366446	4842640	4842640	1834843	430517	430517	1412720	1412720	133544	34435	34435	3247640	3247640	1986234	579954	579954	78556
AVE. MOL. WT	10.75			8.933	8.933	8.933			2.016	2.016	2.016			16.043	16.043	16.043	18.015
SCFM (14.4 PSIA, 60°F)	219974			825617	311075	311075			411680	16640	124660			799050	233306	233306	28143
PRESSURE, PSIG	340			165	145	245			165	45	245			165	145	145	600
TEMPERATURE, °F	460			347	85	85			282	85	85			328	85	85	489
ENTHALPY, MM BTU/HR	95.040			401.361	3.336	3.338			91.250	1.261	1.261			142.104	3.242	3.242	57.257

PROCESS FLOW DIAGRAM
LEGEND

CONTROL FUNCTION
LC-LEVEL
FC-FLOW
TC-TEMPERATURE
PC-PRESSURE
XC-UNCLASSIFIED

UTILITY DESIGNATION
ST-STEAM
CW-COOLING WATER
RW-CHILLED WATER
DA-DRY AIR
PA-PLANT AIR
PW-PROCESS WATER
COND-CONDENSATE

ITEM NO. / F.D. NO. — RAW MAT'L OR PRODUCT ENTERING OR LEAVING AREA
MAIN PROCESS LINE
AUXILIARY OR SERVICE LINE
A AREA NUMBER
B LINE (STREAM) NO.
PRESSURE PSIG
PRESSURE MM HgABS
GALLONS PER MINUTE
STD FT³/MIN. (60°F & 14.7 PSIA)
LBS/HR.
TEMPERATURE, °F

PRELIMINARY NOT APPROVED FOR CONSTRUCTION

A.K.L-ALI 2-16-82
DJFRCHAK 2-24-82
NUKE-HAN 3-26-82
L. SNYDER

HYPERSORPTION PROCESS FLOW DIAGRAM
LURGI MOVING BED COAL GASIFICATION
OXYGEN BLOWN - HYPERSORPTION SYSTEM

DE-AC21-81MC 16447

Dravo Engineers and Constructors

Pittsburgh Pennsylvania

Dravo

AFC 3-83
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TABLE 4-1

GAS COMPOSITION FOR LURGI CASE, FEED TO SEPARATION PROCESSES

Compound	Moles Per Hour	Mole Percent
CH_4	6,464.0	8.69
C_2H_4	192.0	0.26
C_2H_6	293.6	0.39
H_2	31,590.8	42.46
CO	11,315.2	15.21
CO_2	23,077.2	31.02
H_2S	888.8	1.19
COS	38.8	0.05
N_2	277.2	0.37
H_2O	268.0	0.36
Total	74,405.6	100.0

The activated carbon is fed at the top of the hypersorber through four 12" lines. Activated carbon circulation rates are based on the rates and composition of the expected bottom products assuming adsorption equilibrium of an ideal gas mixture. Hypersorber height is composed of several sections; adsorption section above the feed, redistribution of the activated carbon and desorption section at the bottom.

A 60^0 angle of repose was used to calculate the height of activated carbon distribution. Volume for the feed mechanism is provided at the top of the hypersorber. The design of the activated carbon feed to the hypersorber is conceptual and detail design is beyond the scope of this study. Heights of adsorption and desorption sections are the same, 5 ft. For activated carbon redistribution at the bottom, 5 ft. is allowed.

A futher limitation on the carbon flow rate is the commercial availability of a transport system.

The hypersorber diameter design is based on 80% of the fluidization velocity. The capacity of the largest commercially available dense phase conveying system available is 150 tons/hr. for the activated carbon circulation. Therefore, the maximum amount of activated carbon feed to the hypersorber is 600 tons/hr. (1,200,000 lb./hr.) as four transporters and four bins are provided for one hypersorber.

Hypersorber 1: The gas from gas cooling area is fed to the Hypersorber 1 at 165 Psig and 100^0F. The activated carbon cooled to 100^0F is fed to the top of Hypersorber 1. As activated carbon descends it adsorbs all gases except H_2 and CO. Adsorbed gases are withdrawn as bottom product, while H_2 and CO are withdrawn as top product. N_2 in the feed gas is evenly distributed between the top and bottom products. Isotherms of Co and N_2 are nearly identical, the split could not be determined with confidence.

Adsorbed gases on the activated carbon are desorbed at

412^{o}F in the desorption section of the Hypersorber 1. The activated carbon is heated to 412^{o}F by recycling bottom product gas at 700^{o}F. Compressed bottom product gases are heated to 700^{o}F in the heat exchanger E-1* by high pressure steam.

The activated carbon at 412^{o}F is collected in the transporter located at the bottom of Hypersorber 1 and then conveyed to the bin at the top of Hypersorber 1 by using a gas medium of top product composition.

During start-up, top product gas inventory is accumulated to provide conveying gas for the recirculation loops. Top product gases flow from the bottom to the top of the storage bin to cool the hot activated carbon. Fines from the top product gas are removed in the bag filter. This gas is cooled to 85^{o}F in cooler, E-2. Part of the gas is recycled to the bin after compressing to 175 Psig in compressor, C-1, which has an aftercooler to produce gas at 100^{o}F. The remaining gas, compressed to 245 Psig in compressor, C-2, is used to transport the hot activated carbon to the bin, completing a cycle.

The solids in the bottom product gas are removed in a bag filter. Part of bottom product gas is compressed to 170 Psig in the compressor, C-8, and is heated in exchanger, E-2, as described previously. The remaining portion of the bottom product gas is fed to Hypersorber 3 to separate CH_4 from the acid gases.

Hypersorber 2: The top product gas from Hypersorber 1 containing H_2 and CO is fed to Hypersorber 2 to separate the H_2 from the

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* Equipment numbers refer to AFC-3283 flow diagram equipment items. Numbers in parentheses refer to the EPRI reports.

CO. Activated carbon at 100^{o}F adsorbs CO gas leaving H_2 as the top product gas. H_2 in excess of the desired H_2/CO ratio is burned in the high pressure boilers.

The desorption temperature for CO gas is 346^{o}F. Gas heating, carbon cooling and transport are similar to that described for hypersorber 1.

Hypersorber 3: The bottom product gas from the Hypersorber 1 at 165 Psig and 105^{o}F is fed to the Hypersorber 3, where CH_4 is separated as the top product gas and acid gas as the bottom product gas. Desorption of the acid gas takes place at 412^{o} F.

Tables 4-2 to 4-5 are lists of equipment required for the Lurgi Hypersorption separation and purification. Diameters of the desorption sections are based on the heat transfer area required for desorption, while the top section diameters are designed for 80 percent of the fluidization gaseous velocity. Transporter system, supply vessels, bin bag filters, and controls are a packaged item. Thirty-two bins, 12' diameter x 30' high, are also required.

Part of the bottom product CO from the Hypersorber 2, is mixed with the balanced part of the top product H_2 and CO from the Hypersorber 1 to make H_2/CO ratio 2:1 for the feed gas to Fischer-Tropsch type plant. The product gas mixture is compressed from 165 Psig to 340 Psig in compressor, C-7, and heated to 460^{o}F in heat exchanger, E-7, using 600 Psig saturated steam. Gas at 340 Psig and 460^{o}F is the feed gas for the Fischer-Tropsch type plant for further synthesis.

A small quantity of the activated carbon is withdrawn from the bins for regeneration. Activated carbon make-up added into the bins is the total of the carbon withdrawn for regeneration and attrition losses. Regeneration is done off-site.

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TABLE 4-2

HYPERSORBERS: LURGI CASE

Equipment Name	Quantity Required	Size
Hypersorber 1		
Adsorption section	4	13' dia. x 19' high
Desorption section	4	20' dia. x 10' high
Hypersorber 2		
Adsorption section	2	6' dia. x 13' high
Desorption section	2	12' dia. x 10' high
Hypersorber 3		
Adsorption section	3	12' dia. x 18' high
Desorption section	3	20' dia. x 10' high

TABLE 4-3

BAG FILTERS AT HYPERSORBERS: LURGI CASE

Bag Filter At	Quantity Required	Design SCFM Each
Hypersorber 1	16	75,000
Hypersorber 2	4	45,000
Hypersorber 3	12	75,000

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TABLE 4-4

COMPRESSORS: LURGI CASE

Equipment No.	Quantity Required	Design SCFM	Compression Ratio P_2/P_1	BHP Required	Total BHP Required
C-1	4	258,000	1.19	4,200	16,800
C-2	4	77,760	1.63	3,600	14,400
C-3	2	148,500	1.19	2,400	4,800
C-4	2	58,300	1.63	2,700	5,400
C-5	3	205,000	1.14	3,300	9,900
C-6	3	77,760	1.63	3,600	10,800
C-8	4	250,000	1.14	3,600	14,400
C-9	2	76,000	1.14	1,100	2,200
C-10	3	235,000	1.14	3,300	9,900

TABLE 4-5

HEAT EXCHANGERS: LURGI CASE

Equipment No.	Quantity Required	Heat Duty/Exchanger MMBTU/HR.	Surface Area/Exchanger Required Ft.^{2}
E-1	4	207.0	9,100
E-2	4	97.0	11,000
E-3	2	56.4	2,500
E-4	2	43.3	6,200
E-5	3	196.0	8,700
E-6	3	87.0	11,000

CH_4, from Hypersorber 3, is burned in the high pressure boiler while the bottom product acid gas, is fed to the Sulfur Recovery Area for sulfur recovery.

The by-product sulfur is recovered at the rate of 321 tons/day using the Stretford Process.

Air compression, air separation and oxygen compression areas remain the same as described in EPRI Report AF-244 except that additional steam is required to operate the air compressor because the fuel gas for expanders is unavailable. The number of high pressure boilers have been increased to generate a greater quantity of steam. The fuel for the high pressure boilers is supplemented by coal.

Export power is greater than in EPRI Report AF-244 because of the increased amount of low pressure steam available.

The process condensate is processed through a tar oil separation unit and phenol extraction unit identical to that in EPRI Report AF-244.

The production of hydrocarbon liquids and ammonia remains the same as specified in EPRI Report AF-244.

4.1.1.1 CHANGES FROM EPRI REPORT AF-244

Product gas from the Lurgi gasifier in EPRI Report AF-244 is to be used for the production of Fischer-Tropsch synthesis gas rather than fuel for boilers. Therefore, the following changes have been made in the equipment usage.

Oxidant Feed System

Gasifier product gas is not available to run the Gas Expander, (11-1-EX-1), therefore, it is deleted. The high pressure steam (1450 Psig, 900°F) flow rate to the Air Compressor Steam Turbine, (11-1-T-1), has been increased to provide the additional power to drive the Air Feed Compressor, (11-1-C-1).

Gas Cooling

Heat exchanger, (21-1-E-1), previously used to heat the product gas from 103°F to 355°F, is now used to heat the high pressure BFW from 269°F to 297°F for the boilers.

Heat exchanger, (22-1-E-5), previously used to provide stripping steam for the regenerator in the Acid Gas Removal System, is now used to heat vacuum condensate from 109°F to 196°F.

Heat exchangers, (21-1-E-12), and (21-1-E-13), previously used to heat the demineralized water and vacuum condensate from 110°F to 220°F, are now used to heat the demineralized water from 70°F to 226°F.

Acid Gas Removal System

This area has been deleted because the acid gases are separated in the Hypersorption Area.

Gas Expansion

This area has been deleted because product gas is not available for the expansion.

Gas Compression and Heating

The product gas from Hypersorption Area is available at 165 Psig and 100°F. Fischer-Tropsch feed is required at 340 Psig and 460°F for further synthesis. A gas compression and heating area is added to compress gas from 165 Psig to 340 Psig and to heat the gas to 460°F using 600 Psig saturated steam.

Power Recovery

Three additional power recovery units have been added to recover the power from excess 340 Psig and 50 Psig steam. A total of six units are used to recover the power.

High Pressure Boiler

Six additional high pressure boilers have been added to generate the additonal high pressure steam required by the Air Compressor Steam Turbine, (11-1-T-1), and by the heat exchangers in the hypersorption area. A total of ten high pressure boilers are used to generate the required steam.

4.1.1.2 PROCESS ENERGY BALANCE

Table 4-6 presents the overall process energy balance for 100% capacity operation. The boundary for the balance encompasses the process units and steam system. The energy content of the streams crossing the boundary is expressed as the sum of the stream's higher heating value, sensible heat above 60^{0}F, and latent heat of water at 60^{0}F. Electric power is converted to equivalent theoretical heat energy at 3413 Btu/kwh. The energy balance closes to less than 1%. The discrepancies result from the approximations used for some units and from the calculations of some heat loads.

4.1.1.3 STEAM, BOILER FEEDWATER AND CONDENSATE

The steam system operates at five pressure levels:

High Pressure - 1,450 Psig, 900^{0}F
Intermediate Pressure - 600 Psig
Medium Pressure - 340 Psig
Low Pressure - 100 Psig
Low Pressure - 50 Psig

TABLE 4-6

LURGI HYPERSORPTION ENERGY BALANCE

Basis: $60^{o}F$, Water as Liquid, 3413 Btu/kwh

	MMBTU/HR					
HEAT IN	HHV	Sensible	Latent	Radiation	Power	Total
Coal	13,084	5				13,089
Air Compressor Suction Air		12	18			30
Demineralized Water		19				19
Boiler Combustion Air		40	57			97
Auxiliary Power Inputs					373	373
Total	13,084	76	75	0	373	13,608
HEAT OUT						
Ash (Dry Basis)	47	18				65
Gasifier Heat Losses				85		85
Sulfur Product	104	1				105
Ammonia Product	100					100
Naphtha, Tar, Oil, Phenol	913	1				914
Product Gas	4,160	95				4,255
Excess Steam Generator					938	938
Bottoming Cycle Generator					124	124
Steam Surface Condenser			3,330			3,330
Isobutane Condenser			745			745
Boiler Stack gas		394	832			1,226
Gasifier Effluent Trim Cooling		2	11			13
Oxidant Compressor Cooling		104				104
Tar and Phenol Process Heat		467				467
Steam Heat Losses			36			36
Process Condensate Cooling		297				297
Air Separator Power					4	4
Hypersorption Coolers		736				736
Total	5,324	2,115	4,954	85	1,066	13,544

$$\frac{\text{Output - Input}}{\text{Input}} = 0.47\%$$

High pressure steam is produced in ten boilers operating at 1,450 Psig. Boiler feedwater from the deaerator is preheated against hot process condensate from the gasifier effluent cooling area. Two heat exchangers are used to preheat the boiler feed water. Steam is superheated to 900°F. High pressure steam turbines drive the air feed compressors in the Oxidant Feed System. These are back pressure turbines exhausting at 340 Psig and 609°F. High pressure steam is also used to heat the bottom product gas recycle from the hypersorbers. Desorption of the adsorbed gas on activated carbon of the bottom product requires a moderately high temperature. High pressure condensate from the heat exchangers is flashed at 50 Psig to generate 50 Psig steam.

The intermediate pressure steam is generated in one of the high pressure boilers. This steam is used to heat the product gas to 460°F in the Gas Heating Area. Condensate at 600 psig is flashed to generate 50 Psig steam.

Medium pressure steam at 340 Psig is mainly supplied by the back pressure turbines used in the Oxidant Feed System. The balance comes from process waste heat generation jacket steam from the gasifiers and the steam generated in the Sulfur Recovery Area waste heat boilers.

The medium pressure steam header is controlled by feeding the excess steam to a power recovery turbine. Turbogenerators act as a balancing wheel to control the steam system. Swings in steam demand at the different levels are reflected on the power output of the turbogenerator. About 52% of the total medium pressure steam is supplied to the gasifiers and 38% is used for power generation. The balance is used for pump steam turbine drivers which exhaust to the 50 Psig steam level.

The 100 Psig steam is generated in the gasifer effluent waste heat boilers, (21-1-E-2). This steam is used in the Process Water Treating Unit.

The 50 psig steam supply is obtained from small back pressure turbine drivers, gasifier effluent process heat recovery, Sulfur Recovery Unit, flashing 1,450 Psig and 600 Psig condensates, and blowdown flash steam. Blowdown from the high and medium pressure steam generators are combined in a blowdown drum which vents flash steam to the 50 Psig header. The steam is used in the Process Water Treating Unit. About 75% of the steam is used to recover power.

Critical pumps, such as boiler feedwater and condensate pumps, are steam turbine driven. Other pump services considered important but not critical use electrical motors in normal operation and steam driven spares for emergency service. All turbine drivers operate from 340 Psig to 50 Psig back pressure.

Steam condensate from the various process units is flashed at 15 Psig in a hot condensate flashdrum, (30-V-3). Flashed steam is vented to the deaerator where it provides a portion of the deaeration steam. The flashed condensate is pumped to the deaerator tray section.

Raw water is treated in a semiautomatic, resin bed demineralization unit to produce demineralized water suitable for a 1500 Psig boiler system. Storage, equivalent to 24 hours of demineralized water production is provided. The raw water requirement is 3,850 gpm. Some demineralized water is also used to satisfy process requirements.

The demineralized water for the steam system is combined with vacuum condensate returned from surface condenser. The combined stream is heated to 228°F by heat exchange in the Gas Cooling Unit. The heated stream is deaerated in a tray type deaerator operating at 15 Psig.

Excess heat and steam generated in the fuel processing sections of the plant are converted to electrical power. Thus, 312.2 MW of power are generated from the process.Power is generated by the expansion of excess 340 Psig and 50 Psig steam in a steam turbine driven generator. A small amount of power is generated in the air separation plant. The rest is developed in the operation of a low level heat recovery scheme utilizing an organic fluid in a Rankine cycle.

From Steam, MW	274.8
From Low Level Heat, MW	36.3
From Air Separation,MW	1.1
	312.2 MW

A bottoming cycle using isobutane is employed to recover low temperature heat energy and convert it to by-product power. Isobutane vaporized in oxidant feed compressor coolers and gasifier effluent coolers is expanded in a turbine driving a generator (the same generator driven by expansion of steam). The expanded isobutane is condensed against cooling water and recirculated in the cycle. Isobutane surge and storage facilities are provided.

Very low level heat in the plant is rejected to cooling water. This heat comes from:

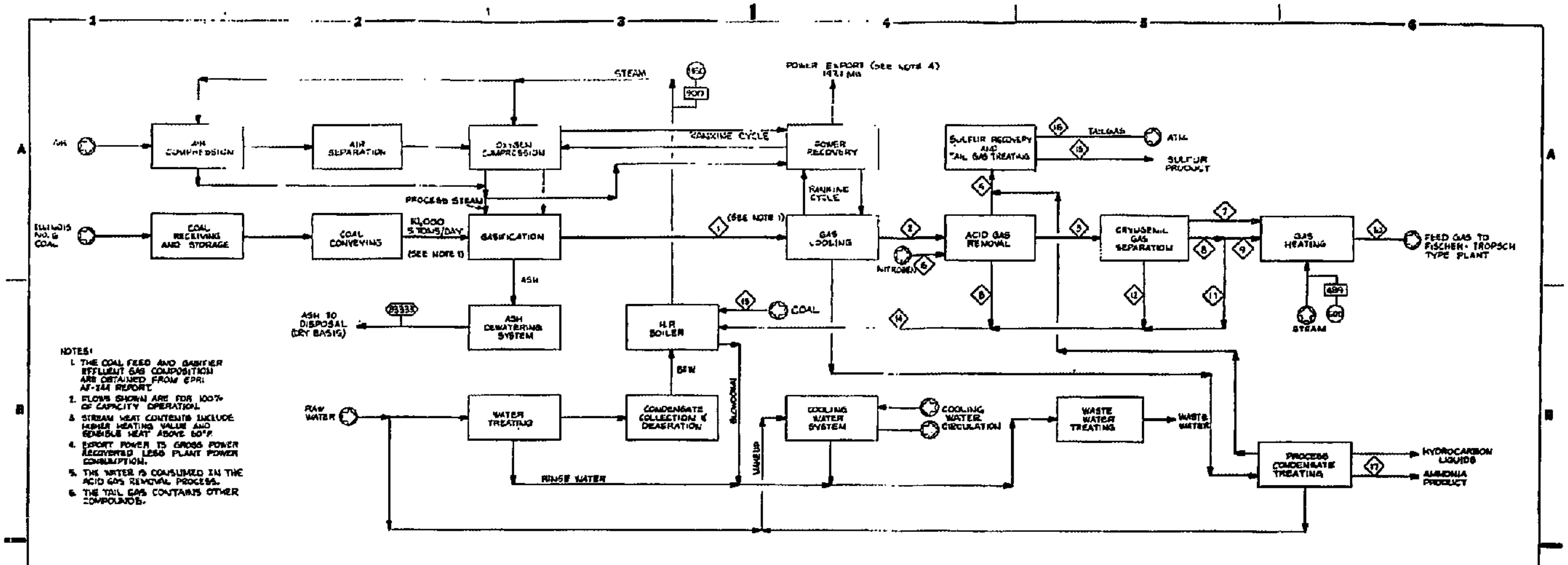
- Process cooling
- Steam turbine surface condenser
- Bottoming cycle condenser

Total heat rejection to cooling water is 5,138 MM Btu/hr.

Cooling water is a circulating system wherein the water is returned to cooling towers, cooled in the towers to 70^{o}F and pumped back to the various users at 60 Psig. The cooling towers are multicell induced draft type. The cooling water flowrate is 608,000 GPM.

4.1.2 CRYOGENIC SEPARATION AND ACID GAS PURIFICATION

The composition of the gas leaving the Gas Cooling Area is shown in Table 4-1. A Block Flow Diagram of the cryogenic system is available as drawing AFC-3283-100-003. Gas is fed to the Acid Gas Removal System at 242 Psig and 100^{o}F. The Acid Gas Removal process Rectisol of Lotepro Corporation is



NOTES:

1. THE COAL FEED AND GASIFIER EFFLUENT GAS COMPOSITION ARE OBTAINED FROM EPRI AF-244 REPORT.
2. FLOWS SHOWN ARE FOR 100% OF CAPACITY OPERATION.
3. STREAM HEAT CONTENTS INCLUDE HIGHER HEATING VALUE AND SENSIBLE HEAT ABOVE 60°F.
4. EXPORT POWER IS GROSS POWER RECOVERED LESS PLANT POWER CONSUMPTION.
5. THE WATER IS CONSUMED IN THE ACID GAS REMOVAL PROCESS.
6. THE TAIL GAS CONTAINS OTHER COMPOUNDS.

MATERIAL / STREAM No.	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17
LB MOLES/HR	GASIFIER EFFLUENT	CRUDE GAS	SWEET PRODUCT GAS	ACID GAS	FLASH GAS	STRIP N_2	CO PRODUCT	H_2 PRODUCT	H_2 TO GAS HEATING	PRODUCT GAS	EXCESS H_2	RESIDUAL GAS	COAL	FUEL GAS	SULFUR PRODUCT	TAIL GAS (NOTE 6)	AMMONIA PRODUCT
CH_4	6464.0	6464.0	6140.8		323.2		11.2	314.6	223.1	234.5	91.7	5814.8		6289.7			
C_2H_4	192.0	192.0	115.2		76.8							115.2		192.0			
C_2H_6	293.6	293.6	176.2		117.4							176.2		293.6			
H_2	31590.8	31590.8	31432.8		158.0		11.2	31275.6	22168.8	22180.0	9106.8	146.0		9410.8			
CO	11315.2	11315.2	11149.6		165.6		11090.0	1ppm		11070.0		52.6		225.7			
CO_2	25017.2	25017.2	0.5	1012.4	21064.3							0.5		21064.8			
H_2S	888.8	888.8	1ppm	888.3	1ppm											4.6	
COS	35.8	35.8		35.5													
SULFUR															852.6		
NH_3	605.8																605.6
N_2, Ar	277.1	277.2	273.2	54.2	3644.8	5100.0	100.8	92.4	87.1	251.9	15.3	30.0		3680.1			
O_2																	
H_2O (NOTE 5)	121145.4	268.0															
COAL LBS/HR													545559				
TOTAL LB MOLES/HR	175888.8	74403.6	49284.3	2999.2	25554.1	5100.0	11313.2	31642.8	22429.0	33742.2	9213.8	6312.3		41096.2	852.6	4.6	605.6
TOTAL LBS/HR	3147312	1509386	490379	122845	1049083	105402	316459	69599	49340	365720	20259	104240	545559	1164597	27338	157	10315
AVE MOL WT	19.13	20.96	9.95	40.96	40.90	28.014	27.97	2.20	2.20	10.84	2.2	16.47		28.46	32.064	34.08	17.03
SCFM (14.7 PSIA, 60°F)	1114154	480201	311669	18960	164921	25879	73015	204217	144793	217767	59404	40897		265225			
PRESSURE, PSIA	302	242	210	40	1	50	385	379	375	340	375	50					
TEMPERATURE, °F	380	100	75	75	75	95	90	90	90	460	90	90		90			
ENTHALPY, MM BTU/HR	2748.6	737.95	[illegible]	0.927	1.767	0.681	2.302	8.641	4.707	47.404	1.354	1.775		1.521			

PROCESS FLOW DIAGRAM
LEGEND

CONTROL FUNCTION
LC-LEVEL
FC-FLOW
TC-TEMPERATURE
PC-PRESSURE
XC-UNCLASSIFIED

UTILITY DESIGNATION
ST-STEAM
CW-COOLING WATER
RW-CHILLED WATER
DA-DRY AIR
PA-PLANT AIR
PW-PROCESS WATER
COND-CONDENSATE

ITEM NO. / F.D. NO. — RAW MAT'L OR PRODUCT ENTERING OR LEAVING AREA
MAIN PROCESS LINE
AUXILIARY OR SERVICE LINE
A AREA NUMBER / B LINE (STREAM) NO.
PRESSURE PSIG
PRESSURE MM HgABS
GALLONS PER MINUTE
STD FT³/MIN (60°F & 14.7 PSIA)
LBS/HR.
TEMPERATURE, °F

PRELIMINARY NOT APPROVED FOR CONSTRUCTION

A.K. SHAH 1-20-82
D.J. PATRICK 1-26-82
R. ROODMAN 5-16-82
L. SAROFF

PROCESS BLOCK FLOW DIAGRAM
LURGI MOVING BED COAL GASIFICATION
OXYGEN BLOWN - CRYOGENIC SEPARATION

DE-AC21-81MC16447

Dravo Engineers and Constructors
Pittsburgh Pennsylvania

AFC-3783 4-18 100-003 A

selected to remove acid gases by using methanol as solvent and N_2 as stripping gas. Four parallel trains are provided. The gas feed is separated into three streams:

Product Gas:	This stream contains most of the H_2 and CO and is fed to the Cryogenic Gas Separation Area to separate H_2 and CO.
H_2S Fraction:	This stream contains most of sulfur compounds and CO_2. It is fed to the Sulfur Recovery Area.
Tail Gas:	This stream contains hydrocarbons, CO_2 and inert gases. The gas is used as fuel in the high pressure boiler.

The product gas at 210 Psig and 95°F from the Acid Gas Removal System is fed to the Cryogenic Gas Separation Area. Two parallel trains are provided. Product gas is separated into a H_2 - product, CO-product and residual gas. The residual gas is burned in the high pressure boiler as fuel. CO product and some of the H_2 are blended to provide a H_2/CO ratio of 2:1 product at 375 Psig and 90°F which is fed to the gas heating area. Excess H_2 is burned in the high pressure boiler to generate process steam.

The 2:1 ratio product gas is heated from 90°F to 460°F by 600 psig saturated steam in the gas heating area. This final product at 340 psig and 460°F is fed to the Fischer-Tropsch type plant for further synthesis.

The H_2S fraction from the Acid Gas Removal system is fed to the Sulfur Recovery Area to recover the sulfur as by-product. The sulfur is recovered by Claus process unit and the tail gas is cleaned by Beavon-Stretford Process Unit. Both the processes are described in the EPRI Report AF-244. About 321 short tons per day of sulfur are recovered as by-product.

The air compression, air separation and oxygen compression areas remain the same as described in EPRI Report AF244 except the steam required to run the air compressor has been increased because the fuel gas expander is deleted.

The number of high pressure boilers has been increased to generate the additional steam. Hydrogen and hydrocarbons are used as fuel the high pressure boilers. This fuel is supplemented by coal.

More power is exported than in EPRI Report AF-244 because an increased amount of excess low pressure steam is available to recover as electric power.

The process condensate is processed through a Tar Oil Separation Unit and Phenol Extraction Unit.

The production of hydrocarbon liquids and ammonia remains the same as in EPRI Report AF-244 as changes did not affect their production. About 124 short tons per day of ammonia are produced.

4.1.2.1 CHANGES FROM EPRI AF-244 REPORT

Product gas from the Lurgi gasifier in EPRI Report AF-244 is to be used as the feed for production of Fischer-Tropsch synthesis gas rather than as a fuel for boilers. Therefore, the following changes have been made in the equipment usage.

Oxidant Feed System

Gasifier product gas is not available to run the Gas Expander, (11-1-EX-1), and therefore it is deleted. The high pressure steam (1,450 Psig, 900°F) flow rate to the Air Compressor Steam Turbine, (11-1-T-1), has been increased to drive the Air Feed Compressor, (11-1-C-1).

Gas Cooling

Heat exchanger, (21-1-E-1), previously used to heat the product gas from 103°F to 355°F is now used to heat the high pressure BFW from 279°F to 322°F.

Heat exchanger, (22-1-E-5), previously used to provide stripping steam for the regenerator in the Acid Gas Removal System, is now used to heat vacuum condensate from 109°F to 228°F.

The heat exchangers, (21-1-E-12), (21-1-E-13), previously used to heat the demineralized water and vacuum condensate from 110°F to 220°F, are now used to heat the demineralized water from 70°F to 228°F.

Acid Gas Removal System

The Acid Gas Removal System using Allied Chemical Corporation's Selexol process has been replaced by Rectisol Acid Gas Removal System of Lotepro Corporation. Four trains have been provided.

Cryogenic Gas Separation

A new area has been added to separate the gas into a H_2 - product, a CO-product and a residual gas. Two trains have been provided.

Gas Heating

This area has been added to heat the product gas from 90°F to 460°F. The product gas has enough pressure and therefore compression is not required.

Power Recovery

One additional power recovery turbine has been added to recover the power from excess 340 Psig and 50 Psig steam. A total of four units are used to recover the power.

High Pressure Boiler

Two additional high pressure boilers have been added to generate additional high pressure steam required by the Air Compressor Steam Turbine (11-1-T-1). A total of six high pressure boilers are used to generate the required steam.

Gas Expansion

The gas expansion area has been deleted as product gas is not available for power recovery.

4.1.2.2 PROCESS ENERGY BALANCE

Table 4-7 presents the overall process energy balance for 100% capacity operation. The boundary for the balance encompasses the process units and steam system. Energy content of the streams crossing the boundary is expressed as the sum of the stream's higher heating value, sensible heat above $60^{o}F$, and latent heat of water at $60^{o}F$. Electric power is converted to equivalent theoretical heat energy at 3413 Btu/kwh. The energy balance closes to less than 1%. The discrepancy results from approximations used for some units and from calculations of some heat loads.

4.1.2.3 STEAM, BOILER FEEDWATER AND CONDENSATE

The steam system operates at six pressure levels:

High Pressure - 1,450 Psig, $900^{o}F$
Intermediate Pressure - 600 Psig
Medium Pressure - 340 Psig
Low Pressure - 100 Psig
Low Pressure - 50 Psig
Low Pressure - 20 Psig

High pressure steam is produced in six boilers operating at 1,450 psig. Boiler feedwater from the deaerator is preheated with hot process condensate from the gasifier effluent cooling area. Two heat exchangers are used to preheat the boiler feed water. Steam is superheated to $900^{o}F$. High pressure steam is used to drive the air feed compressor turbines in the Oxidant Feed System. These are back pressure turbines exhausting at 340 Psig and $609^{o}F$.

TABLE 4-7

LURGI CRYOGENIC ENERGY BALANCE

Basis: 60^{o}F, Water as Liquid, 3413 Btu/kwh

MMBTU/HR

HEAT IN	HHV	Sensible	Latent	Radiation	Power	Total
Coal	10,864	5				10,869
Air Compressor Suction Air		12	18			30
Demineralized Water		19				19
Boiler Combustion Air		25	36			61
Auxiliary Power Inputs					233	233
Total	10,864	61	54	0	233	11,212
HEAT OUT						
Ash (Dry Basis)	47	18				65
Gasifier Heat Losses				85		85
Sulfur Product	104	1				105
Ammonia Product	100					100
Naphtha, Tar, Oil, Phenol	913	1				914
Product Gas	4,167	94				4,261
Excess Steam Generator					623	623
Bottoming Cycle Generator					124	124
Steam Surface Condenser			2,368			2,368
Isobutane Condenser			745			745
Boiler Stack gas		239	505			744
Gasifier Effluent Trim Cooling		2	11			13
Rectisol Solvent Coolers		67				67
Cryogenic Coolers		181				181
Oxidant Compressor Cooling		104				104
Tar and Phenol Process Heat		467				467
Steam Heat Losses			36			36
Process Condensate Cooling		297				297
Air Separator Power					4	4
Total	5,331	1,471	3,665	85	751	11,303

$\frac{\text{Output - Input}}{\text{Input}} = 0.81\%$

The intermediate pressure steam is generated in one of the high pressure boilers. This steam is used to heat the product gas to 460°F in the Gas Heating Area and is flashed to generate 100 psig steam.

The medium pressure steam at 340 Psig is mainly supplied by the back pressure turbines used on the Oxidant Feed System. Additional medium pressure waste heat steam is generated in the jacket on the gasifiers and steam generated in the waste heat boilers in the Sulfur Recovery Unit. The medium pressure steam header is controlled by feeding the excess steam to a power recovery turbine. The turbogenerator also acts as the balancing wheel to control the steam system. Swings in steam demand at the different levels are reflected in the power output of the turbogenerator. About 52% of the total medium pressure steam is supplied to the gasifiers and 38% is used for power generation. The remaining steam is used for steam turbine driven pumps which exhaust to the 50 Psig steam level and in the Cryogenic Gas Separation Area.

The 100 Psig steam is generated in the gasifier effluent waste heat boiler, (21-1-E-2), and a small quantity is supplied by flashing 600 Psig condensate. The steam is used in the Process Water Treating Unit.

The 50 Psig steam is supplied by steam from the small back pressure turbine drivers, gasifier effluent process heat recovery, Sulfur Recovery Unit and blowdown flash steam. Blowdown from the high and medium pressure steam generators is combined in a blowdown drum which vents flash steam to the 50 Psig header. The steam is used in the Process Water Treating Unit and in Acid Gas Removal System Rectisol. About 54% steam is used to recover the power.

The 20 Psig steam is used in the Rectisol Acid Gas Removal process .

Critical pumps such as boiler feedwater and condensate pumps are steam turbine driven. Other pump services considered important but not critical use electrical motors for normal operation and steam driven spares for emergency service. All turbine drivers operate from 340 Psig to 50 Psig back pressure.

Steam condensate from the various process units is flashed at 15 Psig in a hot condensate flashdrum, (30-V-3). Flashed steam is vented to the deaerator where it provides a portion of the deaeration steam. The flashed condensate is pumped to the deaerator tray section.

Raw water is treated in a semiautomatic, resin bed demineralization unit to produce demineralized water suitable for a 1,500 Psig boiler system. Storage equivalent to 24 hours of demineralized water production is provided. The raw water requirement is 3,850 gpm. Some demineralized water is also used to satisfy process requirements.

The demineralized water for the steam system is combined with vacuum condensate returned from surface condenser. The combined stream is heated to 228^{o}F by heat exchange in the Gas Cooling Unit. The heated stream is deaerated in a tray type deaerator operating at 15 Psig.

Excess heat and steam generated in the fuel processing sections of the plant are converted to electrical power. Thus, 219.9 MW of power are generated from the process. Power is generated by the expansion of excess 340 Psig and 50 Psig steam in a steam turbine driving a generator. A small amount of power is generated in the air separation plant. The remaining power generated is developed in the operation of a low level heat recovery scheme utilizing an organic fluid in a Rankine cycle.

From Steam, MW	182.5
From Low Level Heat, MW	36.3
From Air Separation,MW	1.1
	219.9 MW

A bottoming cycle using isobutane is employed to recover low temperature heat energy and convert it to by-product power. Isobutane vaporized in the oxidant feed compressor coolers and the gasifier effluent coolers is expanded in a turbine driving a generator (the same generator driven by expansion of steam). The expanded isobutane is condensed against cooling water and recirculated in the cycle. Isobutane surge and storage facilities are provided.

Very low level heat in the plant is rejected to cooling water. This heat comes from:

- Process cooling below 140°F
- Steam turbine surface condenser
- Bottoming cycle condenser

Total heat rejection to cooling water is 3,688 MM Btu/hr.

Cooling water is a circulating system wherein the water is returned to cooling towers, cooled in the towers to 80°F and pumped back to the various users at 60 Psig. The cooling towers are multicell induced draft type. The cooling water circulation rate is 492,000 GPM.

4.2 PROCESS DESCRIPTION: (TEXACO GASIFICATION CASE)

The coal receiving and conveying, coal grinding and slurry preparation, gasification, steam generation and particulate removal and gas cooling areas are the same as described in the EPRI Report AF-642 except minor changes were made as reported elsewhere in this report.

Gasification effluent from the gas cooling area is the feed for the hypersorption and cryogenic separation and purification area of the plant. The gas composition is shown in Table 4-8.

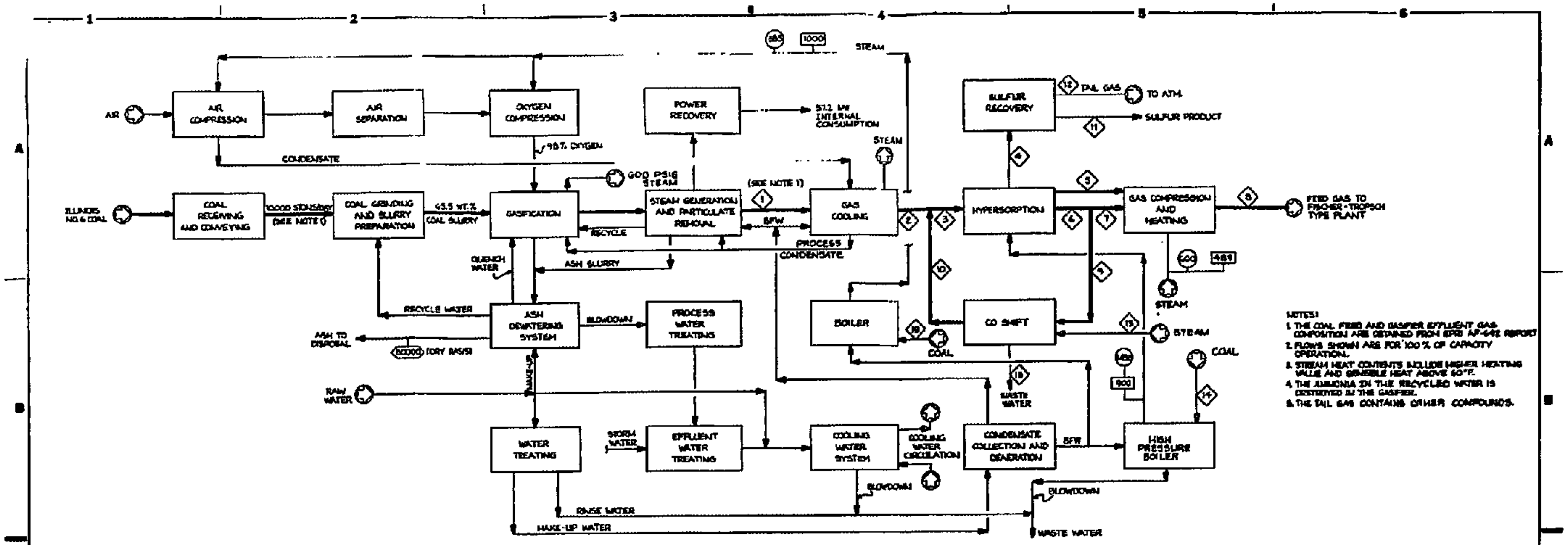
4.2.1 HYPERSORPTION SEPARATION AND PURIFICATION

The general process description is same as it is described for the Lurgi Hypersorption case. A Block Flow Diagram of the Texaco slurry feed gasification with hypersorption separation is shown as AFC-3283-100-004 and the hypersorber Process Flow Diagram as AFC-3283-100-005.

In this case only two hypersorbers are used. Hypersorber 3 is not required because of the low concentration of CH_4 in the gas. Methane remains with the acid gases.

Hypersorber 1: The gas from gas cooling area is fed to Hypersorber 1 at 165 Psig and 100^{o}F. H_2 and CO are separated from the other gases. The desorption temperature is the same as in the Lurgi Hypersorption case, 412^{o}F.

Hypersorber 2: The top product gas from Hypersorber 1 is fed to Hypersorber 2 and H_2 and CO are separated as top and bottom product gases, respectively. The CO desorption temperature is 346^{o}F, same as in the case of Lurgi Hypersorption. In this case the bottom product CO gas is in excess and is fed to the CO Shift Area to produce more hydrogen.



NOTES:
1. THE COAL FEED AND GASIFIER EFFLUENT GAS COMPOSITION ARE OBTAINED FROM EPRI AF-642 REPORT
2. FLOWS SHOWN ARE FOR 100% OF CAPACITY OPERATION.
3. STREAM HEAT CONTENTS INCLUDE HIGHER HEATING VALUE AND SENSIBLE HEAT ABOVE 60°F.
4. THE AMMONIA IN THE RECYCLED WATER IS DESTROYED IN THE GASIFIER.
5. THE TAIL GAS CONTAINS OTHER COMPOUNDS.

MATERIAL / STREAM NO.	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16
LB MOLES/HR	SCRUBBER EFFLUENT	CRUDE GAS	TOTAL CRUDE GAS	ACID GAS	H_2 PRODUCT	CO PRODUCT	CO TO GAS HEATING	PRODUCT GAS	EXCESS CO	CO SHIFT PRODUCT	SULFUR PRODUCT	TAIL GAS (NOTE 5)	STEAM TO CO SHIFT	COAL	WASTE WATER	COAL
CH_4	72.0	72.0	72.0	72.0												
H_2	25974.6	25974.6	48078.4		42807.4			42807.4		16892.8						
CO	38296.5	38296.5	38751.1			38757.1	21405.7	21403.7	17353.4	520.4						
CO_2	7944.5	7944.5	24777.3	24777.3						16892.8						
H_2S	907.0	907.0	907.0	907.0								4.6				
COS	54.0	54.0	54.0	54.0												
N_2, AR	705.5	705.5	908.8	454.4		454.4	251.1	251.1	203.5	203.5						
NH_3 (NOTE 4)	178.0															
SULFUR											849.0					
H_2O	23640.0	185.0	330.0	330.0						189.0			48768.3		31752.8	
COAL LBS/HR														102533		81560
TOTAL LB MOLES/HR	97772.1	74049.1	108621.6	26602.7	42807.4	39211.5	21656.8	64462.7	17556.7	34572.5	849.0	4.6	48768.3		31752.8	
TOTAL LBS/HR	1958281	1532146	2330042	1145413	86300	1099149	607013	693315	492134	1186676	27223	157	878561	102533	572026	81560
AVE. MOL. WT.	20.041	20.691	21.499	43.056	2.016	28.031	28.031	10.755	28.031	34.322	32.064	34.08	18.015		18.015	
SCFM (14.4 PSIA, 60°F)	630618	477900	700624	171690	276272	253064	139754	416020	113305	223126			314741			
PRESSURE, PSIG	557	552	165	165	165	165	165	540	165	165			600		165	
TEMPERATURE, °F	552	100	100	105	100	100	100	460	100	100			650		100	
ENTHALPY, MM BTU/HR	655.275	24.072	38.469	16.079	11.941	10.649	2.061	179.728	4.768	14.987			1195.869		22.846	

PROCESS FLOW DIAGRAM
LEGEND

CONTROL FUNCTION
LC-LEVEL
FC-FLOW
TC-TEMPERATURE
PC-PRESSURE
XC-UNCLASSIFIED

UTILITY DESIGNATION
ST-STEAM
CW-COOLING WATER
RW-CHILLED WATER
DA-DRY AIR
PA-PLANT AIR
PW-PROCESS WATER
(CND)-CONDENSATE

ITEM NO. / F.D. NO. — RAW MAT'L OR PRODUCT ENTERING OR LEAVING AREA
MAIN PROCESS LINE
AUXILIARY OR SERVICE LINE
A AREA NUMBER / B LINE (STREAM) NO.

PRESSURE PSIG
PRESSURE MM HgABS
GALLONS PER MINUTE
STD FT³/MIN (60°F & 14.7 PSIA)
LBS/HR
TEMPERATURE, °F

PRELIMINARY
NOT APPROVED FOR CONSTRUCTION

A.K. SHAH 2-19-82
T.M. REDELL 2-24-82
R. POODMAN 3-26-82
L. SAROFF

PROCESS BLOCK FLOW DIAGRAM
TEXACO (SLURRY FEED)
ENTRAINED BED COAL GASIFICATION
OXYGEN BLOWN-HYPERSORPTION SPRN.

DE-AC21-81MC 16447

Dravo Engineers and Constructors

Pittsburgh Pennsylvania

AFC-3283

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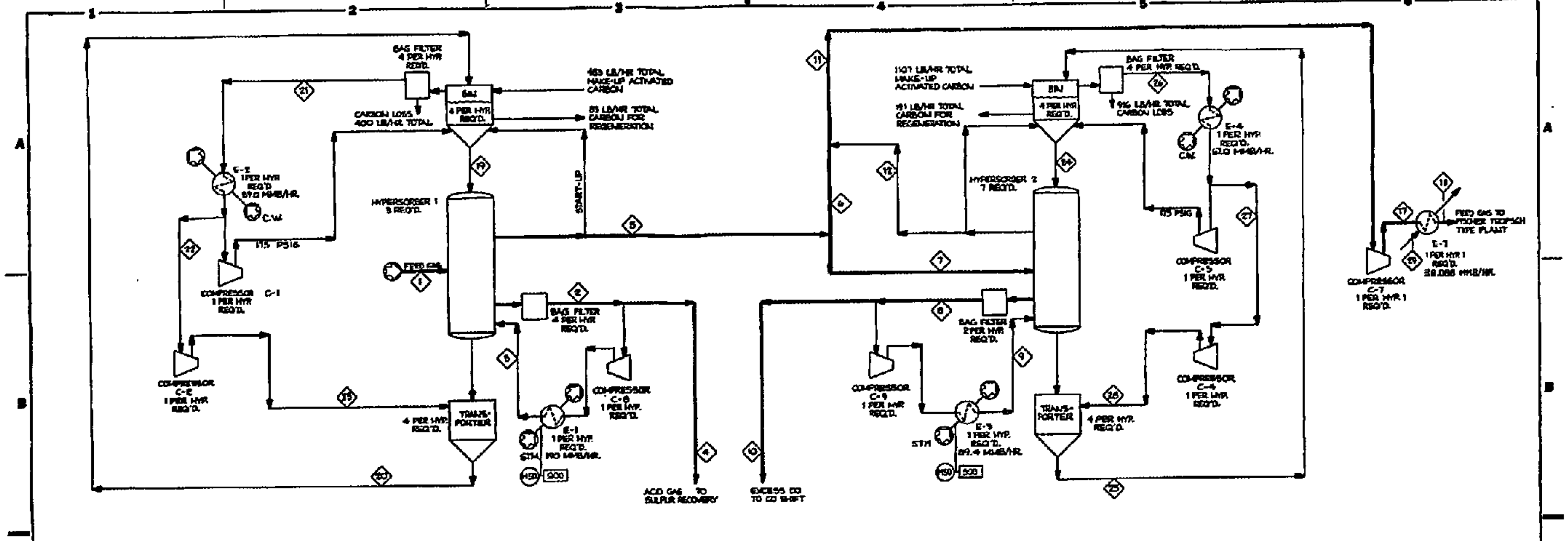

MATERIAL / STREAM No.	1	2	3	4	5	6	7	8	9	10	11	12
LB MOLES/HR.	TOTAL CRUDE GAS	BOTTOM PRODUCT	BOTTOM PRODUCT RECYCLE	ACID GAS	TOP PRODUCT	PRODUCT GAS	HYP. 2 GAS FEED	BOTTOM PRODUCT	BOTTOM PRODUCT RECYCLE	EXCESS CO	FEED TO COMP.	TOP PRODUCT
CH_4	72.0	357.4	285.4	72.0								
H_2	42807.4				42807.4	23640.5	19166.9				42807.4	19166.9
CO	38757.1				38757.1	21405.7	17353.4	[illegible]	[illegible]	17353.4	21405.7	
CO_2	34771.9	119217.0	96498.1	[illegible]								
H_2S	907.0	4510.0	3604.0	907.0								
COS	54.0	255.4	211.4	54.0								
N_2, Ar	408.8	2344.7	1807.1	454.4	454.4	251.1	209.5	1703.5	1509.0	209.5	251.1	
H_2O	538.6	1680.2	1342.2	538.0								
ACTIVATED CARBON LB/HR												
TOTAL LB MOLES/HR	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	17563.7	64462.2	19166.9
TOTAL LB/HR	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	675313	[illegible]
AVG. MOL. WT.	32.499	49.056	49.056	49.056	14.485	14.453	14.453	28.056	28.056	28.081	10.755	2.016
SCFM (14.6 PSIA, 60°F)	701024	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	115508	416028	119700
PRESSURE, PSIG	165	165	170	165	165	165	165	165	170	165	165	165
TEMPERATURE, °F	100	105	100	105	100	100	100	105	900	105	100	100
ENTHALPY, MMBTU/HR	38.649	71.958	[illegible]	16.079	22.890	12.471	10.119	98.964	441.036	4.768	17.922	5.951

MATERIAL / STREAM No.	17	18	19	20	21	22	23	24	25	26	27	28	29
LB MOLES/HR.	FEED TO SYN.	FINAL PRODUCT	CARBON TO HYP. 1	RECYCLE CARBON	GAS FROM BIN HYP. 1	GAS TO COMP. HYP. 1	GAS TO TRANS. HYP. 1	CARBON TO HYP. 2	CARBON RECYCLE	GAS FROM BIN HYP. 2	GAS TO COMP HYP. 2	GAS TO TRANS. HYP. 2	STEAM
CH_4													
H_2	42807.4	42807.4			79635.0	19867.0	19867.0			53180.0	84350.0	84350.0	
CO	21405.7	21405.7			64448.0	17081.0	17081.0						
CO_2													
H_2S													
COS													
N_2, Ar	251.1	251.1			790.0	202.0	202.0						
H_2O													8043.1
ACTIVATED CARBON LB/HR			3531920	3531920				7638190	7638190				
TOTAL LB MOLES/HR	64462.2	64462.2	3531920	3531920	141090.0	36150.0	36150.0	7638190	7638190	53180.0	84850.0	84850.0	8043.1
TOTAL LB/HR	693313	693313			[illegible]	522489	522489			648042	170050	170050	145000
AVG. MOL. WT.	10.735	10.735			14.453	14.453	14.453			2.016	2.016	2.016	18.015
SCFM (14.6 PSIA, 60°F)	416028	416028			[illegible]	233304	233304			[illegible]	344381	344381	51200
PRESSURE, PSIG	340	340	165	165	165	145	245	165	165	165	145	245	600
TEMPERATURE, °F	220	460	100	412	840	85	85	100	346	290	65	85	489
ENTHALPY, MMBTU/HR	71.489	178.728			274.107	2.487	2.487			492.120	5.685	5.885	174.544

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LEGEND

CONTROL FUNCTION
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PC-PRESSURE
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AUXILIARY OR SERVICE LINE
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B LINE (STREAM) NO.

PRESSURE PSIG
PRESSURE MM HgABS
GALLONS PER MINUTE
STD FT³/MIN (60°F & 14.7 PSIA)
LBS/HR.
TEMPERATURE, °F

PRELIMINARY NOT APPROVED FOR CONSTRUCTION

	A.H. SHAH	3-9-82
	CARDOCELL	3-15-82
	R. ROODMAN	3-26-82
	L. SAROFF	

HYPERSORPTION PROCESS FLOW DIAGRAM
TEXACO (SLURRY FEED)
ENTRAINED BED COAL GASIFICATION
OXYGEN BLOWN - HYPERSORPTION SPRN.

DE-AC21-81MC 16447

Dravo Engineers and Constructors

Pittsburgh Pennsylvania

AFC-3283 4-29 100-005 A

TABLE 4-8

GAS COMPOSITION FOR TEXACO CASE, FEED TO SEPARATION PROCESSES

Compound	Moles Per Hour	Mole Percent
CH_4	72.0	0.10
H_2	25,974.6	35.08
CO	38,236.5	51.64
CO_2	7,944.5	10.73
H_2S	907.0	1.22
COS	54.0	0.07
N_2	597.0	0.81
Ar	108.5	0.14
H_2O	155.0	0.21
Total	74,049.1	100.0

Tables 4-9 to 4-12 are equipment lists for the Texaco Hypersorption separation and purification. Transporter system, supply vessels, bin bag filters and controls are a packaged item. Forty bins, 12' diameter x 30' high, are also required.

Excess CO product is shifted in the CO Shift Area to produce more H_2. Exit gas mixture from CO Shift Area is available at about 350 psig and is mixed with the gas from the gas cooling area. The required pressure for the Hypersorption Area is only 165 psig. A gas expander may be used to recover the power. Superheated steam at 600 psig is required in the CO Shift Area and is provided by heat exchangers in the CO Shift Area.

Heat energy available in the CO shift area is utilized to heat the CO Shift Area feed gas from 90°F to 675°F and to produce 100 Psig, 50 Psig, 20 Psig steam as well as to heat the demineralized water to 184°F.

Acid gas from the Hypersorption Area is fed to the Sulfur Recovery Area. Sulfur is recovered in the Stretford process. About 321 short tons per day of sulfur are recovered as by-product.

Air compression, air separation and oxygen compression areas remain the same as described in EPRI Report AF-642 except that additional steam is required to run the air compressor because the fuel gas expanders are unavailable. The number of boilers have been increased to generate the increased amount of steam. The fuel for the boilers is supplemented by coal.

Excess 50 Psig steam is used to recover power in the Power Recovery Area. A total of 57.2 MW is produced for internal consumption. The balance of power required by the processes is imported.

Most of the sour process condensate generated is used for the preparation of the coal slurry feed to the gasifiers. Some of the ammonia present as salts in the water decomposes to nitrogen and hydrogen at the temperature existing in the combustor zone of the gasifiers.

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TABLE 4-9

HYPERSORBERS: TEXACO CASE

Equipment Name	Quantity Required	Size
Hypersorber 1	3	21' Dia. x 36' high
Hypersorber 2	7	14' Dia. x 30' high

TABLE 4-10

BAG FILTERS AT HYPERSORBERS: TEXACO CASE

Bag Filter At	Quantity Required	Design SCFM Each
Hypersorber 1	12	75,000
Hypersorber 2	14	82,000

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TABLE 4-11

COMPRESSORS: TEXACO CASE

Equipment No.	Quantity Required	Design SCFM	Compression Ratio P_2/P_1	BHP Required	Total BHP Required
C-1	3	250,000	1.19	4,050	12,150
C-2	3	77,760	1.63	3,600	10,800
C-3	7	250,000	1.19	4,050	28,350
C-4	7	77,760	1.63	3,600	25,200
C-8	3	250,000	1.14	3,300	9,900
C-9	7	130,000	1.14	1,700	11,900

TABLE 4-12

HEAT EXCHANGERS: TEXACO CASE

Equipment No.	Quantity Required	Heat Duty/Exchanger MMBTU/HR	Surface Area/Exchanger Required Ft.^{2}
E-1	3	190.0	8,400
E-2	3	89.0	11,000
E-3	7	89.4	4,000
E-4	7	67.0	8,300

A small stream of ammonia (as ammonium salts) contaminated effluent leaving the process units is obtained in the gasification area. This effluent is treated in the effluent water treating unit.

Ammonia recovery unit is therefore not provided.

4.2.1.1 CHANGES FROM EPRI REPORT AF-642

Product gas from the Texaco gasifier in EPRI Report AF-642 is to be used for production of Fischer-Tropsch synthesis gas rather than for a combined cycle system. Therefore, the following changes have been made in the equipment.

Oxidant Feed System

Gasifier product gas is not available to run the Fuel Gas Expander, (11-1-Ex-1), and therefore it is deleted. The intermediate pressure steam (385 Psig, 1,000°F) flow rate to the Air Compressor Turbine, (11-1-T-1), has been increased to drive the Air Compressor, (11-1-C-1).

Coal Gasification/Ash Handling

Gasifier heat in EPRI Report AF-642 was used to produce 1,520 Psig steam and 440 Psig steam, and now 1,450 Psig and 600 psig steam is produced.

Gas Cooling

The heat exchanger, (21-1-E-2), previously used to heat the fuel gas from 105°F to 320°F, is now used to produce 50 psig saturated steam.

Heat exchanger, (21-1-E-3), previously used to heat the condensate from 109°F to 203°F, is now used to heat the condensate from 109°F to 254°F.

Acid Gas Removal System

The Acid Gas Removal System using Allied Chemical Corporation's Selexol has been deleted as the acid gases are separated in the Hypersorption Area.

Gas Compression and Heating

Fischer-Tropsch synthesis gas from Hypersorption Area is available at 165 psig and 100°F. This product gas is required at 340 psig and 460°F, therefore a gas compression unit is used to compress the gas. It is heated to 460°F in a heat exchanger using 600 psig saturated steam. The gas expansion area has been deleted.

CO Shift

This area has been added to produce more hydrogen by using the excess CO available in the process. Three trains have been provided.

Steam, Condensate and BFW

Three boilers for 385 Psig and three high pressure boilers for 1,450 Psig steam have been added to produce steam. One steam turbine has been added to recover the power by using excess 50 Psig steam.

Combined Cycle

This area has been eliminated. The equipment on the Drawing EXTC-50/51-1, EPRI Report AF-642 has been eliminated except the Air Compressor Turbine, (11-1-T-1), Oxygen Compressor Turbine, (11-1-T-2), and Surface Condensers, (11-1-E-1), (11-1-E-10), which are included in the oxidant feed system.

4.2.1.2 PROCESS ENERGY BALANCE

Table 4-13 presents the overall process energy balance for Texaco slurry feed Hypersorption System at 100 % capacity operation. The boundary for each balance encompasses the entire plant. Energy content of streams crossing the boundary is expressed as the sum of the stream's higher heating value, sensible heat above 60°F and latent heat of water at 60°F. Electric power is converted to equivalent theoretical heat Btu energy at 3413 Btu/kwh. The energy balance closes to slightly more than 1%. The discrepancies result from approximations used for some process units and from calculations of some heat loads.

4.2.1.3 STEAM, BOILER FEEDWATER AND CONDENSATE

The steam system operates at five pressure levels:

High Pressure - 1,450 Psig, 900°F
Intermediate Pressure - 600 Psig
Medium Pressure - 385 Psig, 1,000°F
Low Pressure - 100 Psig
Low Pressure - 50 Psig

High pressure steam is produced in three high pressure boilers operating at 1,450 psig. The high pressure boiler feed water from a deaerater is preheated in a heat exchanger using 1,450 psig condensate and in the gasifier effluent cooling area. Steam is superheated in the high pressure boiler. High pressure steam is used to heat the bottom product gas in the hypersorption area. High pressure condensate is flashed to produce 600 psig steam.

Most of the intermediate pressure 600 psig saturated steam is produced by using two heat exchangers, one in CO Shift Area and the other in the Gasification Area. The balance of the steam is provided by flashing 1,450 psig condensate at 600 psig. The 600 psig condensate is flashed at 50 psig to produce 50 psig steam.

TABLE 4-13

TEXACO HYPERSORPTION ENERGY BALANCE

Basis: 60^{o}F, Water as Liquid, 3413 Btu/kwh

	MMBTU/HR					
HEAT IN	**HHV**	**Sensible**	**Latent**	**Radiation**	**Power**	**Total**
Coal	12,446	5				12,451
Air Compressor Suction Air		22	53			75
Boiler Combustion Air		13	19			32
Demineralized Water		14				14
Auxiliary Power Inputs					409	409
Total	12,446	54	72	0	409	12,981
HEAT OUT						
Ash (Dry Basis)		81				81
Gasifier Heat Losses				26		26
Gas Cooling		19	6			25
Sulfur Product	105	1				106
Oxidant Compressor Interstage Cooling		535	33			568
Oxidant Compressor Surface Condensers			1,545			1,545
Product Gas	7,870	180				8,050
Sulfur Plant Effluent Gas		2	19			21
Steam Turbines					195	195
Power Block Losses (1)				43	172	215
Steam Turbine Condensers			956			956
Boiler Stack Gas		193	150			343
Steam Heat Losses			22			22
Air Separation Plant Waste Gas		18				18
Waste Water Effluent		19				19
CO Shift Cooler			191			191
CO Shift Waste Water		23				23
Hypersorption Coolers		736				736
Total	7,975	1,807	2,922	69	367	13,140

$\frac{\text{Output - Input}}{\text{Input}} = 1.22\%$

(1) Includes mechanical and electrical losses.

The medium pressure steam is generated in the boiler. Heat available in the Gasification Area and sulfur plant is used to heat the BFW. This steam is used to drive the air feed compressor turbines in the Oxidant Feed System.

The 100 Psig steam is produced by one of the heat exchangers in the CO Shift Area. The sulfur Recovery Area requires 100 psig steam.

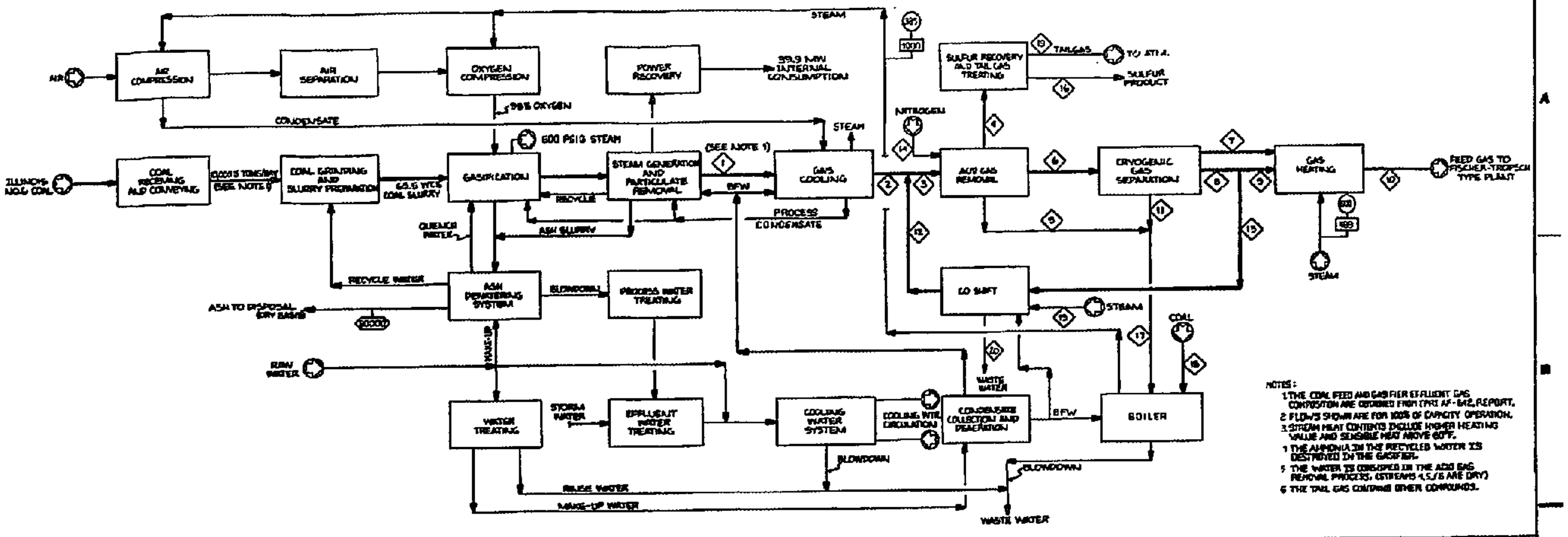
A supply of 50 psig steam is produced by steam generation in process exchangers, (21-1-E-1), generators in the gas cooling unit, (20-1-ME-3), steam generators in the sulfur plant, heat exchange in the CO shift area and 600 Psig condensate flashed at 50 psig. About 81% of the steam is used in the steam turbine, (51-T-4), for power recovery. The balanced steam is used in condensing turbine, (51-T-3), driving the boiler feedwater pump, cryogenic gas separation area, sulfur pit and for steam tracing.

The vacuum condensate from turbines, (11-1-T-1), (11-1-T-2), (51-T-3), (51-T-4), is combined and after heat recovery from the crude gasifier effluents, in (21-1-E-3) flows to the deaerator.

Excess 50 Psig steam generated in the fuel processing sections of the plant are converted to electrical power. Thus, 57.2 MW (195 MMB/hr) of power are generated for internal comsumption. Total power requirement is 119.7 MW (409 MMB/hr). Therefore, 62.5 MW (214 MMB/hr) power must be imported.

4.2.2 CRYOGENIC SEPARATION AND ACID GAS PURIFICATION

Gas composition of the gas leaving the Gas Cooling Area is shown in Table 4-8. A Block Flow Diagram for the Texaco slurry feed gasifier with cryogenic separation is shown in AFC-3283-100-006. Gasifier gas and the gas from CO Shift Area are fed to the Acid Gas Removal System at 532 psig and 100°F. The Rectisol acid gas removal process of Lotepro Corporation, same as in Lurgi cryogenic case, is selected to remove acid gases. Three parallel trains are required. The gas feed is separated into three streams; product gas, H_2S fraction and tail gas.



NOTES:
1 THE COAL FEED AND GASIFIER EFFLUENT GAS COMPOSITION ARE OBTAINED FROM EPRI AF-642 REPORT.
2 FLOWS SHOWN ARE FOR 100% OF CAPACITY OPERATION.
3 STREAM HEAT CONTENTS INCLUDE HIGHER HEATING VALUE AND SENSIBLE HEAT ABOVE 60°F.
4 THE AMMONIA IN THE RECYCLED WATER IS DESTROYED IN THE GASIFIER.
5 THE WATER IS CONSUMED IN THE ACID GAS REMOVAL PROCESS. (STREAMS 4,5,6 ARE DRY)
6 THE TAIL GAS CONTAINS OTHER COMPOUNDS.

MATERIAL / STREAM No.	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18	19	20
LB MOLES/HR	SCRUBBER EFFLUENT	CRUDE GAS	TOTAL CRUDE GAS	ACID GAS	FLASH GAS	SYNFUEL PRODUCT GAS	H₂ PRODUCT	CO PRODUCT	H₂ TO GAS HEATING	PRODUCT GAS	RESIDUAL GAS	CO SHIFT PRODUCT	EXCESS CO	STRIP N₂	STEAM TO CO SHIFT	SULFUR PRODUCT	FUEL GAS	COAL	TAIL GAS (NOTE 6)	WASTE WATER
CH_4	72.0	72.0	72.0		3.4	48.4		0.4	0.2	0.2	68.0		0.2				[illegible]			
H_2	25,576.6	25,576.6	41,263.6		206.3	41,061.3	40,102.9	0.5	0.3	40,211.2	843.5	15,296.0	0.2				1054.3			
CO	38,236.5	38,236.5	38,707.5		580.6	38,126.9	160.2	35,781.5	19,475.5	20,105.7	2,298.1	478.0	13,164.0				1,857.8			
CO_2	7,941.5	7,941.5	23,737.5	3,330.6	20,406.3	0.8					0.8	15,295.0					20,412.1			
H_2S	507.0	507.0	507.0	507.0	<5 PPM	<1 PPM													6.7	
COS	54.0	54.0	54.0	54.0																
N_2, Ar	705.5	705.5	1195.5	67.7	4722.9	1184.9	45.3	1,094.3	604.1	649.8	45.1	490.0	490.0	4,380.0			474.0			
NH_3 (NOTE 4)	171.0																			
SULFUR																877.7				
H_2O (NOTE 5)	234,400.0	150.0	1302									65.2			46,768.5					33410.5
COAL LBS/HR																		96855		
TOTAL LB MOLES/HR	97,712.1	74,048.1	105,633.5	3,891.0	25,525.7	80,494.9	40,346.8	36,276.5	20,020.1	60,965.3	3,332.0	31,449.3	16,754.9	4,380.0	46,168.3	877.7	23,952.1		4.7	33410.5
TOTAL LBS/HR	1,958,139	1,552,364	2,106,077	160,778	1,047,532	1,187,271	101,765	1,016,106	561,372	663,452	67,398	782,511	456,296	123,904	840,564	28,142	1,284,568	96855	160	601,290
AVE. MOL. WT.	20.041	20.968	19.937	41.614	40.400	14.759	2.485	28.065	28.065	10.885	20.226	23.188	28.066	28.014	18.075	32.064	53.346		34.08	18.016
SCFM (14.7 PSIA, 60°F)	660,611	477,900	667,553	24,649	167,350	519,174	254,262	234,112	129,225	389,473	20,788	204,058	104,964	30,845	294,342		155,135			—
PRESSURE, PSIG	557	532	532	35	1	500	465	675	415	465	30	550	475	30	600		1			446
TEMPERATURE, °F	352	100	100	95	95	95	90	90	90	460	80	100	90	95	680		95			100
ENTHALPY, MMBTU/HR	255.275	34,073	35,452	0.944	7.806	19.574	8.566	7.392	4.060	110.154	0.645	11.980	3.372	1.757	1,106,849		8.501			24.039

PROCESS FLOW DIAGRAM
LEGEND

CONTROL FUNCTION
LC-LEVEL
FC-FLOW
TC-TEMPERATURE
PC-PRESSURE
XC-UNCLASSIFIED

UTILITY DESIGNATION
ST-STEAM
CW-COOLING WATER
RW-CHILLED WATER
DA-DRY AIR
PA-PLANT AIR
PW-PROCESS WATER
COND-CONDENSATE

ITEM NO. / P.D. NO. — RAW MAT'L OR PRODUCT ENTERING OR LEAVING AREA
MAIN PROCESS LINE
AUXILIARY OR SERVICE LINE
A AREA NUMBER B LINE(STREAM)NO
PRESSURE PSIG
PRESSURE MM HgAbS
GALLONS PER MINUTE
STD FT³/MIN (60°F & 14.7 PSIA)
LBS/HR
TEMPERATURE, °F

PRELIMINARY
not approved for construction

DRAWN	A.M. SUAH	1-21-82
	R.J. FLETCHER	1-25-82
CHECKED	R. RODMAN	1-26-82
APPROVED	L. SAROFF	

PROCESS BLOCK FLOW DIAGRAM
ENTRAINED EFD TEXACO (SLURRY FEED)
OXYGEN COAL GASIFICATION
BLOWN-CRYOGENIC SEPARATION

DE-AC21-81MC16447

Dravo Engineers and Constructors
Pittsburgh Pennsylvania

AFC-3283 4-39 100-006 A

For further description, refer to the process description of Lurgi Cryogenic case.

The product gas at 500 psig and 95°F from the acid gas removal system is fed to the Cryogenic Gas Separation Area. Two parallel trains are provided. Product gas is separated into a H_2-product, CO-product and residual gas. The residual gas is burned as fuel in the boilers. H_2 and CO are then mixed in a 2:1 ratio of H_2/CO.

Excess CO product is shifted in the CO Shift Area to produce more H_2. The resultant gas mixture of H_2, CO_2 and unreacted CO, H_2O is fed to the Acid Gas Removal System to remove CO_2. Superheated steam at 600 psig required in the CO Shift Area is provided by superheating 600 Psig saturated steam in one of the CO shift heat exchangers. The heat available in the CO Shift Area is utilized to heat the feed gas from 90°F to 675°F for CO Shift Area, to produce 100 Psig, 50 Psig, 20 Psig steam and to heat demineralized water to 184°F.

The final product at 465 Psig and 460°F having a H_2/CO ratio of 2:1, is the feed gas for the Fischer-Tropsch type plant. Synthesis gas is available at a higher pressure than is required, and gas expansion may be considered to recover the power.

The H_2S fraction from the Acid Gas Removal System is fed to the Sulfur Recovery Area to recover the sulfur as by-product. The sulfur is recovered by a Claus process and the tail gas is cleaned by the Beavon-Stretford Process Unit. Both the processes are described in the EPRI Report AF-642. About 338 short tons per day of sulfur are produced.

The air compression, air separation and oxygen compression areas remain the same as described in EPRI Report AF-642 except that the steam required to run the air compressor has been increased because the fuel gas expander is deleted. The number of boilers have been increased to generate the increased amount of steam. Boiler fuel is supplemented by coal.

Excess 50 Psig steam is used to recover the power in the Power Recovery Area. A 39.9 MW of power is produced for internal consumption. The balance of power required by the process is imported.

Most of the sour process condensate generated is used for the preparation of the coal slurry feed to the gasifiers. Some of the ammonia present as salts in the water decomposes to nitrogen and hydrogen at the temperature existing in the combustor zone of the gasifiers.

A small stream of ammonia (as ammonium salts) contaminated effluent leaving the process units is obtained in the gasification area. This effluent is treated in the effluent water treating unit.

A unit for the recovery of by-product ammonia is therefore not provided.

4.2.2.1 CHANGES FROM EPRI REPORT AF-642

Product gas from the Texaco gasifier in EPRI Report AF-642 is to be used for a production of Fischer-Tropsch synthesis gas rather than for a combined cycle system. Therefore, the following changes have been made in the equipment.

Oxidant Feed System

The product gas is not available to run the Fuel Gas Expander, (11-1-Ex-1), therefore it is deleted. The intermediate pressure steam (385 Psig, 1,000°F) flow rate to the Air Compressor Turbine, (11-1-T-1), has been increased to drive the Air Compressor, (11-1-C-1).

Coal Gasification/Ash Handling

The gasifier heat, previously used to produce 1,520 Psig steam and 440 Psig steam, is used to produce 600 Psig steam and to heat BFW from 250°F to 458°F.

Gas Cooling

The heat exchanger, (21-1-E-2), previously used to heat the fuel gas from 105°F to 320°F, is now used to produce 50 Psig saturated steam.

The heat exchanger, (21-1-E-3), previously used to heat the condensate from 109°F to 203°F, is now used to heat the condensate from 109°F to 270°F.

Acid Gas Removal System

The Selexol Acid Gas Removal System of Allied Chemical Corporation has been replaced by the Rectisol Acid Gas Removal System of Lotepro Corporation. Three trains of Rectisol have been provided.

Cryogenic Gas Separation

This new area has been added to separate the gas into H_2-product, CO-product and residual gas. Two trains have been provided.

Gas Heating

Gas Heating area has been added to heat the synthesis gas from 90°F to 460°F. Since the synthesis gas has a higher pressure than required, pressure energy recovery may be considered in the future.

CO Shift

This area has been added to produce more hydrogen by using excess CO available in the process. Three trains have been provided.

Steam, Condensate and BFW

Three boilers have been added to produce 385 Psig, 1,000°F steam. One steam turbine has been added to recover the power by using excess 50 Psig steam.

Combined Cycle

This area has been eliminated. The equipment on the Drawing EXTC-50/51-1 has been eliminated except the Air Compressor Turbine, (11-1-T-1), Oxygen Compressor Turbine, (11-1-T-2), and Surface Condensers, (11-1-E-1), (11-1-E-10), which are included in the oxidant feed system.

Gas Expansion

This area has been deleted. The product gas is available at 465 Psig. Power recovery from 460 psig to 340 psig could be considered in future.

4.2.2.2 PROCESS ENERGY BALANCE

Table 4-14 presents an overall process energy balance for the Texaco slurry feed case at 100 % operation. The boundary for each balance encompasses the entire plant. Energy content of the streams crossing the boundary is expressed as the sum of the stream's higher heating value, sensible heat above 60^{0}F and latent heat of water at 60^{0}F. Electric power is converted to equivalent theoretical heat energy at 3413 Btu/kwh. The energy balance closes to less than 1%. The discrepancies result from approximations used for some units and from calculations of some heat loads.

4.2.2.3 STEAM, BOILER FEEDWATER AND CONDENSATE

The steam system operates at five pressure levels:

High Pressure - 600 Psig
Intermediate Pressure - 385 Psig, $1,000^{0}$F
Low Pressure - 100 Psig
Low Pressure - 50 Psig
Low Pressure - 20 Psig

High pressure 600 psig saturated steam is produced by two heat exchangers, one in the CO Shift Area and the other in Gasification Area. The high pressure

TABLE 4-14

TEXACO CRYOGENIC ENERGY BALANCE

Basis: 60°F, Water as Liquid, 3413 Btu/kwh

	MMBTU/HR					
HEAT IN	HHV	Sensible	Latent	Radiation	Power	Total
Coal	11,405	6				11,411
Air Compressor Suction Air		22	53			75
Boiler Combustion Air		10	14			24
Demineralized and Raw Water		14				14
Auxiliary Power Inputs					297	297
Total	11,405	52	67	0	297	11,821
HEAT OUT						
Ash Slurry		81				81
Gasifier Heat Losses				26		26
Gas Cooling		19	6			25
Sulfur Product	105	1				106
Oxidant Compressor Interstage Cooling		535	33			568
Oxidant Compressor Surface Condensers			1,545			1,545
Product Gas	7,393	170				7,563
Sulfur Plant Effluent Gas		2	19			21
Steam Turbines					136	136
Power Block Losses (1)				43	172	215
Steam Turbine Condenser			717			717
Boiler Stack Gas		145	112			257
Steam Heat Losses			22			22
Rectisol Solvent Coolers		96				96
Cryogenic Coolers		210				210
Air Separation Plant Waste Gas		18				18
Waste Water Effluent		19				19
CO Shift Cooler			191			191
CO Shift Waste Water		24				24
Total	7,498	1,320	2,645	69	308	11,840

$$\frac{\text{Output} - \text{Input}}{\text{Input}} = 0.16\%$$

(1) Includes mechanical and electrical losses.

steam is used in the Cryogenic Gas Separation, Gas Heating and CO Shift Areas. Saturated steam is superheated to 650°F by one of the heat exchangers in the CO Shift Area. This superheated steam is used in the CO shift converter. The 600 psig condensate is flashed to produce 50 psig steam.

Intermediate pressure steam is generated in the boilers, in the Gasification Area, and the sulfur plant. This steam is used to heat BFW, and to drive air feed compressor turbines in the Oxidant Feed System.

The 100 psig steam is produced by one of the heat exchangers in the CO Shift Area, and is used in the Rectisol acid gas removal system and the Sulfur Recovery Area.

The 50 psig steam is supplied by steam generation in process exchangers, (21-1-E-1), generators in the gas cooling unit, (20-1-ME-3), steam generators in the sulfur plant, heat exchange in the CO Shift Area and 600 Psig condensate flashed at 50 psig. About 75% of the steam is used in the steam turbine, (51-T-4), for power recovery. The balance of the steam is used in condensing turbine, (51-T-3), driving the boiler feedwater pump, Cryogenic Gas Separation Area, sulfur pit and for steam tracing.

The 20 psig steam is generated in one of the heat exchangers in the CO Shift Area. The 20 psig steam is used in the Rectisol Acid Gas Removal System.

Raw water is treated in a semiautomatic, resin bed demineralization unit, to produce demineralized water suitable for a 385 psig boiler. Demineralized water from the storage tank is heated in the CO shift heat exchanger and pumped to the deaerator. The condensate from the 100 psig, 50 psig steam and 20 psig users also flows to the deaerator.

The vacuum condensate from turbines, (11-1-T-1), (11-1-T-2), (51-T-3), (51-T-4), is combined and after heat recovery, (21-1-E-3), from the crude gasifier effluents flows to the deaerator.

Excess 50 Psig steam generated in the fuel processing sections of the plant are converted to electrical power. Thus, 39.9 MW (136 MMB/hr) of power is generated, for internal consumption. Total power requirement is 87.0 MW (297 MMB/hr). Therefore, 47.1 MW (161 MMB/hr) power must be imported.

4.3 PROCESS DESCRIPTION: (FOSTER - WHEELER GASIFICATION CASE)

The coal receiving and conveying, coal pulverization and transportation, coal feeding system, gasification, char recovery, gas cooling and transport gas compression areas are the same as described in the EPRI Report AF-642 except that minor changes were made as reported elsewhere in this report.

Gasification effluent from the gas cooling area is the feed for the hypersorption and cryogenic separation and purification area of the plant. The gas composition is shown in Table 4-15.

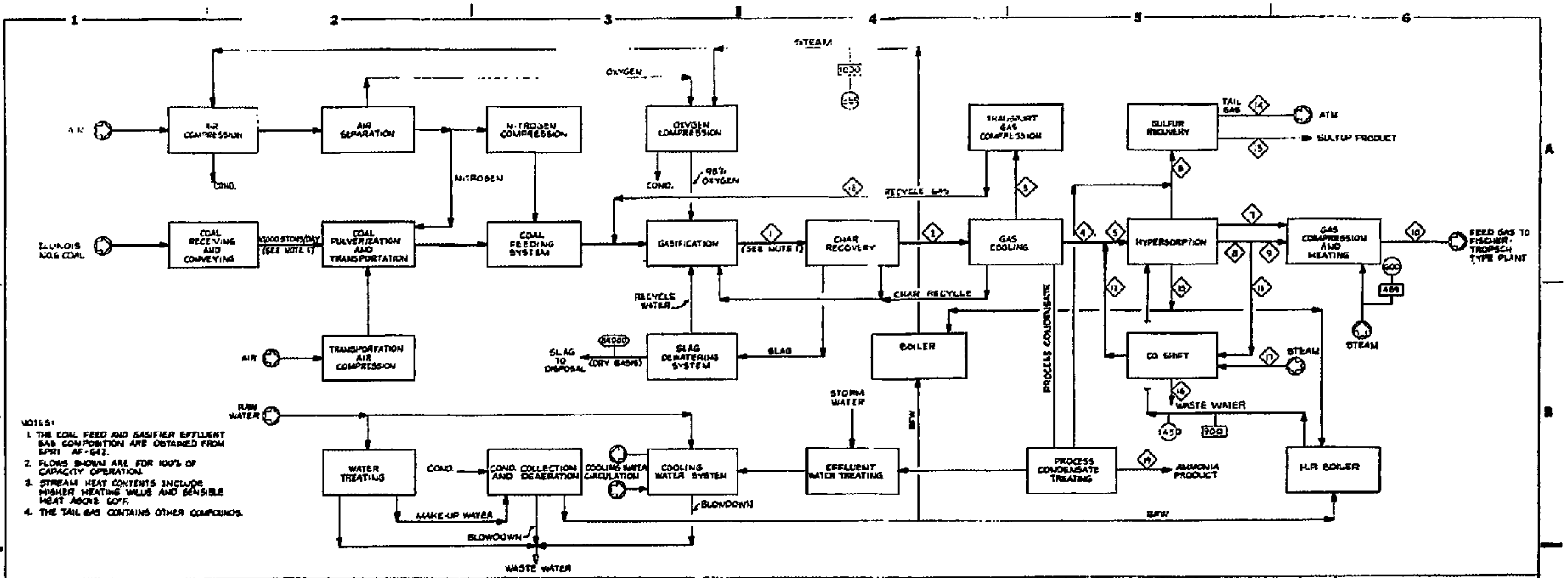
4.3.1 HYPERSORPTION SEPARATION AND PURIFICATION

The general process description is same as described for the Lurgi Hypersorption case. A Block Flow Diagram of the Foster-Wheeler gasification process with hypersorption separation and purification is shown as AFC-3283-100-007 and the hypersorber Process Flow Diagram as AFC-3283-100-008.

Hypersorber 1: The gas from the gas cooling area is fed to the Hypersorber 1 at 165 Psig and 100^{0}F. The H_2 and CO are separated from the other gases. The desorption temperature is the same as in the Lurgi Hypersorption case 412^{0}F.

The top product is feed to Hypersorber 2. The bottom gas product flows to Hypersorber 3.

Hypersorber 2: The top product gas from Hypersorber 1 is fed to Hypersorber 2, and H_2 and CO are separated as top and bottom product gases, respectively. The CO desorption temperature is 346^{0}F, same as in case of the Lurgi Hypersorption. In this case, bottom product CO gas is in excess and therefore it is fed to the CO Shift Area to produce more hydrogen.



NOTES:

1. THE COAL FEED AND GASIFIER EFFLUENT GAS COMPOSITION ARE OBTAINED FROM EPRI AF-642.
2. FLOWS SHOWN ARE FOR 100% OF CAPACITY OPERATION.
3. STREAM HEAT CONTENTS INCLUDE HIGHER HEATING VALUE AND SENSIBLE HEAT ABOVE 60°F.
4. THE TAIL GAS CONTAINS OTHER COMPOUNDS.

MATERIAL / STREAM No.	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18	19		
LB MOLES/HR	GASIFIER EFFLUENT	GAS TO GAS COOLING	TRANSPORT GAS	CRUDE GAS	TOTAL CRUDE GAS	ACID GAS	H_2 PRODUCT	CO PRODUCT	CO TO GAS HEATING	PRODUCT GAS	EXCESS CO	CO SHIFT PRODUCT	SULFUR PRODUCT	TAIL GAS (NOTE 4)	METHANE	WASTE WATER	STEAM TO CO SHIFT	RECYCLE GAS	AMMONIA PRODUCT		
CH_4	5333.2	5333.7	237.9	5730.3	5950.3										5730.5			232.9			
H_2	26306.7	26306.7	1186.0	25010.1	36716.4		36726.4			36716.4		11655.1						1136.0			
CO	51498.5	51418.5	1479.6	50018.1	50578.4			30579.4	18563.7	18363.7	12016.2	360.5						1419.6			
CO_2	11164.3	11164.8	933.8	10630.9	22164.5	22864.2						11655.7						579.8			
H_2S	917.1	917.1	43.2	873.9	875.9	873.4								4.4				49.2			
COS	91.5	91.5	4.2	87.5	87.8	87.5												4.2			
N_2, AR	400.5	400.5	18.9	361.6	475.6	237.8		32.8	143.3	143.8	94.0	94.0						18.4			
NH_3	570.3	570.3																	570.9		
SULFUR													818.1								
H_2O	12641.5	12641.5	12.6	257.4	583.8	383.8						126.4				24146.5	36048.4	12.6			
COAL LBS/HR																					
TOTAL LB MOLES/HR	89634.6	89634.6	3511.2	72450.6	96342.9	23866.0	36726.4	30612.2	18507.0	55233.4	12110.3	23892.5	818.1	4.4	5730.5	24146.5	36048.4	3571.2	570.3		
TOTAL LBS/HR	1601586	1601586	73522	1491685	2045358	1029420	74040	857587	518381	559427	339209	531473	26137	199	82309	437162	649416	75522	9713		
AVE. MOL. WT.	17.870	17.870	20.940	20.590	17.109	43.128	2.016	28.010	28.040	10.126	28.010	22.062	32.064	34.08	16.043	18.015	18.015	20.590	17.031		
SCFM (14.4 PSIA, 60°F)	575447	575447	75048	467584	621781	154047	231026	197598	119441	350467	76157	154197			33110		232692	23048			
PRESSURE, PSIG	358	314	198	198	165	165	168	165	165	340	165	165			165	300	600	532			
TEMPERATURE, °F	1700	387	105	100	100	105	100	100	100	460	100	100			100	105	650	105			
ENTHALPY, MM BTU/HR	15694.94	5913.916	6.476	26.314	56.404	15.978	10.145	8.519	5.028	154.023	3.291	10.080			1.862	19.641	858.65	1.428			

PROCESS FLOW DIAGRAM
LEGEND

CONTROL FUNCTION
LC-LEVEL
FC-FLOW
TC-TEMPERATURE
PC-PRESSURE
XC-UNCLASSIFIED

UTILITY DESIGNATION
ST-STEAM
CW-COOLING WATER
RW-CHILLED WATER
DA-DRY AIR
PA-PLANT AIR
PW-PROCESS WATER
COND-CONDENSATE

ITEM NO. / F.O. NO. — RAW MAT'L OR PRODUCT ENTERING OR LEAVING AREA
MAIN PROCESS LINE
AUXILIARY OR SERVICE LINE
A AREA NUMBER / B LINE (STREAM) NO.
PRESSURE PSIG
PRESSURE MM HgABS
GALLONS PER MINUTE
STD FT³/MIN (60°F & 14.7 PSIA)
LBS/HR
TEMPERATURE, °F

PRELIMINARY NOT APPROVED FOR CONSTRUCTION

A.K. SHAH	3-8-82
R.J. PATRICK	3-12-82
N. ROODMAN	3-24-82
L. SAROFF	

PROCESS BLOCK FLOW DIAGRAM
FOSTER WHEELER
ENTRAINED BFD COAL GASIFICATION
OXYGEN BLOWN-HYPERSORPTION SEPN.

DE-AC21-81MC 16447

Dravo Engineers and Constructors

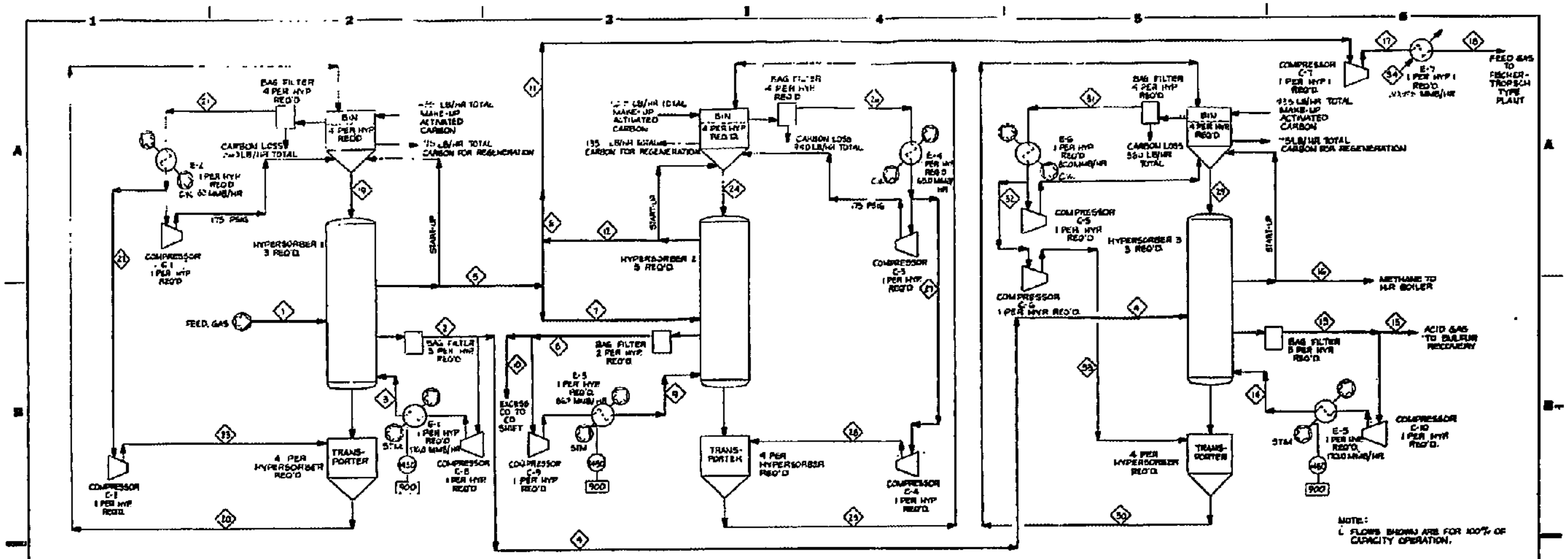
Pittsburgh Pennsylvania

AFC-3283 4-48 100 007 A

TABLE 4-15

GAS COMPOSITION, FOSTER-WHEELER CASE, FEED TO SEPARATION PROCESSES

Compound	Moles Per Hour	Mole Percent
CH_4	5,130.3	7.08
H_2	25,070.7	34.61
CO	30,018.9	41.43
CO_2	10,630.5	14.67
H_2S	873.9	1.21
COS	87.3	0.12
N_2, Ar	381.6	0.53
H_2O	257.4	0.35
Total	72,450.6	100.0



NOTE:
1. FLOWS SHOWN ARE FOR 100% OF CAPACITY OPERATION.

MATERIAL / STREAM No.	1	2	3	4	5	6	7	8	9
LB MOLES/HR	TOTAL CRUDE GAS	BOTTOM PRODUCT	BOTTOM PRODUCT RECYCLE	ACID GAS	TOP PRODUCT	PRODUCT GAS	HYP 2 GAS FEED	BOTTOM PRODUCT	BOTTOM PRODUCT RECYCLE
CH_4	5130.5	22055.4	16925.1	5130.5					
H_2	36726.4				36726.4	22199.7	14526.7		
CO	30579.4				30539.4	18565.2	12016.2	131473.3	103920.0
CO_2	27718.7	95809.4	73528.7	22286.2					
H_2S	873.9	3756.9	2883.0	873.9					
COS	87.5	375.5	288.0	87.1					
N_2+Ar	478.6	1022.5	784.5	237.8	237.8	143.8	94.0	787.0	705.0
H_2O	383.6	16500.0	12662.7	383.6					
ACTIVATED CARBON LB/HR									
TOTAL LB MOLES/HR	96547.9	124669.5	95670.0	28999.5	67543.6	40706.7	26636.9	132260.3	104000.0
TOTAL LBS/HR	1563158	4779312	3667844	1113729	931619	563136	368493	2652919	1523711
AVG MOL WT	21.209	38.354	38.556	38.356	13.854	13.834	13.834	28.010	28.040
SCFM (14.7 PSIA, 60°F)	611181	804595	617638	187157	454624	267114	171910	652647	581490
PRESSURE, PSIG	169	65	175	165	105	165	165	165	115
TEMPERATURE, °F	120	100	700	104	100	100	100	105	700
ENTHALPY, MMBTU/HR	36.404	16.675	568.995	1.840	11.564	1.771	7.545	27.776	457.985

MATERIAL / STREAM No.	10	11	12	13	14	15	16	17	18
LB MOLES/HR	EXCESS CO	FEED TO COMP	TOP PRODUCT	BOTTOM PRODUCT	BOTTOM PRODUCT RECYCLE	ACID GAS	METHANE	FEED TO EXCH	FINAL PRODUCT
CH_4							5130.5		
H_2		36726.4	14526.3					36728.4	36726.0
CO	12016.2	18362.2						18362.2	18361.2
CO_2				111642.1	89325.4	22286.2			
H_2S				4576.6	3502.7	873.9			
COS				459.2	349.8	87.3			
N_2+Ar	94.0	43.8		1191.0	953.2	237.8		143.8	143.8
H_2O				1932.1	1656.9	383.6			
ACTIVATED CARBON LB/HR									
TOTAL LB MOLES/HR	12110.2	55135.4	14576.7	119359.0	95610.0	23469.0	5130.8	55233.4	55233.4
TOTAL LBS/HR	339206	592421	29285	5159467	4126047	1029420	82309	592421	592421
AVG MOL WT	28.010	10.726	2.016	43.128	43.128	43.128	16.045	10.776	10.726
SCFM (14.7 PSIA, 60°F)	76191	356461	93753	771465	617438	154047	33110	356467	356467
PRESSURE, PSIG	165	65	165	165	170	165	165	340	340
TEMPERATURE, °F	105	100	100	105	700	105	100	220	460
ENTHALPY, MMBTU/HR	3.291	15.275	4.052	80.020	514.042	15.378	1.562	61.348	154.013

MATERIAL / STREAM No.	19	20	21	22	23	24	25	26
LB MOLES/HR	CARBON TO HYP 1	CARBON RECYCLE	GAS FROM BIN HYP 1	GAS TO COMP HYP 1	GAS TO TRANS HYP 1	CARBON TO HYP 2	CARBON RECYCLE	GAS FROM BIN HYP 2
CH_4								
H_2			71272.9	19716.2	19716.7			251350.0
CO			59469.7	16507.3	16807.3			
CO_2								
H_2S								
COS								
N_2+Ar			457.4	126.5	126.5			
H_2O								
ACTIVATED CARBON LB/HR	2996940	2996940				5189000	5189000	
TOTAL LB MOLES/HR			130600.0	36150.0	36150.0			251350.0
TOTAL LBS/HR	2996940	2996940	1507822	500099	500098	5189000	5189000	466402
AVG MOL WT			15.834	13.836	13.836			2.016
SCFM (14.7 PSIA, 60°F)			843587	233306	233306			1493094
PRESSURE, PSIG			65	145	245			165
TEMPERATURE, °F			333	85	85			188
ENTHALPY, MMBTU/HR			149.001	2.490	2.490			341.142

MATERIAL / STREAM No.	27	28	29	30	31	32	33	34
LB MOLES/HR	GAS TO COMP HYP 2	GAS TO TRANS HYP 2	CARBON TO HYP 3	CARBON RECYCLE	GAS FROM BIN HYP 3	GAS TO COMP HYP 3	GAS TO TRANS HYP 3	STEAM
CH_4					171150.0	36150.0	36150.0	
H_2	60150.0	60150.0						
CO								
CO_2								
H_2S								
COS								
N_2+Ar								1066.1
H_2O								
ACTIVATED CARBON LB/HR			1996940	1996940				
TOTAL LB MOLES/HR	60150.0	60150.0			171150.0	36150.0	36150.0	7066.1
TOTAL LBS/HR	121464	121464	2996940	1996940	2794438	579954	579954	127296
AVG MOL WT	2.016	2.016			16.043	16.043	16.043	18.015
SCFM (14.7 PSIA, 60°F)	388643	388643			256067	233506	233306	45403
PRESSURE, PSIG	145	245			165	145	245	600
TEMPERATURE, °F	85	85			321	85	88	489
ENTHALPY, MMBTU/HR	4.204	4.204			250.506	3.242	3.242	149.621

PROCESS FLOW DIAGRAM
LEGEND

CONTROL FUNCTION
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PC-PRESSURE
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UTILITY DESIGNATION
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CW-COOLING WATER
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ITEM NO. / F.D. NO. — RAW MAT'L OR PRODUCT ENTERING OR LEAVING AREA
MAIN PROCESS LINE
AUXILIARY OR SERVICE LINE
A AREA NUMBER / B LINE (STREAM) NO.

PRESSURE PSIG
PRESSURE MM HG ABS.
GALLONS PER MINUTE
STD. FT³/MIN. (60°F & 14.7 PSIA)
LBS/HR.
TEMPERATURE, °F

PRELIMINARY NOT APPROVED FOR FABRICATION

DRAWN	A.K.SHAH	5-11-82
CHECKED	D.L.KERIDO	5-16-82
CHECKED	R.BOODMAN	5-26-82
APPROVED	L. SAROFF	

HYPERSORPTION PROCESS FLOW DIAGRAM
FOSTER WHEELER
ENTRAINED BED COAL GASIFICATION
OXYGEN BLOWN-HYPERSORPTION SPRN.

DE-AC21-81 MC 18447

Dravo Engineers and Constructors
Pittsburgh Pennsylvania

AFC-3283 4-60 100-008 A

Hypersorber 3: The bottom product gas from Hypersorber 1 at 165 psig and 100^oF is fed to Hypersorber 3, where CH_4 is separated as the top product gas and acid gases as the bottom product gas. The desorption temperature for the acid gases is 412^oF.

The top product gas, CH_4, is burned in the high pressure boiler while the bottom product acid gas is fed to the Sulfur Recovery Area for sulfur recovery.

Tables 4-16 to 4-19 are equipment lists for the Foster-Wheeler hypersorption separation and purification. Transporter system, supply vessels, bin bag filters and controls are packaged items. Forty-four bins, 12' diameter x 30' ft. high, are also required.

Excess CO product is shifted in the CO Shift Area to produce more H_2. The gas mixture from the CO Shift Area is available at about 350 psig and is mixed with the gas from the gas cooling area. The required pressure for the Hypersorption Area is only 165 psig. A gas expander may be used to recover the power. The 600 Psig superheated steam required in the CO Shift Area is provided by superheating 600 Psig saturated steam in one of the CO shift heat exchangers. The heat available in the CO shift is utilized to heat the feed gas from 90^oF to 675^oF for CO Shift Area, to produce 100 psig, 50 psig, 20 psig steam and to heat demineralized water to 184^oF.

The H_2S fraction from the Hypersorption Area is fed to the Sulfur Recovery Area to recover the sulfur as by-product. Because of very low concentration of sulfur compounds, a Stretford Process is used for sulfur recovery. About 315 short tons per day of sulfur are recovered as by-product.

The air compression, air separation, nitrogen compression and oxygen compression areas remain the same as described in the EPRI report. The number of boilers have been increased to generate the additional steam required.

The ammonia production remains the same as in EPRI report AF-642 because the process changes did not affect ammonia formation. Process condensate

TABLE 4-16

HYPERSORBERS: FOSTER-WHEELER CASE

Equipment Name	Quantity Required	Size
Hypersorber 1	3	20' Dia. x 36' high
Hypersorber 2	5	14' Dia. x 30' high
Hypersorber 3	3	24' Dia. x 40' high

TABLE 4-17

BAG FILTERS AT HYPERSORBERS: FOSTER-WHEELER CASE

Bag Filter At	Quantity Required	Design SCFM/ Each
Hypersorber 1	9	75,000
Hypersorber 2	10	75,000
Hypersorber 3	9	75,000

TABLE 4-18

COMPRESSORS: FOSTER-WHEELER CASE

Equipment No.	Quantity Required	Design SCFM	Compression Ratio P_2/P_1	BHP Required	Total BHP Required
C-1	3	230,000	1.19	3,700	11,100
C-2	3	77,760	1.63	3,600	10,800
C-3	5	240,000	1.19	3,900	19,500
C-4	5	77,760	1.63	3,600	18,000
C-5	3	190,000	1.19	3,100	9,300
C-6	3	77,760	1.63	3,600	10,800
C-8	3	230,000	1.14	2,800	8,400
C-9	5	130,000	1.14	1,700	8,500
C-10	3	230,000	1.14	2,800	8,400

TABLE 4-19

HEAT EXCHANGERS: FOSTER-WHEELER CASE

Equipment No.	Quantity Required	Heat Duty/Exchanger MMBTU/HR	Surface Area/Exchanger Required Ft.^{2}
E-1	3	170.0	7,500
E-2	3	80.0	9,890
E-3	5	86.7	4,000
E-4	5	65.0	8,280
E-5	3	170.0	7,500
E-6	3	80.0	9,890

treatment includes U. S. Steel's proprietary PHOSAM-W process. About 117 short tons per day of ammonia are produced.

A Power Recovery Area is not provided as there is no excess steam available to recover the power. All auxiliary power (513 MMB/hr) requirements are imported.

4.3.1.1 CHANGES FROM EPRI REPORT AF-642

Product gas from the Foster-Wheeler gasifier in EPRI Report AF-642 is to be used for a production of a Fischer-Tropsch synthesis gas rather than for a combined cycle system. Therefore, the following changes have been made in the equipment.

Coal Preparation

N_2 is heated from 90°F to 400°F in the Transport Gas Heater (10-1-V-1) by cooling BFW from 423°F to 247°F. The required BFW, which was supplied by the Heat Recovery steam generator, (51-1-B-1:E-6), is now heated from 250°F to 423°F in one of the CO shift heat exchangers and then supplied to the Transport Gas Heater.

Gas Cooling

Heat exchanger, (21-1-E-1), previously used to produce high pressure 1,450 psig steam, is used for producing 600 psig steam.

Heat exchanger, (21-1-E-2), previously used to heat fuel gas from 105°F to 800°F, is used to generate 420 Psig steam.

Heat exchanger, (21-1-E-5), previously used to heat demineralized water and condensate from 95°F to 153°F, is used to heat demineralized water from 70°F to 128°F.

p

Acid Gas Removal System

The Selexol Acid Gas Removal System of Allied Chemical Corporation has been deleted because the acid gases are separated in the Hypersorption Area.

Gas Compression and Heating

This area is added and is similar to the Texaco hypersorption case. The product gas from Hypersorption Area is available at 165 Psig and 100^{0}F. The product gas is required at 340 Psig and 460^{0}F for further synthesis. Therefore, this area is added to compress the gas from 165 Psig to 340 Psig and to heat the gas to 460^{0}F in a heat exchanger using 600 Psig saturated steam. The Gas Expansion Area is deleted as the product gas is required to be compressed to 340 Psig.

CO Shift

This area has been added to produce more hydrogen from the excess CO available in the process. Two trains have been provided.

Steam, Condensate and BFW

Two boilers operating 385 Psig steam and four high pressure boilers generating 1,450 Psig steam have been added.

Combined Cycle

This area has been eliminated. Equipment changes are similar to the Texaco hypersorption case. The equipment on the Drawing EXTC-50/51-1 EPRI Report AF-642 has been eliminated except the Air Compressor Turbine, (11-1-T-1), Oxygen Compressor Turbine, (11-1-T-2), and Surface Condensers, (11-1-E-1), (11-1-E-10), which are included in the oxidant feed system.

4.3.1.2 PROCESS ENERGY BALANCE

Table 4-20 presents the overall process energy balance at 100% capacity operation. The boundary for the balance encompasses the entire plant. Energy content of the streams crossing the boundary is expressed as the sum of the stream's higher heating value, sensible heat above 60°F, and latent heat of water at 60°F. Electric power is converted to equivalent theoretical heat energy at 3413 Btu/kwh. The energy balance closes to slightly less than 2%. The discrepancies result from approximations used for some process units and from calculations of some heat loads.

4.3.1.3 STEAM, BOILER FEEDWATER AND CONDENSATE

The steam system operates at six pressure levels:

High Pressure - 1,450 Psig, 900°F
Intermediate Pressure - 600 Psig
Medium Pressure - 420 Psig
Medium Pressure - 385, Psig, 1,000°F
Low Pressure - 100 Psig
Low Pressure - 50 Psig

High pressure steam is produced in four high pressure boilers operating at 1,450 Psig. High pressure boiler feedwater from a deaerater is preheated in a heat exchanger using 1,450 psig condensate and in the gasifier effluent cooling area. Steam is superheated in the high pressure boiler. High pressure steam is used to heat the bottom product gas in the Hypersorption Area. High pressure condensate is flashed to produce 600 psig steam.

Most of the intermediate pressure 600 psig saturated steam is produced by two heat exchangers, one in the CO Shift Area and the other in the Gasification Area. The additional steam required is provided by flashing 1,450 psig condensate at 600 psig. The 600 psig steam is used in the Gas Heating and CO Shift Areas. Saturated steam is superheated to 650°F by one of the heat

TABLE 4-20

FOSTER-WHEELER HYPERSORPTION ENERGY BALANCE

Basis: 60°F, Water as Liquid, 3413 Btu/kwh

				MMBTU/HR		
HEAT IN	**HHV**	**Sensible**	**Latent**	**Radiation**	**Power**	**Total**
Coal	10,196	5				10,201
Air Compressor Suction Air		16	39			55
Boiler Combustion Air		11	28			39
Transport Air Compressor Suction Air		1	3			4
Demineralized Water		12				12
Auxiliary Power Inputs					513	513
Total	10,196	45	70	0	513	10,824
HEAT OUT						
Ash/Slag	58	101				159
Gasifier Heat Loss				24		24
Gasifier Cooling		34				34
Sulfur Product	106	1				107
Ammonia Product	100					100
Product Gas	6,752	154				6,906
Power Block Losses (1)				47	133	180
Oxidant Compressor Surface Condensers			855			855
Boiler Stack Gas		218	201			419
Gasifier Effluent Cooling		63	205			268
Oxidant Compressor Interstage Cooling		304	1			305
Process Condensate Treating		50	50			100
Transport Air Vent		14	3			17
Steam Heat Losses		18	3			21
Waste Water Effluent		12				12
Pulverizer Vent Gas		48	21			69
Air Separation Plant Waste Gas		14				14
CO Shift Cooler			214			214
CO Shift Waste Water		20				20
Hypersorption Coolers		805				805
Total	7,016	1,856	1,553	71	133	10,629

$$\frac{\text{Input - Output}}{\text{Inputs}} = 1.80\%$$

(1) includes mechanical and electrical losses.

exchangers in the CO Shift Area. This superheated steam is used in the CO shift converter. The 600 psig condensate is flashed to produce 420 psig steam.

Saturated medium pressure, 420 psig, steam is generated by using the heat available in the CO Shift and Gas Cooling Areas. A small quantity is also obtained from flashing 600 psig condensate at 420 psig. Most of the saturated steam is used in the gasification area to meet the process steam demand of the gasifiers. The balance of the steam is used in the process exchangers located in the Process Condensate Treating Area. The 420 psig condensate is flashed to produce 100 Psig steam.

The medium pressure 385 psig steam is generated in the boiler. The heat available in the gasification area and sulfur plant is used to heat BFW. Additional heat is provided by firing coal in the boilers. This steam is used to drive the air feed compressor turbines in the Oxidant Feed System.

The 100 psig steam is supplied by flashing 420 psig condensate. This steam is used in the Sulfur Recovery Area, and Process Condensate Treating. Steam at 20 psig is obtain by flashing the 100 psig condensate.

The 50 Psig steam is generated in the CO shift area heat exchanger. This steam is mainly used in the Process Condensate Treating Area process exchangers.

Raw water is treated in a semiautomatic, resin bed demineralization unit to produce demineralized water suitable for 1,450 and 385 psig boilers. Demineralized water from the storage tank is heated in Gas Cooling Area and pumped to the deaerator.

The vacuum condensate from turbines, (11-1-T-1), (11-1-T-2), is combined and after heat recovery, (21-1-E-3), from the crude gasifier effluents flows to the deaerator.

4.3.2 CRYOGENIC SEPARATION AND ACID GAS PURIFICATION

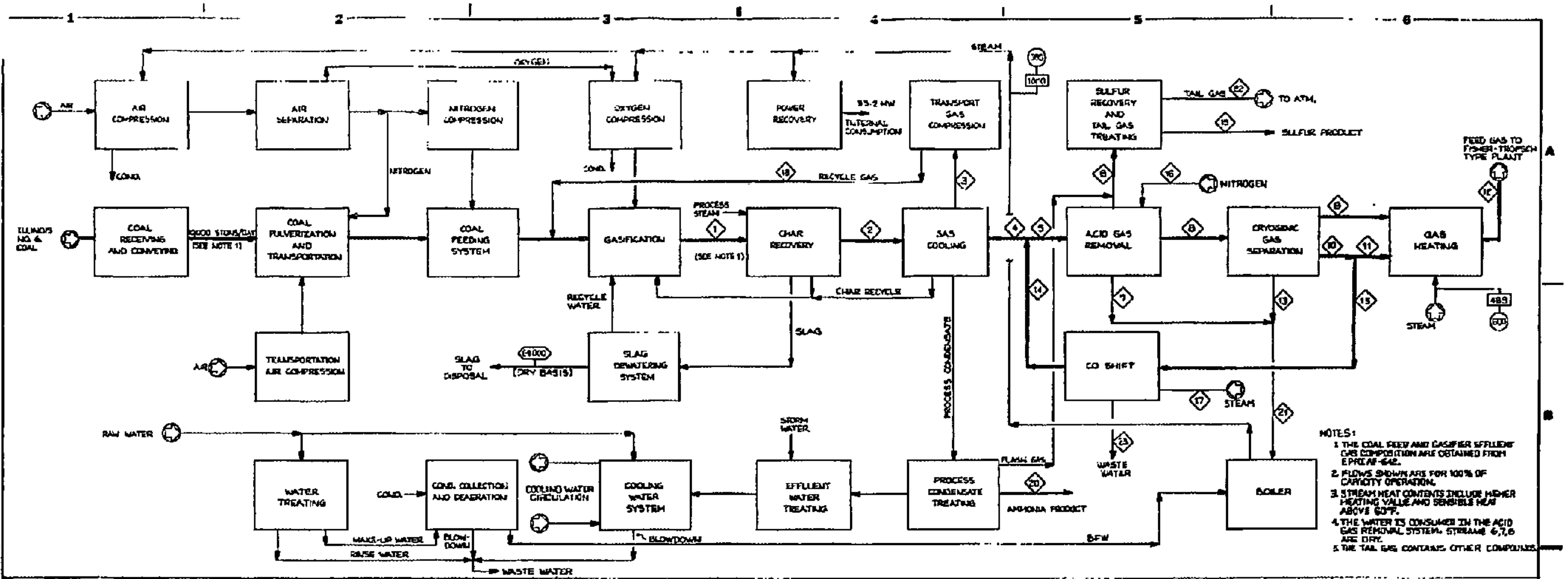
The composition of the gas leaving the Gas Cooling Area is shown in Table 4-15. A Block Flow Diagram for the Foster-Wheeler gasifier with cryogenic separation is shown in AFC-3283-100-009. Gasifier gas and the gas from CO Shift Area are fed to the Acid Gas Removal System at 298 psig and $105^{o}F$. The Rectisol acid gas removal process of Lotepro Corporation, same as the Lurgi and the Texaco Cryogenic Cases, is selected to remove acid gases. Three parallel trains are required. The gas feed is separated into three streams: product gas, H_2S, fraction, and tail gas

For further description, refer to the process description of the Lurgi Cryogenic case.

The product gas at 270 Psig and $100^{o}F$ is fed to the Cryogenic Gas Separation Area. Two parallel trains are provided. Product gas is separated into a H_2-product, CO product and residual gas. The residual gas contains a very high concentration of CH_4 and is burned as fuel in the boilers. Therefore, supplemental fuel coal for the boilers is not required as it was required for the Texaco cryogenic case. H_2 and CO are mixed to produce a stream containing H_2/CO ratio of 2:1.

Excess CO Product is shifted in the CO Shift Area to produce more H_2. The resultant gas mixture of H_2, CO_2 and unreacted CO, H_2O is fed to the Acid Gas Removal System to remove CO_2. The 600 psig superheated steam required in the CO Shift Area is provided by the boiler. Heat available in the CO Shift Area is utilized to heat the feed gas from $90^{o}F$ to $650^{o}F$ for the CO Shift Area, to produce 50 and 20 psig steam, to heat demineralized water to $184^{o}F$, and to heat BFW for the boilers.

The final product at 360 psig and $460^{o}F$ has a H_2/CO ratio of 2:1 is the feed gas to a Fischer-Tropsch type plant.



NOTES:

1. THE COAL FEED AND GASIFIER EFFLUENT GAS COMPOSITION ARE OBTAINED FROM EPRI AF-642.
2. FLOWS SHOWN ARE FOR 100% OF CAPACITY OPERATION.
3. STREAM HEAT CONTENTS INCLUDE HIGHER HEATING VALUE AND SENSIBLE HEAT ABOVE 60°F.
4. THE WATER IS CONSUMED IN THE ACID GAS REMOVAL SYSTEM. STREAMS 6,7,8 ARE DRY.
5. THE TAIL GAS CONTAINS OTHER COMPOUNDS.

MATERIAL / STREAM No.	1	2	3	4	5	6	7	8	9	10	11	12
LB MOLES/HR	GASIFIER EFFLUENT	GAS TO GAS COOLING	TRANSPORT GAS	CRUDE GAS	TOTAL CRUDE GAS	ACID GAS	FLASH GAS	SWEET PRODUCT GAS	H_2 PRODUCT	CO PRODUCT	CO TO GAS HEATING	PRODUCT GAS
CH_4	5383.7	5383.7	252.9	5130.8	5383.5		262.5	4799.0	856.0	320.4	994.3	4450.2
H_2	25306.7	25306.7	1256.0	25010.7	26452.1		187.0	36770.1	36080.0	72.8	4.4	36084.4
CO	31448.5	31498.5	1479.6	50018.9	50170.9		455.8	29915.1		29747.2	18041.2	18042.1
CO_2	11154.3	11154.5	523.8	10630.5	15303.5	2593.3	19409.5	0.1				
H_2S	917.1	911.1	43.2	873.9	873.9	873.9	<5 PPM	<1 PPM				
COS	91.5	91.5	4.2	87.3	87.3	87.3						
N_2, Ar	400.5	400.5	18.9	381.6	537.6	64.1	4407.6	531.9	102.0	395.4	239.4	341.4
NH_3	570.5	570.5										
SULFUR												
H_2O (NOTE 4)	72847.5	72841.5	12.6	857.4	539.7							
TOTAL LB MOLES/HR	89468.1	89468.6	8517.1	72456.6	95921.5	3618.6	24751.4	77216.8	36438.0	50495.8	18480.2	54418.2
TOTAL LB/HR	1801596	1801956	78622	1491853	2033093	150955	995426	1006098	79701	1500722	515193	594094
AVE. MOL. WT.	20.210	20.210	20.590	20.590	21.195	41.716	40.250	14.030	2.187	27.870	27.870	10.850
SCFM (14.4 PSIA, 60°F)	575447	575447	25148	467584	619061	23354	156612	462816	235163	196815	119268	354433
PRESSURE, PSIG	350	319	298	298	298	35	1	270	575	385	385	360
TEMPERATURE, °F	1700	282	105	105	105	100	100	100	90	90	90	450
ENTHALPY, MMBTU/HR	1542.94	392.994	2.428	24.010	36.072	1.289	8.571	33.195	7.440	8.229	3.171	58.741

MATERIAL / STREAM No.	13	14	15	16	17	18	19	20	21	22	23
LB MOLES/HR	RESIDUAL GAS	CO SHIFT PRODUCT	EXCESS CO	STRIP N_2	STEAM TO CO SHIFT	RECYCLE GAS	SULFUR PRODUCT	AMMONIA PRODUCT	FUEL GAS	TAIL GAS (NOTE 5)	WASTE WATER
CH_4	4417.6	124.2	126.8			252.9			4600.1		
H_2	177.3	11380.4	8.4			1256.0			5513		
CO	147.9	352.0	11725.0			1479.6			6013.7		
CO_2	0.1	11973.0				523.8			19410.2		
H_2S						43.2				4.7	
COS						4.2					
N_2, Ar	54.5	156.0	156.0	4460.0		18.9			4456.1		
NH_3								570.3			
SULFUR							864.4				
H_2O (NOTE 4)		61.9			36046.8	12.6					24591.5
TOTAL LB MOLES/HR	4178.0	234109	12015.6	4460.0	36046.8	3571.2	864.4	570.3	39501.4	4.7	24591.5
TOTAL LB/HR	76373	541808	334829	125503	649383	75223	27118	5713	1071799	160	443011
AVE. MOL. WT.	15.770	23.059	27.830	28.014	18.018	20.990	32.064	17.031	26.920	34.08	18.015
SCFM (14.4 PSIA, 60°F)	30834	151477	77547	28913	232640	23048			190448		—
PRESSURE, PSIG	90	300	385	35	460	325			1		300
TEMPERATURE, °F	90	105	90	100	450	105			100		105
ENTHALPY, MMBTU/HR	1.372	10.063	2.452	1.219	671.570	1.428			9.843		19.909

PROCESS FLOW DIAGRAM
LEGEND

CONTROL FUNCTION
LC-LEVEL
FC-FLOW
TC-TEMPERATURE
PC-PRESSURE
XC-UNCLASSIFIED

UTILITY DESIGNATION
ST-STEAM
CW-COOLING WATER
RW-CHILLED WATER
DA-DRY AIR
PA-PLANT AIR
PW-PROCESS WATER
COND-CONDENSATE

ITEM NO. / F.D. NO. — RAW MAT'L OR PRODUCT ENTERING OR LEAVING AREA
MAIN PROCESS LINE
AUXILIARY OR SERVICE LINE
A AREA NUMBER
B LINE (STREAM) NO.

PRESSURE PSIG
PRESSURE MM HgABS
GALLONS PER MINUTE
STD FT³/MIN (60°F & 14.7 PSIA)
LBS/HR.
TEMPERATURE, °F

PRELIMINARY NOT APPROVED FOR CONSTRUCTION

A.K. SHAH
T.M. RODEU 1-30-82
R. ALGMAN 2-26-82
L. SAROFF

PROCESS BLOCK FLOW DIAGRAM
FOSTER WHEELER
ENTRAINED BED COAL GASIFICATION
OXYGEN BLOWN—CRYOGENIC SEPARATION

DE-AC21-81MC 16447

Dravo Engineers and Constructors
Pittsburgh Pennsylvania

AFC-3283
100-009
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The H_2S fraction from the Acid Gas Removal System is fed to the Sulfur Recovery Area to recover the by-product sulfur. The sulfur is recovered in a Claus process and the tail gases are treated in a Beavon-Stretford Process. Both the processes are described in the EPRI Report AF-642 .

The air compression, air separation, nitrogen compression and oxygen compression areas remain the same as described in EPRI Report AF-642. The number of boilers has been increased to generate an increased amount of steam.

The ammonia production remains the same as in EPRI Report AF-642 since process changes did not affect ammonia generation. The process condensate treatment uses the U. S. Steel's proprietary PHOSAM-W process. About 117 short tons per day of ammonia are produced.

4.3.2.1 CHANGES FROM EPRI REPORT AF-642

Product gas from the Foster-Wheeler gasifier in EPRI Report AF-642 is to be used for a production of Fischer-Tropsch synthesis gas rather than for a combined cycle system. Therefore, the following changes have been made in the equipment.

Coal Preparation

Same as in Foster-Wheeler hypersorption case.

Gas Cooling

Same as in Foster-Wheeler hypersorption case.

Acid Gas Removal System

The Selexol Acid Gas Removal System of Allied Chemical Corporation has been replaced by the Rectisol Acid Gas Removal System of Lotepro Corporation. Three trains of Rectisol have been provided.

Cryogenic Gas Separation

This new area has been added to separate the gas into a H_2 product, CO product and residual gas. Two trains have been provided.

Gas Heating

This area has been added to heat the product gas from 90°F to 460°F. The product gas has proper pressure required for the further synthesis.

CO Shift

This area has been added to produce more hydrogen by using the excess CO available in the process. Two trains have been provided.

Steam, Condensate and BFW

Two boilers have been added to produce 385 psig, 1,000°F steam. One steam turbine has been added to recover the power by using the excess 385 psig steam.

Combined Cycle

This area has been eliminated. The equipment on the Drawing EXHC-50/51-1 has been eliminated with the exception of the Air Compressor Turbine, (11-1-T-1), Oxygen Compressor Turbine, (11-1-T-2), and Surface Condenser, (11-1-E-7), which are included in the Oxidant Feed System.

Gas Expansion

This area has been deleted as the product gas has only enough pressure for further gas synthesis.

4.3.2.2 PROCESS ENERGY BALANCE

Table 4-21 present overall process energy balance at 100% capacity operation. The boundary for the each balance encompasses the entire plant. Energy content of the streams crossing the boundary is expressed as the sum of the stream's higher heating value, sensible heat above 60°F, and latent heat of water at 60°F. Electric power is converted to equivalent theoretical heat energy at 3413 Btu/kwh. The energy balance closes to less than 1%. The discrepancies result from approximations used for some units and from calculations of some heat loads.

4.3.2.3 STEAM, BOILER FEEDWATER AND CONDENSATE

The steam system operates at six pressure levels:

High Pressure - 600 psig
Intermediate Pressure - 400 psig
Intermediate Pressure - 385, psig, 1,000°F
Low Pressure - 100 psig
Low Pressure - 50 psig
Low Pressure - 20 psig

High pressure 600 psig saturated steam is produced and used as described in the Texaco cryogenic case. The 600 psig condensate is flashed to produce 50 psig steam.

Saturated intermediate pressure 400 psig steam is generated by using heat available in the CO Shift and Gas Cooling Areas. A small quantity is also obtained from flashing 600 psig condensate at 400 psig. Most of the saturated steam is used in the Gasification Area to meet the process steam demand of the gasifiers. The balance of the steam is used in the process exchangers located in the Process Condensate Treating Area. The 400 psig condensate is flashed to produce 100 psig steam.

TABLE 4-21

FOSTER-WHEELER CRYOGENIC ENERGY BALANCE

Basis: 60°F, Water as Liquid, 3413 Btu/kwh

				MMBTU/HR		
HEAT IN	HHV	Sensible	Latent	Radiation	Power	Total
Coal	10,196	5				10,201
Air Compressor Suction Air		16	39			55
Boiler Combustion Air		11	27			38
Transport Air Compressor Suction Air		1	3			4
Demineralized Water		12				12
Auxiliary Power Inputs					350	350
Total	10,196	45	69	0	350	10,660
HEAT OUT						
Ash/Slag	58	101				159
Gasifier Heat Loss				24		24
Gasifier Cooling		34				34
Sulfur Product	106	1				107
Ammonia Product	100					100
Product Gas	6,806	154				6,960
Generated Power					113	113
Power Block Losses (1)				47	133	180
Oxidant Compressor Surface Condensers			855			855
Power Surface Condensers			223			223
Boiler Stack Gas		134	123			257
Gasifier Effluent Cooling		63	205			268
Oxidant Compressor Interstage Cooling		304	1			305
Rectisol Solvent Coolers		119				119
Cryogenic Coolers		392				392
Process Condensate Treating		50	50			100
Transport Air Vent		14	3			17
Steam Heat Losses		18	3			21
Waste Water Effluent		12				12
Pulverizer Vent Gas		48	21			69
Air Separation Plant Waste Gas		14				14
CO Shift Cooler			214			214
CO Shift Waste Water		20				20
Total	7,070	1,478	1,698	71	246	10,563

$\frac{\text{Input - Output}}{\text{Input}} = 0.90\%$

(1) includes mechanical and electrical losses.

The intermediate pressure steam, 385 psig and 1,000°F, is generated in the two boilers. The heat required for the boilers is provided by the flash gas from Acid Gas Removal System and the residual gas from Cryogenic Gas Separation. Some heat from CO Shift Area is also used. The major part of the steam is used in the turbines to drive the air feed compressors in the Oxidant Feed System. The excess steam available is used for power recovery. In this case, it is possible to generate the excess steam because the fuel gas is available from the Acid Gas Removal System and Cryogenic Gas Separation Area.

The 100 psig steam is supplied by the flashing of 400 psig condensate. This steam is used in the Acid Gas Removal System, Sulfur Recovery Area, and Process Condensate Treating. The 100 psig condensate is flashed to produce 20 psig steam.

The 50 psig steam is generated in the CO Shift Area heat exchanger. This steam is mainly used for the process exchangers in Process Condensate Treating Area, Cryogenic Gas Separation and steam tracing. Hot condensate from the 50 psig steam is flashed to generate 20 psig steam.

The 20 psig steam is mainly generated in the CO Shift Area heat exchanger. The balance is supplied by flashing the hot condensate available in the 100 and 50 psig Steam users. This steam is used in the Rectisol Acid Gas Removal System.

Raw water is treated in a semiautomatic, resin bed demineralization unit to produce demineralized water suitable for 385 psig boiler. Demineralized water from the storage tank is heated in the Gas Cooling Area and pumped to the deaerator.

The vacuum condensate from turbines (11-1-T-1), (11-1-T-2), (51-T-3) is combined and after heat recovery, (21-1-E-3), from the crude gasifier effuents flows to the deaerator. The condensate from the 100, 50 and 20 psig steam users, and the cold BFW from the process heater, (10-1-V-1), also flows to the deaerator.

5. CAPITAL INVESTMENT AND OPERATING CHARGES

Capital investment and operating charge bases are the same as those presented in EPRI Reports AF-244 and AF-642, with adjustment made for increases in the cost of utilities and labor rates applicable in 1981. Each case may have a different basis depending on the base in the EPRI report. Plant areas that were unchanged or minimumly modified have not been repriced as it is outside the scope of the present work. Care must therefore be exercised in evaluating the cost analysis. Pricing of certain areas do not readily lend themselves to indexing. In the EPRI report for the Lurgi oxygen-blown gasification plant, the acid gas and sulfur clean up and recovery have been reported as a single unit price. It was necessary to retain the sulfur clean up. However, a different acid gas removal system was required. Extracting the cost of the sulfur recovery system was accomplished, but with a degree of uncertainty. The estimated cost in 1975 dollars was $19 MM. Escalating the $19 MM price using the CEP index projects a 1981 price of $30 MM. A vendor quotation for a similar system in 1981 was $59 MM. Indexing those portions of the plant that have not been repriced reduces the probability of obtaining meaningful capital investment and plant operating costs (many operating charges are percentages of the capital investment) for comparative evaluation.

No attempt was made to optimize either the hypersorption or the cryogenic facilities. Process designs have been done with the best engineering judgement applied, alternative studies were made where it had been deemed desireable.

Capital charges have not been included in the economic analysis.

Plant investments have been done in accordance with Dravo Standard IV, an accuracy of ± 30% within 95% probability.

5.1 LURGI GASIFICATION CASES

Tables 5-1 and 5-2 show the breakdown of plant area investments for the hypersorption and cryogenic plants, respectively. Areas unaltered from the EPRI Report AF-244 are in 1975 dollars, and those areas which have been added during this study are in 1981 dollars. The areas in which partial changes have been made, cost of the changes have been back calculated in 1975 dollars and then added or subtracted from the original costs. A contingency of 13.2% is identical to that in EPRI Report AF-244. No further estimates of the contingencies were made. Escalating the costs to 1981, with due regard to the inaccuracies that develop, using the CEP index, plant investment for the hypersorption separation and purification is $953 MM and that for the cryogenic separation with acid gas removal is $905 MM.

Operating costs are shown in Table 5-3. These are based on the plant investment costs and the cost of service basis, Table 5-4. Table 5-5 is an estimate of the operating labor required to run the plants. Operating labor requirements are functions of the number of areas and trains within each area. Additional labor is required to operate the hypersorption system due to the number of units involved. A total of 9 hypersorbers are required with 36 transport lines. Labor rates can be found in Table 5-6. The unit cost for the operating charges is based on the net production of products, Table 5-7.

TABLE 5-1

CAPITAL INVESTMENT AT 100% OPERATING LOAD FACTOR

Case: Lurgi Gasifier - Hypersorption

Plant Investment	\$1000(1)	\$1000(2)
Coal Preparation	12,441	
Oxidant Feed System	61,859	
Gasification and Ash Handling	66,212	
Gas Cooling	50,044	
Hypersorption		76,604
Sulfur Recovery		69,684
Gas Compression and Heating		6,664
Process Condensate Treating	76,659	
Power Recovery	47,067	
Utility and Offsite Facilities	100,766	
Subtotal	415,048	152,952
Contingency	54,828	19,099
Total Plant Investment	469,876	172,051

(1) Mid - 1975 Dollars

(2) July - 1981 Dollars

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TABLE 5-2

CAPITAL INVESTMENT AT 100% OPERATING LOAD FACTOR

Case: Lurgi Gasifier - Cryogenic

Plant Investment	$1000(1)	$1000(2)
Coal Preparation	12,441	
Oxidant Feed System	61,859	
Gasification and Ash Handling	66,212	
Gas Cooling	50,054	
Acid Gas Removal System		61,500
Sulfur Recovery and Tail Gas Treating	18,294	
Cryogenic Gas Separation		40,000
Gas Heating		507
Process Condensate Treating	76,659	
Power Recovery	41,606	
Utility and Offsite Facilities	92,642	
Subtotal	419,767	102,007
Contingency	55,456	13,473
Total Plant Investment	475,223	115,480

(1) Mid - 1975 Dollars

(2) July - 1981 Dollars

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TABLE 5-3

LURGI GASIFICATION CASE

OPERATING CHARGES, 330 DAYS/YR., $1,000/YR

	Hypersorption	Cryogenic
Coal at $42/Ton	177,859	147,674
Operating Labor	9,333	8,528
Catalyst and Chemicals	711	749
Utilities	783	800
Activated Carbon	13,171	0
Maintenance, Labor	15,436	14,664
, Materials	23,155	21,996
Administrative and Support Labor	7,431	6,958
General and Administrative Expense	19,057	18,103
Ash Disposal	542	542
Ad Valorem Taxes and Insurance	23,821	22,629
By-Product Electric Power	(67,584)	(50,096)
By-Product Sulfur	0	0
By-Product Ammonia	(7,760)	(7,760)
Total Operating Charges/year	215,955	184,787
Per Unit Production, $/MM Btu	5.37	4.59

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TABLE 5-4

LURGI GASIFICATION CASE

COST OF SERVICE BASIS

Coal	$42.0/ton ($1.72/MM Btu)
Raw Make-Up Water	$0.50/1,000 gallons
Ash Disposal	$1.65/ton
By-Product Power Credit	$0.043/kwh
By-Product Hydrocarbons	Treated as product
By-Product Ammonia Credit	$190.0/ton
Activated Carbon	$1.56/lb.
Regenerated Activated Carbon	$0.55/lb.
By-Product Sulfur Credit	None
Maintenance	4.5%/yr. of on-site plant investment
<u>Maintenance Labor</u>	
Material Ratio	40:60
Labor Burden	50%
Administrative and Support Labor	30% of operating and maintenance labor
General and Administrative Expense	2%/yr. of plant investment

TABLE 5-5

LURGIC GASIFICATION CASE

OPERATING LABOR REQUIREMENTS/SHIFT

	Hypersorption	Cryogenic
"A" Operators	10	7
"B" Operators	31	28
Day Labors (One Shift)	20	18
Lab Technicians	3	2
Foreman	4	4
Superintendent	1	1

TABLE 5-6

LURGI GASIFICATION CASE

OPERATING LABOR GRADES AND RATES

Grade	Wage/Salary Rate
"A" Operator	$13.20/hr.
"B" Operator	$11.10/hr.
Day Laborer	$9.00/hr.
Lab Technician	$11.10/hr.
Foreman	$29,200/yr.
Shift Superintendent	$30,000/yr.

TABLE 5-7

LURGI GASIFICATION CASE

OPERATION FACTORS

Net Production	Hypersorption	Cryogenic
Product Gas, MM Btu/day	99,840	100,008
By-product Liquid Hydrocarbons, MM Btu/day	21,912	21,912
By-product Electric Power, KW	198,450	147,100
By-product Sulfur, St/day	320	328
By-product Ammonia, St/day	124	124

Comparing the hypersorption to the cryogenic facility in 1981 dollars, initial plant investments are within 5% of one another. Capital investments can be considered identical within the accuracy of the feasibility study and the problems associated with cost escalation. The difference in operating charges of $33 MM is primarily due to the additional $30 MM for the coal required to produce high pressure steam of which $17 MM is recovered as an electric power production credit, and the $15 MM yearly requirement for activated carbon make-up and regeneration. The activated carbon amounts to 6.1% of the operating charges. If the attrition of the activated carbon has been overstated to obtain a conservative estimate, the operating charges for the hypersorption case could be reduced. Recent experiment by Yang (16) at Suny Buffalo have shown that it may be possible to use a devolitilized coal char instead of activated carbon as the adsorption medium. This development has the potential to reduce the capital as well as the operting charges. Operating charges per million BTU's produced are 17% higher for the hypersorption system, but within the 30% variation allowed.

The investment cost for the new areas added in this study, excluding the sulfur purification, is $83,200,000 for hypersorption versus $102,007,000 for cryogenic separation and purification, again within the ± 30% accuracy. Sulfur purification is excluded because of the inaccuracy in estimating the investment for the cryogenic case.

Neither system incidates a clear economic advantage.

5.2 TEXACO GASIFICATION CASES

Tables 5-8 and 5-9 show the plant investment using a process area breakdown for the hypersorption and cryogenic plants, respectively. Areas unaltered from EPRI Report AF-642 are in 1976 dollars and those areas which have been added for this study are in 1981 dollars. For the areas in which partial changes have been made, the cost of the changes have been back calculated in 1976 dollars and then added or subtracted from the original costs. The contingency is 19.07%, identical to that in the EPRI Report AF-642. No further estimates of the contingencies were made. Escalating the cost to 1981, with due regard to the inaccuracies that develop, using the CEP index, plant investment for the hypersorption separation and purification is $774 MM and that for the cryogenic separation with acid gas removal is $741 MM.

Operating costs are shown in Table 5-10. These are based on the plant investment costs and the cost of service basis, Table 5-11. Table 5-12 is an estimate of the operating labor required in the plant.

Operating labor requirements are a function of the number of areas and trains within each area.

The unit cost for the operating charges is based on the net production of products, Table 5-13.

TABLE 5-8

CAPITAL INVESTMENT AT 100% OPERATING LOAD FACTOR

Case: Texaco Gasifier - Hypersorption

Plant Investment	$1,000(1)	$1,000(2)
Coal Handling	22,061	
Oxidant Feed	114,122	
Gasification and Ash Handling	24,261	
Gas Cooling	66,986	
Hypersorption		86,193
Sulfur Recovery		69,684
Gas Compression and Heating		10,166
CO Shift		15,630
Steam, Condensate and BFW	14,127	
Support Facilities	55,205	
Subtotal	296,762	181,673
Contingency	56,593	34,645
Total Plant Investment	353,355	216,318

(1) Mid - 1976 Dollars

(2) July - 1981 Dollars

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TABLE 5-9

CAPITAL INVESTMENT AT 100% OPERATING LOAD FACTOR

Case: Texaco Gasifier - Cryogenic

Plant Investment	\$1,000(1)	\$1,000(2)
Coal Handling	22,061	
Oxidant Feed	114,122	
Gasification and Ash Handling	24,261	
Gas Cooling	66,986	
Acid Gas Removal System		71,000
Sulfur Recovery and Tail Gas Treating	19,267	
Cryogenic Gas Separation		44,800
Gas Heating		897
CO Shift		14,616
Steam, Condensate and BFW	9,043	
Support Facilities	55,205	
Subtotal	310,945	131,313
Contingency	59,311	25,047
Total Plant Investment	370,256	156,360

(1) Mid - 1976 Dollars
(2) July - 1981 Dollars

TABLE 5-10

TEXACO GASIFICATION CASE

OPERATING CHARGES, 330 DAYS/YR., $1,000/YR.

	Hypersorption	Cryogenic
Coal at $42/Ton	168,592	155,042
Operating Labor	7,148	5,659
Catalyst and Chemicals	870	892
Utilities	2,735	2,770
Activated Carbon	15,193	0
Maintenance, Labor	8,198	7,913
, Materials	12,298	11,971
Administrative and Support Labor	4,604	4,072
General and Administrative Expense	9,208	8,183
Ash Disposal	523	523
Ad Valorem Taxes and Insurance	19,182	18,514
Electric Power	21,305	16,065
By-Product Sulfur	0	0
By-Product Ammonia	0	0
Total Operating Charges, $/year	269,856	231,604
Per Unit Production, $/MM Btu	4.34	3.96

TABLE 5-11

TEXACO GASIFICATION CASE

COST OF SERVICE BASIS

Coal	$42.00/ton ($1.72/MM Btu)
Raw Make-Up Water	$0.50/1,000 gallons
Ash Disposal	$1.65/ton
Electric Power	$0.043/kwh
Activated Carbon	$1.56/lb.
Regenerated Activated Carbon	$0.55/lb.
By-Products Sulfur Credit	None
Maintenance	2.7%/yr. of on-site plant investment
Maintenance Labor Material Ratio	40:60
Administrative and Support Labor	30% of operating and maintenance labor
General and Administrative Expense	60% of operating and maintenance labor
Property Taxes and Insurance	2.5%/yr. of plant investment
Operating Labor	$17 per manhour average (labor rate includes 35% payroll burden)

TABLE 5-12

TEXACO GASIFICATION CASE

OPERATING LABOR REQUIREMENTS/SHIFT

	Hypersorption	Cryogenic
"A" Operators	10	7
"B" Operators	20	16
Day Labors (One Shift)	11	9
Lab Technicians	3	2
Foreman	3	3
Superintendent	1	1

TABLE 5-13

TEXACO GASIFICATION CASE

OPERATION FACTORS

Net Production	Hypersorption	Cryogenic
Product Gas, MM Btu/day	188,880	177,432
By-Product Sulfur, T/day	327	338

Comparing the plant investment for each facility indicates that the difference is, as in the Lurgi case, within 5% of one another. The projected accuracy of the feasibilty study is ± 30%. Again plant investments for both hypersorption plant or a cryogenic plant are virtually identical. The difference in operating costs is $42 MM, and as with the Lurgi case, is primarily due to the high pressure steam requirements and the cost of replacing the attrited activated carbon. Activated carbon charges amount to 5.6 % of the yearly operating charges. Operating cost to produce one milion Btu's is 9.6% greater for the hypersorption system but within the ±30% variation.

Examination of the investment costs for the new areas added in this study, excluding the sulfur recovery system for reasons previously discussed, shows the investment for hypersorption is $110,702,000 versus $90,993,000 for the cryogenic separation and purification.

Due to the 30% level of this study, neither system indicates a clear economic advantage.

5.3 FOSTER-WHEELER GASIFICATION CASES

Tables 5-14 and 5-15 show the plant investment using a process area breakdown for the hypersorption and cryogenic plants, respectively. Areas unaltered from EPRI Report AF-642 are in 1976 dollars and areas which have been added for this study are in 1981 dollars. For the areas in which partial changes have been made, the cost of the changes have been back calculated in 1976 dollars and then added or subtracted from the original costs. The contingency is 21.5% and is identical to that in EPRI Report AF-642. No further estimates of the contingencies were made. Escalating the cost to 1981, with an understanding that inaccuracies develop, and using the CEP index, plant investment for the hypersorption separation and purification is $667 MM and that for the cryogenic separation with acid gas removal is $621 MM.

Operating costs are shown in Table 5-16. These are based on the plant investment costs and the same cost of services base that were used for the Texaco case Table 5-11. Table 5-17 is an estimate of the operating labor required in the plant. Operating labor requirements are a function of the areas and the trains within each area.

The unit cost for the operating charges is based upon the net production of products, Table 5-18.

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TABLE 5-14

CAPITAL INVESTMENT AT 100% OPERATING LOAD FACTOR

Case: Foster-Wheeler Gasifier - Hypersorption

Plant Investment	\$1,000 (1)	\$1,000 (2)
Coal Handling	29,320	
Oxidant Feed	66,891	
Gasification and Ash Handling	35,603	
Gas Cooling	19,388	
Hypersorption		97,002
Sulfur Recovery		69,684
Gas Compression and Heating		9,150
CO Shift		10,901
Process Condensate Treating	7,648	
Steam, Condensate and BFW	13,453	
Support Facilities	57,551	
Subtotal	229,854	186,737
Contingency	49,356	40,098
Total Plant Investment	279,210	226,835

(1) Mid - 1976 Dollars

(2) July - 1981 Dollars

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TABLE 5-15

CAPITAL INVESTMENT AT 100% OPERATING LOAD FACTOR

Case: Foster-Wheeler Gasifier - Cryogenic

Plant Investment	\$1,000(1)	\$1,000(2)
Coal Handling	29,320	
Oxidant Feed	66,891	
Gasification and Ash Handling	35,603	
Gas Cooling	19,388	
Acid Gas Removal System		69,000
Sulfur Recovery and Tail Gas Treating	19,267	
Cryogenic Gas Separation		44,200
Gas Heating		837
CO Shift		8,712
Process Condensate Treating	7,648	
Steam, Condensate and BFW	10,371	
Support Facilities	57,551	
Subtotal	246,039	122,749
Contingency	52,831	26,357
Total Plant Investment	298,870	149,106

(1) Mid - 1976 Dollars

(2) July - 1981 Dollars

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TABLE 5-16

FOSTER-WHEELER GASIFICATION CASE

OPERATING CHARGES, 330 DAYS/YR., $1,000/YR.

	Hypersorption	Cryogenic
Coal at $42/Ton	138,600	138,600
Operating Labor	7,148	5,361
Catalyst and Chemicals	955	988
Utilities	2,425	2,489
Activated Carbon	18,935	-
Maintenance, Labor	7,258	6,632
, Materials	10,886	9,948
Administrative and Support Labor	4,322	3,598
General and Administrative Expense	8,644	7,196
Ash Disposal	549	549
Ad Valorem Taxes and Insurance	16,443	15,517
Electric Power	51,189	23,968
By-Product Sulfur	0	0
By-Product Ammonia	(7,308)	(7,308)
Total Operating Charges, $/year	260,046	207,538
Per Unit Production, $/MM Btu	4.88	3.85

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TABLE 5-17

FOSTER-WHEELER GASIFICATION CASE

OPERATING LABOR REQUIREMENTS/SHIFT

	Hypersorption	Cryogenic
"A" Operators	10	6
"B" Operators	20	15
Day Labors (One Shift)	11	9
Lab Technicians	3	2
Foreman	3	3
Superintendent	1	1

TABLE 5-18

FOSTER-WHEELER GASIFICATION CASE

OPERATION FACTORS

Net Production	Hypersorption	Cryogenic
Product Gas, MM Btu/day	162,048	163,344
By-Product Sulfur, /day	315	333
By-Product Ammonia, /day	117	117

Comparing the plant investment for separating and purifying the gasification product gas from the Foster-Wheeler gasification process, the investment by a hypersorption system is 7% higher than for cryogenic separation. Again this value is within the limits of accuracy of the study. The difference in operating cost is $64 MM per year, primarily due to the additional electric power used by the hypersorption compressors. In all the other cases sufficient residual steam was available to produce most of the electric power within the plant. However, for this case no internal power generation was possible. Early in the design it was decided to heat and cool the activated carbon by direct contact rather than extending the tower height to include internal heat exchangers. The result of a study projected that 5 to 7 million dollars could be saved per tower configuration for direct heating and cooling. A plant investment saving of approximately $66 MM resulted for the Foster-Wheeler Hypersorption study. However, the operating costs for this case was adversly affected by that decision. Operation charges for hypersorption separation and purification is $4.88/MMBtu as opposed to $3.85/MM Btu for cryogenic separation or 26.8% greater. In addition the plant cost for the separation, acid gas removal, CO shift and gas compression for hypersorption is $114,120,000 versus $122,749,000, for cryogenic, which are almost identical.

Considering the additional equipment required for indirect heating and cooling the hypersorber area investment would be $180,120,000; considerably more than that for cryogenic separation and acid gas purification areas. Economic evaluations for the Foster-Wheeler case leads to the conclusion that hypersorptive separation would be more costly.

6. RESEARCH RECOMMENDATIONS

Detailed design of a Hypersorption Unit requires the knowledge of: maximum loading on the adsorbent of various components from the multicomponent mixtures; dynamics of absorption, as external mass transfer coefficients, mass transfer within the adsorbent pores, and adsorption mechanism (competition for sites from the mixture); thermal considerations, heats of adsorption and desorption temperature, heat transfer methodology for accomplishing the desorption in the rectification zone and cooling the adsorbent; methods of feeding to minimize attrition; and adsorbent transport system, again to minimize attrition. Much of this information is not available at present, however some items can be properly estimated using standard engineering techniques. In addition pilot plant studies would be invaluable in the absence of fundamental data and mathematical simulations.

Equilibrium adsorptions for binary mixtures on molecular sieves have been recently reported by Danner and Co-workers (17, 18). Their techniques can be extended for multicomponent mixtures for any adsorbent. Measurement of mass transfer coefficients both external and internal can be accomplished in numereous ways, many experimental techniques are no doubt available particularly from catalyst mechanistic studies. Dynamic behavior of the adsorption process is described by Szirmay (13). Szirmay's method can be used to determine the lengths of the adsorption and rectification section of the hypersorber. The main thrust of the research recommendation is a plan for building a pilot plant which can be used to obtain adequate data to minimize the risk of the design of hypersorption units.

6.1 HYPERSORPTION PILOT PLANT

It is proposed that a 6 inch diameter moving bed adsorption system be constructed to separate and purify the gas compositions that occur as products from an oxygen - blown gasifier. This plant would require 3,000 lb. inventory of activated carbon or an equal volume of other adsorbant as devolitilized coal char and 3,000 scf of mixed gases. Separated products can be remixed and returned to the system to provide a continuous feed. The length of the hypersorption unit (moving adsorption bed) would depend upon whether direct or indirect heating (desorption section) and cooling of the adsorbent would be used. In general, the design should be useable at pressure up to 30 atmospheres and temperatures to 650°F, provide for sharp separation, evaluate ability to obtain side cuts and determine separation techniques for azeotropic gas mixtures.

6.1.1 DESIGN OF HYPERSORPTION TOWER

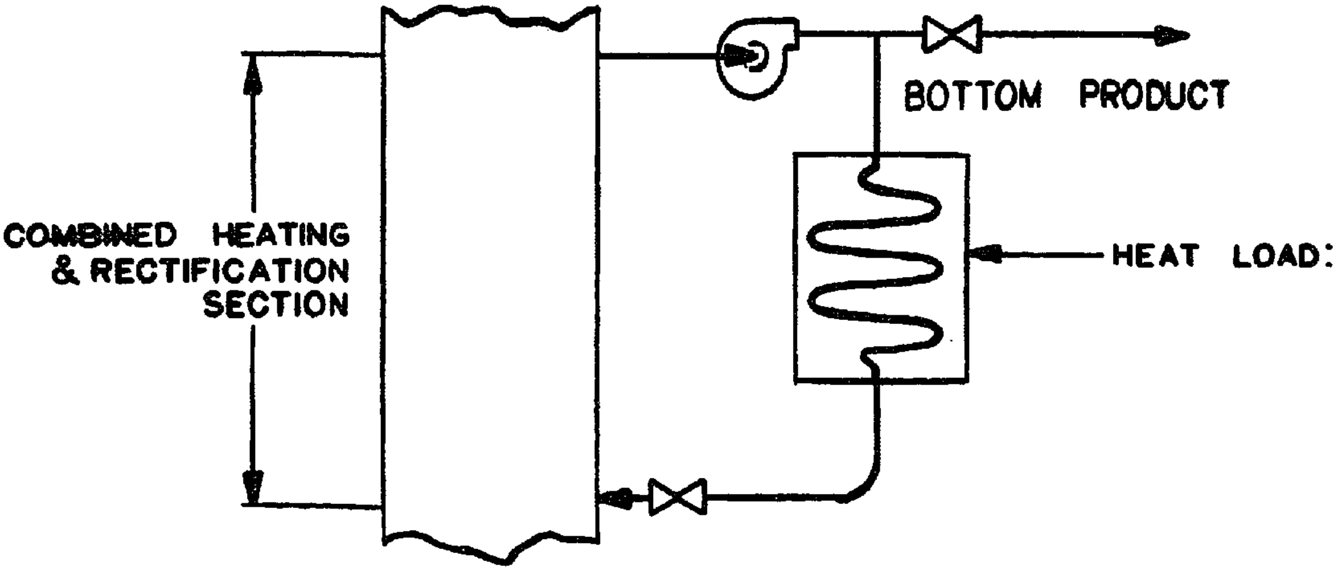
6.1.1.1 COLUMN DIAMETER

The diameter of 6" is based on the work of Szirmay (7). Szirmay's previous work was with a 1" and 8" column. A six inch column in sufficiently large to minimize channelling along the wall and small enough to minimize gas and solids distribution problems. Larger diameters would also entail greater gas flowrates.

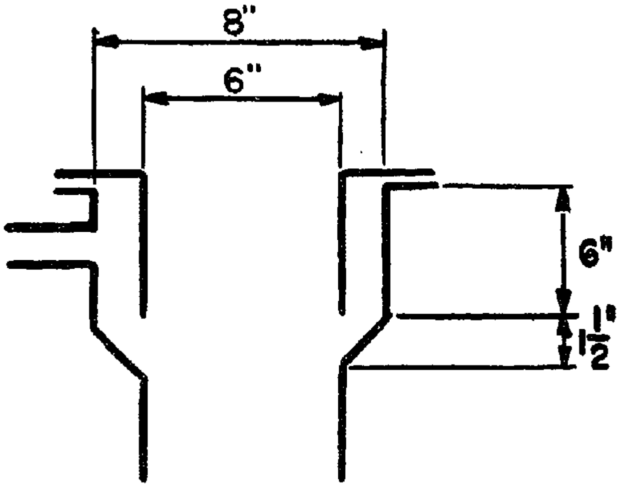
6.1.1.2 COLUMN LENGTH

Column length is composed of many sections; adsorption, desorption and rectification, feeder transition, engaging and disengaging trays, heat transfer transitions, heater and cooler. A summary of these lengths are in Table 6-1.

p



ARRANGEMENT FOR DIRECT HEATING
SK-2



ENGAGING AND DISENGAGING TRAYS
SK-3

In conventional hypersorber design heating and cooling were done within the tower by indirect heat transfer means. Direct heat transfer requires considerably less height. Design of the hypersorber units for the separations discussed in the main body of the report assumed direct heat transfer. Cost studies indicated installed cost advantages for direct heating and cooling versus indirect heat exchangers. A pilot plant incorporating the direct heating for the desorption zone and cooling in a bin before feeding the adsorbant to the hypersorber would reduce the column height to 20 ft. Cooling must be done with the top product gas. If sufficient gas is not available a recycle loop has to be established. A similar situation exists for the direct heating section, however, an additional 2 feet length is required, SK-2.

6.1.2 MECHANICAL AND CONTROL CONSIDERATION

In addition to the following, proper recording instruments should be included.

6.1.2.1 FEED AND DISENGAGING TRAYS

It is important to provide a mechanism for the introduction of the feed and for the withdrawal of top and bottom products, with minimum disturbance to the gas and adsorbent flow pattern. For a 6" ID column, a simple arrangement as shown in SK-3 is adequate.

6.1.2.2 CYCLONES AND BAG FILTERS

Both top and bottom product withdrawals require bag filters for dust control. A cyclone with minimum impingment velocity should be part of the adsorbent transport system. Flat plate impingments should be avoided to minimize attrition losses.

TABLE 6-1

SUMMARY FOR THE DESIGN OF THE ADSORBER COLUMN LENGTH

Adsorbent feeder and transition section from the column to the adsorbent conveying system; estimated length with contingencies	1 ft
Heating Section	20 ft
Desorption and Rectification Sections	3 ft
Adsorption Section	4 ft
Cooling Section	30 ft
Three engaging and disengaging trays, with contingencies @ 1 ft each	3 ft
Four transition sections from heat exchanger to column, with contingencies @ 1 ft	4 ft
Total	68 ft

6.1.2.3 ADSORBENT FEEDER

Only hopper type feeders are recommended, similar to the vibrating screen type. Excellent results have been achieved with this type of feeder in other experiments in obtaining low attrition and steady flow. None of the star, rotary slide valve and similar feeders are recommended because of their manner of operation and possible high attrition rate.

6.1.2.4 ADSORBENT FLOW CONTROL

Monitoring the adsorbent recirculation rate can be accomplished with an Auburn capacitance meter. Proper adsorbent feed and steady operation can be achieved with this type of flow measuring device.

6.1.2.5 ADSORBENT LEVEL CONTROL

Although one definite type of controller cannot be recommended at this point a number of candidates are available with a good chance of success, such as: the electronic type; measuring the change of conductivity or capacitance; or the optical type; and isotope techniques. Optical or isotope types are favored.

6.1.2.6 CONTINUOUS QUALITY CONTROL

Sample ports along the adsorption portion of the column should be installed. The composition should be monitored at these points. This will aid in determining the rate of feed to obtain the proper separation at the exit ports.

Several types of instruments are available for composition monitoring. Infra-red can be used for monitoring some of the gases, continuously. Production type mass-spectrometers are also available for continuous monitoring. Gas chromatography with a direct computer link and automatic sampling can analyze the product gases within 2-3 minutes. Gas analysis should not present any problems.

6.1.2.7 TEMPERATURE CONTROL

Temperature control is needed because improper cooling decreases the capacity of the unit and low desorption temperature also decreases the capacity, as well as contaminates the top product. Proportional controllers with automatically adjustable band control should be adequate, with thermocouples as the measurement device.

6.1.2.8 PRESSURE CONTROL

Pressure should be kept reasonably constant with back pressure controllers. Fast operating units should be employed. Poor pressure control will result in decreased separation reliability. Pressure transducers of the Setra type should not be affected by the solids.

6.1.2.9 PRODUCT CONTROL

A main objective of any pilot program is to determine if sufficient component separation is possible based upon bench scale separations from experimental data. If a component exits at the top but should be a bottom product, based up the original design of the pilot plant tests, several operating condition may be changed to correct the situation:

- decrease the gaseous feed rate
- increase the rate of product withdrawal
- increase the adsorbent recirculation rate
- decrease the desorption temperature

If perchance the opposite occurs, the bottom product is contaminated with a component that should exist with the top product, correction can be made by reversing the above changes in conditions either singly or by an appropriate combination of responses.

6.2 ATTRITION

A better understanding of attrition of activated carbon and other adsorbants as devolitilized coal char is necessary in order to reducing the risk of applying hypersorption processing techniques to separating gaseous components from oxygen-blown gasifiers. Mechanical degradation of activated carbon, particularly in the feed mechanism and the transport system, will increase the operating cost of the hypersorption units.

Typical methods of measuring degradation are: the jet test in a small fluidized bed, tumbling the adsorbant, and determining the crushing strength. The small fluidized bed method requires a one liter vessel with a gas nozzle in the bottom and a screened exit port. Pellets are weighed and loaded into the vessel where a high velocity jet of gas contacts the pellets causing violent motion. Difference in weight before and after the test is a measure of the attrition losses. Tumbling test is accomplished by filling a tube with a weighed amount of pellets, the ends of the tube is closed and tumbled end on end. Again attrition is determined by the weight differences. Catalyst attrition is sometimes measured by determining the force required to crush the pellet. Other variations of attrition test are also used.

The above tests are accelerated attrition tests and are primarily used for screening. Once screening is accomplished it may be desirable to determine attrition at operating conditions. A dense phase flow system is more complicated to operate than a dilute phase transport system and was chosen for the design criteria of the hypersorption separation area in order to minimize attrition losses. A transport loop which would be useable as a dilute or dense phase transport could be built and the adsorbents that showed promise in the accelerated attrition test would be tested in the loop to obtain actual attrition rates.

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