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CRYOGENIC METHANE SEPARATION/ CATALYTIC HYDROGASIFICATION PROCESS ANALYSIS

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INTERIM REPORT

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ABSTRACT

Screening type evaluations of acid gas removal and methane separation were performed for two coal gasification substitute natural gas (SNG) processes: the Exxon Catalytic Coal Gasification (CCG) and Rockwell Hydrogasification processes. The objective of this study was to determine the best combination of acid gas removal and methane separation for each of the coal gasification processes.

Information on raw gas conditions, system pressure, and recycle gas and product SNG specifications, furnished by the primary coal gasification contractors was used as the design basis. Three acid gas removal systems, Selexol, Rectisol, and Benfield HiPure were evaluated for integration with cryogenic methane separation for the Exxon CCG process. MEA, DEA, improved DGA, and Sulfinol acid gas treating systems were evaluated for integration with cryogenic and hypersorption methane separation for the Rockwell Hydrogasification process.

The economic analysis methods employed to evaluate gas treating costs for the different integrations of acid gas removal and methane separation were: utility financing and private investor financing via A.G.A. - CCU and Air Products DCF methods. Based on gas treating cost and the availability of commercial technology, the best acid gas removal/methane separation alternative was chosen for each coal gasification process.

The integrations of Selexol/cryogenic and improved DGA/cryogenic were selected for the Exxon CCG and Rockwell Hydrogasification processes, respectively.

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1. SUMMARY OF RESULTS

The objective of this program is to recommend the most attractive combinations of acid gas removal and cryogenic methane separation for both the Exxon Catalytic Coal Gasification (CCG) and the Rockwell Hydrogasification processes currently undergoing development supported by DOE. This program is comprised of the following three tasks:

- Task I - Screening, to define the most promising integration scheme for each gasification process.
- Task II - Developing a process flowsheet, heat and material balance, P&ID, equipment specifications, utility summary, and plot plan for the selected scheme of each process.
- Task III - Preparing the detailed economics and issuing the final report.

This interim report documents the results of the Task I study. The acid gas removal processes under study included Selexol, Rectisol, and Benfield HiPure for the Exxon CCG process; MEA, DEA, improved DGA, and Sulfinol for the Rockwell Hydrogasification process. The evaluations were performed using data from the prime coal gasification contractors and from the vendors of proprietary acid gas removal processes. This information, combined with Air Products' in-house capabilities in acid gas processes and cryogenic separation, allowed preparation of process designs and cost estimates for each of the integrated schemes. The gas treating cost was then determined for each combined system.

The design criteria used in preparing the process designs are listed in the criteria section of this report. The basis of economics for the capital cost and the gas treating cost estimate are also given in the criteria section.

Exxon

For the Exxon CCG plant producing 257 MMSCFD of SNG, the integrated Selexol/cryogenic separation process has been selected over the Benfield and Rectisol approaches. Results of the screening study indicate that the Selexol/cryogenic separation alternative is the most economically attractive method for H_2S and CO_2 removal and CH_4 purification. Although the integrated Selexol/cryogenic process requires a slightly larger initial capital investment than the integrated Benfield process, the operating cost savings are more than sufficient to overcome this penalty. On a 20-year average utility financing basis, the costs associated with purifying and separating the feed gas stream (treating costs) are $0.856/MSCF for the Selexol/cryogenic separation alternative; 11.5% lower than the costs for Benfield, and 14.9% lower than Rectisol. Evaluating the alternative processes using private investor financing methods produces similar results. Sensitivity analysis of investment, power, and steam costs indicates that the selection is relatively insensitive to the estimated values for these costs. Therefore, on the basis of requiring the lowest treating costs, the Selexol/cryogenic alternative has been chosen as the best process to use with the Exxon CCG scheme.

The Selexol/cryogenic separation process currently proposed represents significant capital and utility savings over the Selexol/cryogenic separation process included by Exxon in its own total facility estimate(1). A capital cost of $156,300,000 for the Air Products' designed process is approximately 25% lower than that proposed by Exxon. In addition, annual power, steam and cooling water costs, totaling $40,320,000/year, also represent a reduction of approximately 25% of the corresponding utility costs presented by Exxon. Based on the work that has been completed for the screening study, capital and operating costs can be reduced further through additional optimization steps in both the acid gas removal and cryogenic sections of the plant. These optimization steps will be evaluated in Task II.

Rockwell

For the Rockwell Hydrogasification process producing 259 MMSCFD of SNG, the improved DGA/cryogenic separation process has been selected. Results of the screening study indicate that the DGA/hypersorption combination is closely competitive with the DGA/cryogenic separation alternative. Based on a 20-year average utility financing analysis, the gas treating costs are $0.691/MSCF and $0.660/MSCF for DGA/hypersorption and DGA/cryogenic, respectively. Because of the uncertainty associated with the adsorption operation of a large moving carbon bed, the high fuel consumption, and the slightly higher treating cost, the DGA/hypersorption process is judged less attractive. Therefore, the DGA/cryogenic alternative has been chosen as the best combined process for the Rockwell Hydrogasification scheme.

2. INTRODUCTION

The purpose of this interim report is to document the Task I evaluation of the combinations of acid gas removal and methane Catalytic separation steps for both the Exxon Coal Gasification (CCG) and the Rockwell Hydrogasification processes. Progress to date includes completion of the screening type process designs, preliminary equipment sizes, utility requirements, and cost estimates for seven acid removal processes and the integrated methane separation schemes.

For the Exxon CCG process, one chemical absorption process, Benfield HiPure, and two physical absorption processes, Selexol and Rectisol, were evaluated to determine the feasibility of their integration with cryogenic methane separation.

Four chemical absorption processes, MEA, DEA, improved DGA, and Sulfinol were evaluated for the Rockwell Hydrogasification process due to the low acid gas partial pressure, where physical absorption processes are not attractive. Two types of methane separation, cryogenic and moving bed hypersorption were studied for integration with acid gas removal. This resulted in eight (8) different combinations for screening for the Rockwell case.

This report describes the design criteria, process block diagrams, heat and mass balance at the battery limits, economic criteria, integration basis, and the supporting economics. Trade-off parameters to be considered for Task II are also discussed.

2.1 PROCESS DESCRIPTION AND DESIGN CRITERIA

2.1.1 EXXON CCG PROCESS

In the Exxon CCG process, catalyst is added to the feed coal and the mixture is gasified with the preheated recycle H_2/CO stream at about 1300°F and 500 psia. [With the promotion of K_2CO_3 catalyst, the gasification rate is high enough to yield a high methane concentration.] Since the amount of H_2/CO fed balances the amount of H_2/CO leaving the gasifier, the net products of gasification are only methane and CO_2 along with smaller amounts of H_2S, NH_3 and other trace components. Disregarding heat loss and the energy for coal and CO/H_2 preheating, the overall reaction is thermally neutral. The raw gas leaving the gasifier is cooled, scrubbed, and then further cooled and fed to the acid gas removal unit in which H_2S and CO_2 are removed down to acceptable levels for methane purification in the cryogenic unit. Carbon monoxide and hydrogen are separated from methane and recycled back to the gasifier. Product methane is compressed and sent to the pipeline for sale. A simple schematic flow plan of this process is shown in Figure 2.1.1.

2.1.1.1 PROCESS DESIGN CRITERIA

The process design criteria included definition of the coal, the cooled raw gas composition, recycle gas condition, product gas condition, environmental emission standards, pipeline specifications, and plant site characteristics.

Illinois No. 6 coal was used and is defined by the analysis shown in Table 2.1.1.1a. This coal satisifies the requirement by DOE to consider an Eastern bituminious coal. The cooled raw gas composition and recycle gas condjtion are given in Table 2.1.1.1b. The above data were provided by Exxon (2). Table 2.1.1.1c shows the pipeline gas specifications developed by Air Products.

Plant site characteristics are shown in Table 2.1.1.1d. The emission standards are shown in Table 2.1.1.1e. Both tables have been submitted by Air Products to DOE (3).

Fuel, steam, and electric power with appropriate characteristics were assumed to be available at the plant site. Cooling water was also assumed to be available to the plant at 85°F, with 20°F rise and 20 psi allowable pressure drop.

2.1.1.2 GAS IMPURITIES

In addition to carbon dioxide and hydrogen sulfide, there are many other trace impurities produced in the gasifier, depending on the gasifier operating conditions and the type of coal. These possible impurities include carbonyl sulfide, carbon disulfide, mercaptans, organic acids, light oils, phenols, ammonia, nitrogen oxides, hydrogen cyanide, chlorides and particulates. Among these, carbonyl sulfide and carbon disulfide have significant impact on some acid gas treating systems.

For the Exxon gasification process, no information on trace impurities is available at present from experimental or pilot plant tests. However, the process includes two stages of water scrubbing and one COS hydrolysis unit suggested by Exxon to condition the raw acid gas prior to entering the acid gas removal system. With this arrangement, it is believed that ammonia, organic acids, phenols, chlorides, hydrogen cyanide, and dust particles, etc., will be removed down to very low levels by water scrubbing; carbonyl sulfide, carbon disulfide and mercaptans will be converted to hydrogen sulfide by the COS converter. As a result, the raw acid gas to the acid gas removal unit will be relatively clean. Therefore, the gas composition given by Exxon was adapted and used for the study of the acid gas removal/cryogenic methane separation integration.

2.1.2 ROCKWELL HYDROGASIFICATION PROCESS

The Rockwell process is based on a concept of producing high yield methane by direct reaction of coal with hydrogen at high mass flux and short residence time. The unique feature is the application of injection/mixing techniques originally used by Rocketdyne in very high throughput rocket combustors, to the rapid mixing and reaction of hot hydrogen and pulverized coal.

Figure 2.1.2 shows the simple block flow diagram proposed by Rockwell. The combined make-up and recycle hydrogen is preheated to 1600°F and then partially combusted with O_2 raising its temperature to a level such that the hydropyrolysis reaction rate can be achieved when it is mixed with pulverized coal. The reactor effluent is rapidly quenched after char separation. The gas leaving the separator is processed through the water scrubber to remove solids and condensable components.

The cooled raw gas is then conditioned in the acid removal unit to remove CO_2 and sulfur compounds. The treated gas is further processed to recover H_2 for recycle. A trim methanator is employed to convert trace CO and H_2 to CH_4 to meet the pipeline specifications.

2.1.2.1 PROCESS DESIGN CRITERIA

Most of the process design criteria used for the Exxon CCG process are applied here, except for the unique coal data, the cooled raw gas composition, and the recycle gas condition.

The coal used was Kentucky #9/14 and is defined by the analysis shown in Table 2.1.2.1a. This coal is similar (on a dry basis) to the Illinois #6 used in the Exxon CCG study and satisfies the requirement by DOE to consider an Eastern bituminous coal. The cooled raw gas composition and recycle gas conditions are given in Table 2.1.2.1b. The above data were provided by Rockwell International (4).

2.1.2.2 GAS IMPURITIES

At present, no information on trace impurities is available from experimental or pilot plant tests for the Rockwell Hydrogasification process. In order to investigate the quantitative impact of the trace impurities on the acid gas removal system, a NASA JANAF chemical equilibrium program was used to estimate the trace impurities. The estimate is as follows:

	ppmv
COS	10.5
CS_2	0.13
C_2H_2	0.88
HCN	20.9

These impurities were used to supplement the cooled raw gas composition, provided by Rockwell, to study the combination of acid gas removal and methane separation.

2.2 ECONOMIC CRITERIA

2.2.1 CAPITAL INVESTMENT

The capital requirements for all processes were estimated based on a consistent set of criteria. Table 2.2.1a shows these criteria. Total capital for the combined system of acid gas removal and methane separation includes process plant investment, initial chemicals and catalyst charges, paid-up royalties, allowance for funds during construction, start-up costs, and working capital.

All of the process plant investments were estimated according to the preliminary equipment sizing information provided by vendors and Air Products. Most of the estimate was developed by scaling from in-house cost information. Cost information from Air Products recently completed analysis of a Selexol process for a large coal gasification facility was scaled. Vendor quotations for major

major equipment items (columns and heat exchangers) confirmed Air Products' estimate. These quotations also served as a basis for the current estimates. Cryogenic section capital costs were based on process designs and equipment sizing information developed in-house. Estimates were developed using historical information for similar adsorber/drier systems, cold boxes, compression trains and their component parts. For the downstream sulfur recovery plant, an estimate solicited from the Ralph M. Parsons Co. was incorporated directly in determining the sulfur plant penalty assigned to the Benfield process. With Parsons' experience in this area, it was judged that they would be able to provide the most accurate sulfur plant estimate.

The Lotepro Corporation and the Benfield Corporation provided capital investment estimates for the Rectisol and Benfield acid gas removal processes, respectively. However, these estimates were not incorporated directly into Air Products' estimate. In order to ensure that each process was being evaluated on the same basis, Air Products independently prepared the estimate for each process. With this approach a greater degree of credibility can be placed on the results; the estimates are totally consistent. The Lotepro and Benfield estimates were used as a means of checking Air Products' estimate.

2.2.2 OPERATING COST

The criteria used in the annual operating cost estimate are given in Table 2.2.2a. To develop operating costs necessary to evaluate the alternative processes, values were assigned to each utility. Table 2.2.2b lists the values assumed for the utilities, absorbents, adsorbents, and catalysts. The values assigned for power, steam, cooling water, fuel, and boiler feed water represent current values (March 1979) typically used by Air Products for this type of facility. Refinement of utility costs would require considerably more information than is currently available, especially with regard to plant location and utility sources. It is believed, however, these values represent reasonable assumptions for the screening analysis.

2.2.3 FINANCING CRITERIA

In order to determine the most promising process for each coal gasification scheme, capital and operating costs must be put on the same basis. One method of accomplishing this is to annualize the capital investments by applying capital charges that effectively convert capital investment into pseudo-operating costs. Capital charges include depreciation, interest, taxes, and return on equity. The financing method affects the calculation of the capital charges.

There are several methods available for determining capital charges. For a large coal gasification facility, utility financing is the method most often used. Coal gasification facilities usually require enormous initial capital outlays. Utility financing methods tend to reduce the importance of capital expenditures because the required return on capital invested is a mixture of the returns required for debt and equity. The required return for debt is substantially lower than that for equity. In addition, the return on debt capital is tax deductable. Table 2.2.3a summarizes the assumptions that form the basis of the utility financing method (CCU-A.G.A) used in this evaluation.

An alternative approach for comparing competing processes is discounted cash flow analysis. Table 2.2.3b describes the assumptions used for this private investor financing method.

The CCU-A.G.A. private investor financing method, which does not account for escalation, results in significantly higher capital charges than those resulting from utility financing methods. Generally, the price of SNG or any other industrial or commercial gas includes an escalation formula related to some base; for example, energy costs; or treating costs or some general price index. To account for this, Air Products' DCF method includes reasonably expected escalation rates of 8% on operating revenues and 7% on operating costs. With revenues increasing at a greater rate than costs, yearly cash flows increase over the non-escalated case. As a result, gas treating costs are reduced. In addition, Air Products' DCF calculation assumes cash flows occur continuously throughout the year, as opposed to the year end, as is assumed in the CCU-A.G.A method. The effect of increased discounted cash flows is a reduction in the required treating costs associated with reduced capital charges.

APPENDIX A

FIGURE 2.1.1:
EXXON CATALYTIC GASIFICATION PROCESS

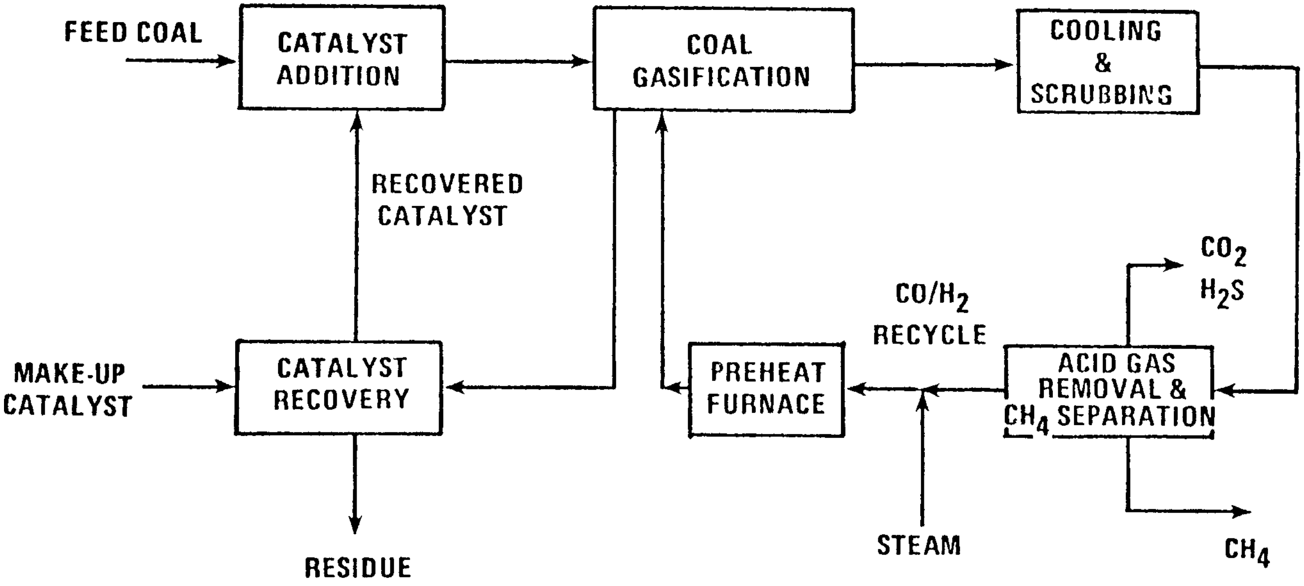

Table 2.1.1.1a: Illinois #6 Coal Analysis (2, 3)

		wt %, dry
C		69.67
H		5.05
O		9.45
N		1.84
S		4.19
Cl		0.08
Ash		9.72
	Total	100.00

Moisture (as received) 16.5 wt %

Table 2.1.1.1b: Cooled Raw Gas, Fuel Gas, and Recycle Gas Conditions (2, 3)

Cooled Raw Gas:

Temp. 120°F
Pressure 450 psia

	lb-mole/hr	mole %
CO	10,510.0	9.524
CO_2	21,633.0	19.603
H_2	37,671.0	34.136
H_2O	416.0	0.377
CH_4	35,657.0	32.311
C_2H_6	5.0	0.004
H_2S	878.0	0.796
COS	2.0	0.002
N_2	3,585.0	3.247
Total	110,357.0	100.000

Recycle Gas:

Temp. 175°F
Pressure 600 psia

	lb-mole/hr	mole %
CO	10,472.0	18.204
CO_2	44.0	0.076
H_2	37,665.0	65.476
H_2O	12.0	0.021
CH_4	5,747.0	9.990
N_2	3,585.0	6.233
Total	57,525.0	100.000

Fuel Gas:

Temp. 105°F
Pressure 45 psia

	lb-mole/hr	mole %
CO	2.0	0.124
CH_4	1,611.0	99.876
Total	1,613.0	100.000

Table 2.1.1.1c: Pipeline Gas Specification

Pressure	1,015 psia
Temperature	120° F

Max. CO	0.5 mole %
Max. CO_2	0.1 mole %
Max. H_2	5.0 mole %
Max. H_2O	1.8 lb/MMSCF* (0°F dewpoint @ 1,015 psia)
Max. H_2S	0.25 Grain/100 SCF
Max. Total Sulfur	1.00 Grain/100 SCF
Max. N_2	2.0 mole %
Min. HHV	950 Btu/SCF*

*Btu and SCF measured at 60°F and 14.7 psia

Table 2.1.1.1d: Plant Site Characteristics

Based on Middletown, U.S.A. site.

Site elevation 0 - 50 ft above river level.

Adjacent to navigable river.

5 miles from main railroad line.

Design ambient pressure at 14.7 psia.

Design ambient temperature conditions:

Standard Day

52°F Wet Bulb
59°F Dry Bulb
60% Relative Humidity

Hot Day

76°F Wet Bulb
94°F Dry Bulb
45% Relative Humidity

Table 2.1.1.1e: Emission Standards

Contaminant	Project Emission Limit	Basis
CO	500 - 2000 vppm	1,2
CO_2	assumed no restrictions	
H_2	assumed no restrictions	
CH_4	assumed no restrictions	
H_2S	10 vppm for direct venting 160 vppm if gas is burned or flared	1,2,3
H_2S, COS, CS_2	300 vppm total reduced sulfur compounds (100 vppm COS alone)	1
NOx (as NO_2)	125 - 300 vppm	2,4,5
NH_3	25 vppm	3
HCN	10 vppm	3
C_2H_2, C_2H4 C_2H_6+	25 vppm total nonmethane hydrocarbons*	4
Benzene	10 vppm	6
Mercaptans (as CH_3SH)	0.2 vppm	4
Particulates	0.01-0.08 gr/dscf corrected to 12% CO_2 Ringlemann 1 (20% opacity)	2,3,5,6,7
SO_2	250 - 500 vppm	3
HCl	5 vppm	3
*Quantities greater than 100 lb/day may have to be flared or incinerated, e.g., as a minimum 1300°F and 0.3 seconds residence time.		8

Notes:

1. EPA NSPS Petroleum Refining
2. Los Angeles Air Pollution Control District
3. New Mexico Gasification Plant Standards
4. San Francisco Bay Area Air Pollution Control District
5. EPA NSPS for Fossil Fuel Steam Generators (see discussion)
6. EPA NESHAPS for vinyl chloride
   OSHA limit for worker exposure for 8-hr. day is 1 vppm ambient concentration of benzene.
7. Pennsylvania Standards
8. Texas Air Control Board Regulations

FIG. 2.1.2: ROCKWELL HYDROGASIFICATION PROCESS DIAGRAM

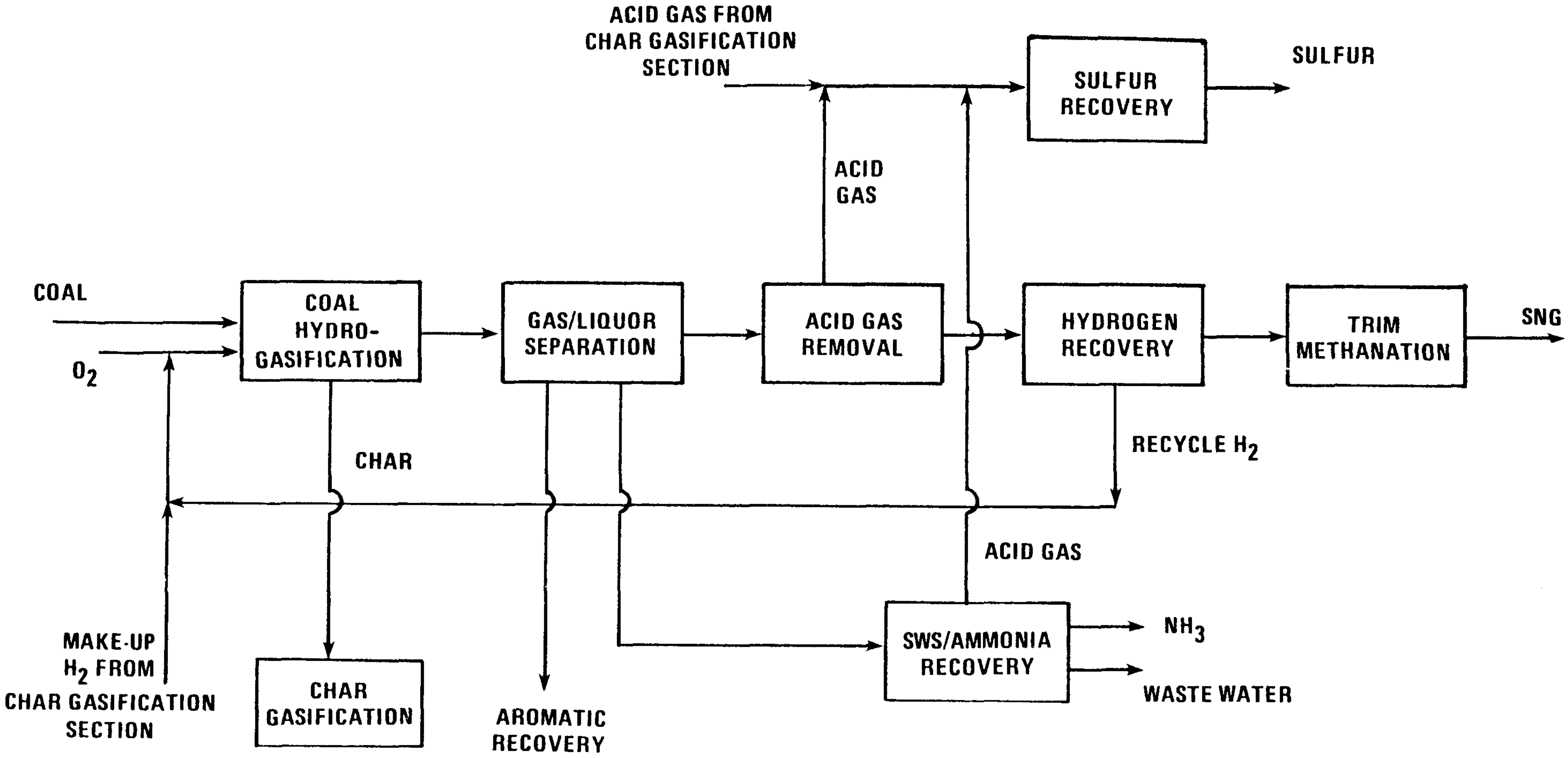

Table 2.1.2.1a: Kentucky #9/14 Coal Analysis (4)

	wt %, dry
C	69.77
H	4.90
O	8.42
N	1.53
S	4.30
Cl	0.05
Ash	11.03
Total	100.00

HHV: 12,567 Btu/lb dry

Moisture (as received) 1.88 wt %

Table 2.1.2.1b: Cooled Raw Gas and Recycle Gas Conditions (4)

Cooled Raw Gas:

Temp. 110 °F
Pressure 1,420 psia

	lb-mole/hr	mole %
CO	2,230.0	2.388
CO_2	117.0	0.125
H_2	63,493.0	67.986
H_2O	4.0	0.004
CH_4	26,872.0	28.773
H_2S	392.0	0.420
N_2	281.0	0.301
NH_3	3.0	0.003
Total	93,392.0	100.000

Recycle Gas:

Temp. Maximum
Pressure 1,615 psia

	lb-mole/hr	mole %
H_2	55,743.0	96.192
N_2	211.0	0.364
CH_4	1,996.0	3.444
Total	57,950.0	100.000

Table 2.2.1a: Capital Cost Estimate Criteria

Basis	. 1979 dollars . Clear and level plant site . Excluding land cost

Item	
Total Plant (process) Investment (TPI)	(1) Installed equipment cost (2) Engineering fee (3) Contractor's fee at 3% of (1) + (2) (4) Contingency at 15% of (1) + (2) + (3) TPI = (1) + (2) + (3) + (4)
Paid-up Royalty (PR)	From Process Developer or Vendor
Allowance for Funds used during Construction (STI)	STI (Short-Term Interest) = 10% (average spending period) x (TPI+PR+SC) Average spending period = 0.9825 yrs. for a 3-yr construction period
Start-up Costs (SC)	20% of gross operating cost
Working Capital (WC)	W.C. = 0.9% (TPI) + 1/24 (annual treating cost)
Total Capital Requirement (TCR)	TCR = TPI + PR + STI + SC + WC

Table 2.2.2a: Operating Cost Estimate Criteria

Basis: 90% on-stream factor

Labor	
Operating	20 Men @ $30M/Man/yr
Maintenance	1.8% of Total Plant Investment (TPI)
Supervision	20% of (Operating and Maint. Labor)
Supplies	
Operating	30% of Operating Labor
Maintenance	1.2% of Total Plant Investment (TPI)
Taxes and Insurance	2.7% of Total Plant Investment (TPI)
Administration & General Overhead	60% of Total Labor

Table 2.2.2b
Assumed Values for Utilities, Absorbents and Adsorbents and Catalysts

	Value
Utilities	
Power	$ .035/kwh
Steam - 250 psig, 700°F	$4.78/M lbs
500 psig, Sat.	$4.37M lbs
150 psig	$3.50/M lbs
50 psig	$2.80/M lbs
15 psig	$2.30/M lbs
Cooling Water	$ .10/M gal circ.
Fuel	$2.50/MM Btu
Hydrogen	$2.50/MM Btu
Carbon Monoxide	$2.50/MM Btu
Methane	$3.50/MSCF
Boiler Feed Water	$ .15/M gal circ.
Absorbents	
Selexol	$8.75/gal
Methanol	$ .55/gal
MEA	$ .41/lb
DEA	$ .42/lb
DGA	$0.67/lb
Sulfonane	$1.15/lb
DIPA	$ .46/lb
K_2CO_3	$ .22/lb
Stretford Solution	$ .23/gal
Adsorbents and Catalysts	
Molecular Sieve	$1.90/lb
Iron Sponge	$5.00/$ft^3$
Claus Catalyst	$600/ton
Beavon Catalyst	$ 63/ft^3
Methanation Catalyst	$150/$ft^3$
Activated Carbon - Calgon BPL 4 x 10 mesh	$1.10/lb
- 6-32W 4 x 10 mesh	$1.73/lb
Alumina H-151	$0.79/lb
F-1	$0.34/lb

Table 2.2.3a: Utility Financing Assumptions

1. Project Life - 20 years
2. Depreciation - 20 year straight line depreciation on plant investment including allowance for funds used during construction and capitalized portion of start-up costs.
3. Financing - 75% Debt/25% Equity
4. Interest Rate - 10% of Average Net Investment
5. Return on Equity - 15%
6. Tax Rate (Fed., St., Loc.) - 49.5%
7. Average Net Investment - .75 (.5 (Investment) + Working Capital)
8. Method - CCU - A.G.A. Method as described by C. F. Braun and Company (5)

Table 2.2.3b: Private Investor Financing

1. Project Life - 20 years
2. Depreciation - 16 years sum-of-years-digits on plant investment including return on investment during construction and capitalized portion of start-up costs.
3. Financing - 100% Equity
4. DCF Return - 15%
5. Tax Rate (Fed., St., Loc.) = 49.5%
6. Escalation - Treating Revenues - 8%/year
   Gross Operating Costs - 7%/year (excluding property taxes at 2.2% of TPI)
7. Method - APCI analysis (with escalation) and CCU - A.G.A. Method (without escalation)

3. ACID GAS REMOVAL AND CRYOGENIC METHANE SEPARATION

3.1 EXXON CCG PROCESS

The acid gas partial pressure of the Exxon raw acid gas stream, about 92 psia, is believed to be in the intermediate range between the high partial pressures, for which physical absorption is favored, and low partial pressures, which favors chemical absorption. However, the ratio of CO_2 to H_2S is about 25. Therefore, the physical absorption processes are more favorable due to their selective removal of H_2S, COS and CO_2, thereby meeting the requirements for sulfur concentration in the off-gas stream for conventional claus sulfur recovery. The capability to removal total sulfur down to very low levels is another advantage of using physical absorption. On the other hand, the improved Benfield (selective H_2S removal plus Hi-Pure unit) chemical absorption process is worth investigating, because it is capable of removing sulfur down to very low levels and it requires low capital cost. For process screening, the field of acid removal processes was narrowed down to Selexol, Rectisol, and improved Benfield Hi-Pure.

Total sulfur and CO_2 concentrations leaving in the treated gas from the acid gas removal system were specified at 1 and 100 ppmv, respectively. Of course, CO_2 and water vapor have to be further conditioned to acceptable levels prior to charging to the cryogenic unit. The sulfur content of the SNG product must also meet the pipeline gas specifications. The above limits of sulfur and CO_2 from the acid gas system are practical and satisfactory, though not necessarily optimum. These limits may be considered as a separate parameter for futher optimization study.

3.1.1 SELEXOL AND CRYOGENIC INTEGRATION

3.1.1.1 SELEXOL ACID GAS REMOVAL

Selexol is a physical absorption process which employs dimethyl ether of polyethylene glycol as a solvent. Air Products has developed one of the most sophisticated Selexol computer simulation routines capable of performing rigorous tray-to-tray column and various process unit calculations for detailed design.

For this application, the Selexol design is a four train design involving two stages of acid gas removal. In the first column, H_2S is selectively removed from the CO_2-rich feed gas ($CO_2/H_2S = 25$) by CO_2 preloaded solvent in such a way that the H_2S content in the off-gas stream is somewhat higher than 35%, suitable for conventional Claus plant sulfur recovery. The remaining CO_2 is then removed by the second column of solvent absorption. The absorbed CO_2 is stripped by using air and is directly vented into the atmosphere.

The coabsorbed H_2, CO, and CH_4 are recovered by intermediate pressure flashing. The flash gas is then compressed and recycled back to the absorber. The flash pressure level may be considered as a parameter for optimization study in Task II.

3.1.1.2 CRYOGENIC METHANE SEPARATION

The process gas from the Selexol unit is sufficiently cooled (5°F) to be fed directly to an adsorption bed system to remove residual water vapor and carbon dioxide. Removal of these contaminants is necessary to prevent freezing in the cryogenic unit, or cold box.

The dry feed gas is partially condensed as it is cooled against the cold box effluent streams. The feed also derives refrigeration from a stripper column by supplying the heat duty for the reboiler. The cold feed vapor phase, which is the major portion of the H_2/CO recycle stream, is separated from the liquid phase, which is principally CH_4. The liquid stream containing dissolved H_2 and CO, enters a stripping column where the H_2 and CO are removed from the CH_4. The column overheads warms against feed flow and, after compression, joins the H_2/CO recycle stream. The column bottoms which is 99.9% CH_4 is the SNG product. A small portion of bottoms CH_4 is flashed to slightly below atmospheric pressure to supply low temperature refrigeration to the feed gas. The remaining portion is pumped to a higher pressure before being vaporized and warmed against the feed.

The two H_2/CO effluent streams from the cold box are compressed to 600 psig for return to the gasifier. A portion of the two effluent SNG streams is supplied as fuel to the CCG process while the remainder is compressed to 1015 psig for delivery to the pipeline.

3.1.1.3 INTEGRATION SCHEME

The simple process block flow diagram of the Selexol and cryogenic integration scheme is shown in Figure 3.1.1.3a. Table 3.1.1.3a and Table 3.1.1.3b show the material balance at the battery limits and the utility requirements, respectively, for this combined system.

3.1.1.4 DISCUSSION

Selexol is a commercially proven process developed by the Allied Chemical Corporation. Selexol has been extensively applied to gas treatment of natural gas and synthesis gas from POX (partial oxidation) since 1965. However, there is no commercial experience on the operation of a Selexol plant associated with coal-derived fuel gas. The effects of the trace impurities on the performance cannot be fully assessed.

If the COS content is high in the acid gas feed, it usually presents two unfavorable design problems:

- High solvent circulation rate resulting in high utility consumption.
- Less flexibility on sulfur absorption selectivity, resulting in CO_2 dilution of the acid off-gas for sulfur recovery.

In the CCG process as presented by Exxon, COS has been hydrolyzed prior to the acid gas removal unit. As a result, COS does not cause any extraordinary problem for direct CO_2 venting.

For utility savings, air has been proposed for stripping since there is no nitrogen available for the Exxon CCG process. Air stripping can present two potential problems, elemental sulfur precipitation and O_2 slippage into the downstream cryogenic unit.

With regard to the potential sulfur problem, Allied's current guideline is that they believe there is no technical risk with air stripping when the feed gas contains up to 4 ppm total sulfur. As Allied has no firm commerical operating experience, this is their best recommendation. Air Products has estimated that the Allied feed criterion applied to existing commercial plants would translate to a rich solvent concentration of between 3 and 20 molar ppm total sulfur. The Exxon CCG design considered here had 15 ppm total sulfur in the solvent. This was considered reasonable by both Allied and Air Products, and the design continued on this basis, acknowledging some small uncertainty here. Allied noted that the addition of a scraped wall chiller could be considered as a contingency measure to deal with sulfur formation, should it present a problem.

With regards to the O_2 safety concern, normal operation would be totally safe; the heat and mass balance simulation shows that the amount of oxygen dissolved in the Selexol and subsequently transferred to the treated gas is nil. While abnormal operation could result in higher oxygen levels, it is believed that an in-depth process hazards review of the engineering flowsheet would minimize risks to acceptable levels. ASED, N.V., an affiliate of Carbo-Chemique at the Willebreak complex, Belgium, has a facility of POX, Selexol, and cryogenic separation. Waste N_2 containing about 4% O_2 is used for Selexol solvent stripping. No safety problems have been experienced in the cryogenic N_2 wash unit downstream. This is, of course, a single positive experience. The risk assessment of air stripping shall be further investigated. In any event, demonstration tests would ultimately be in order before firm committment to an air stripping design.

As an alternative to avoid the above-mentioned concerns, vacuum stripping may be considered as an alternate.

The cycle used for the cryogenic separation of CH_4 employs proven technology. All of the equipment used here is commercially available.

3.1.2 RECTISOL AND CRYOGENIC INTEGRATION

3.1.2.1 RECTISOL ACID GAS REMOVAL

Rectisol is a physical absorption process utilizing methanol as the solvent to absorb acid gases at a relatively low temperature. Because of the low temperature operation, the solubilities of CO_2, H_2S and COS in methanol are high, and result in a low solvent circulation rate.

The raw acid gas, saturated with water vapor, is indirectly cooled by the SNG product from the cryogenic unit, and then cooled by the cold purified gas. Icing is prevented by methanol injection before cooling. Carbon dioxide, COS and H_2S are removed in the absorber. The rich solvent is further cooled, and then stripped with a gas stream (of low CH_4 content) at intermediate pressure to recover CH_4. The stripped gas is compressed back to the absorber. As with Selexol, gas solubility is a function of the solvent, pressure, temperature, and the feed gas composition, therefore the difference in solubility between CO_2 and sulfur compounds in the physical solvent makes it possible to separate the CO_2, suitable for direct venting, and to concentrate the H_2S in the acid off-gas stream, suitable for conventional Claus treatment. Since no inert gas, such as nitrogen, is used in the Exxon CCG process, vacuum flash is utilized to remove the bulk CO_2 in order to minimize steam consumption. Vacuum flash was recommended by Lotepro, as air stripping of methanol is not considered a safe alternative. The rich solvent leaving the flash tank is pumped to the stripper. The remaining acid gas is stripped by using steam, and then treated by a conventional Claus plant to recover sulfur.

A quotation for a four train Rectisol design was obtained from the Lotepro Corporation. Lotepro provided the installed capital cost, utility requirements, mass balance at the battery limits, and equipment sizes. Air Products used this information to estimate the total capital requirement on the same basis as that used for the Selexol and Benfield estimates.

3.1.2.2 CRYOGENIC METHANE SEPARATION

The adsorption bed system and cold box are identical to the system used with the Selexol unit (see 3.1.1.2) integration, with one exception. The effluent streams from the cold box are returned to the Rectisol unit for recovery of refrigeration before being compressed and delivered to the gasifier and pipeline.

3.1.2.3 RECTISOL AND CRYOGENIC INTEGRATION SCHEME

Figure 3.1.2.3a shows the simple process block diagram of the integrated scheme for Rectisol and cryogenic separation. Table 3.1.2.3a and Table 3.1.2.3b show the mass balance at the battery limits and the utility requirements, respectively, for this combined scheme.

3.1.2.4 DISCUSSION

Rectisol is a commercial, proven process and has been employed in coal and oil gasification plants producing fuel gas, synthesis gas, and SNG. It is also widely used in conjunction with low temperature plants for hydrogen purification.

Rectisol has many significant advantages. The solvent, methanol, is inexpensive, generally non-corrosive, readily available, and is a drying agent. Equally important, it is not degraded by trace components such as HCN, NH_3, CS_2, COS, etc., produced from coal gasification. A 0.1 ppmv total sulfur (including H_2S, COS, CS_2 mercaptan, etc.) specification in the treated gas stream is easily achievable. From this viewpoint, Rectisol is particularly attractive for high pressure absorption applications in conjunction with synthesis gas production which requires high purity.

Because of low temperature operation, the capital investment and utility costs for Rectisol are relatively higher than the other processes. Generally, an inert gas, such as N_2, can be used for CO_2 stripping to save steam consumption. Nitrogen has to be very pure (O_2 3 ppm, guideline from Lotepro) due to the following concerns:

a. Potential safety hazard for the mixture of methanol and O_2

b. Elemental sulfur precipitation

c. Safety problem associated with O_2 getting into the downstream cryogenic unit.

Since N_2 is not available, vacuum stripping is employed to remove the bulk of the CO_2 absorbed by the Rectisol solvent.

3.1.3 BENFIELD AND CRYOGENIC INTEGRATION

3.1.3.1 BENFIELD ACID GAS REMOVAL

Benfield is a chemical absorption process which employs potassium carbonate as the absorbent. An activated hot potassium carbonate solution is used in the Benfield Hi-Pure process to remove sulfur down to very low levels.

The raw acid gas is fed into the absorber of a selective sulfur removal unit. The reaction rate for H_2S absorption is 15 to 40 times faster than that of CO_2. This permits selective absorption of the H_2S. The absorbed rich solution is stripped by steam, and the acid off-gas containing about 12% of H_2S is treated with a Claus plant followed by tail gas treatment. The gas leaving the absorber of the selective sulfur removal unit is further conditioned with the Hi-Pure unit. The high purity of the treated gas is achieved in a special arrangement in which two independent solutions are used in two countercurrent circuits. The absorbed acid gas is also stripped here using steam. Because of the low H_2S content, this acid off-gas is combined with the Claus tail gas, and the combined flow is treated with the Beavon process to recover the sulfur. The treated gas leaving the Hi-Pure unit contains concentration of total sulfur and CO_2 less than 4 and 100 ppmv, respectively. Potential additional features that can be incorporated in the designs to reduce operating costs include flash cooling techniques and the use of hydraulic turbines.

A quotation for a selective sulfur plus Hi-Pure system for the Exxon CCG application was received from the Benfield Corporation. Benfield furnished the installed capital cost, utility requirements, mass balance, equipment sizes, and economic breakdown information on their proposed three train basis. Because of the large column sizes, Air Products consulted with Benfield and redesigned the Hi-Pure unit on a four train basis. Also, the recommended polypropylene pall ring packing was changed to stainless steel packing due to the high temperature operation. Air Products reestimated the installed capital cost based on the same economic criteria as that used for the Selexol and Rectisol processes. Therefore, all three processes were evaluated on a comparable basis.

3.1.3.2 SULFUR TRIM

Because the Hi-Pure system cannot reduce the sulfur to meet the SNG product specifications, an iron sponge bed sulfur trimmer was employed to reduce the sulfur content in the treated gas stream leaving the Benfield process to less than 1 ppmv. The design and cost estimate of the iron sponge unit were prepared in-house.

3.1.3.3 CRYOGENIC METHANE SEPARATION

The nature of the Benfield system makes it difficult to integrate with a cryogenic unit. The feed gas from the Benfield unit is relatively warm (132°F), therefore it is cooled in three stages before entering the adsorption bed system. The feed is precooled first against cooling water, then the cold box effluents, and finally against an external refrigeration system.

The cold box separation process is identical to the system used with the Selexol unit (see 3.1.2) integration.

The cold box effluent streams are used to supply refrigeration to precool the feed to the adsorber, before being compressed for delivery to the gasifier and pipeline.

3.1.3.4 INTEGRATION SCHEME

The simple process block diagram of the integrated scheme for Benfield and cryogenic methane separation is shown in Figure 3.1.3.4a. Table 3.1.3.4a and Table 3.1.3.4b show the mass balance at the battery limits and the utility requirements, respectively, for this integrated system.

3.1.3.5 DISCUSSION

The Benfield process has been widely used for effective and economic removal of CO_2 and H_2S. So far about 380 Benfield plants have been built or are under construction. About 165 of them are using the activated Benfield process. Principal applications have been in plants producing ammonia, hydrogen, town gas, and SNG from coal and oil gasification. The Benfield process was used to treat raw gas produced from coal by the Lurgi process at Westfield, Scotland for over 13 years. Pilot plant and commercial data have verified that contaminants such as COS, CS_2, mercaptans, HCN, NH_3, etc., can be handled without causing significant corrosion and solution degradation.

Carbonyl sulfide and carbon disulfide are hydrolyzed in activated potassium carbonate to form CO_2 and H_2S. The rate of hydrolysis is highly dependent on temperature. The absorption of CS_2 is expected to be slower than COS because of two steps of hydrolysis involved. The removal efficiencies of COS and CS_2 have varied from 70 to 99%. The degree of mercaptan absorption is reported to be dependent on the equilibrium vapor pressure of the solution. Therefore, the absorption efficiency depends on the quantity of mercaptans in the feed gas and on the solution circulation rate. Other trace components, such as SO_2, HCN, and NH_3 are removed at high efficiencies.

For some applications, the Benfield process requires less thermal energy than amine scrubbing systems and significantly lower capital investment than processes using physical solvents. Because of its capability to handle trace components and operate at high temperature, the Benfield process is attractive in the low Btu coal-fired combined cycle power plant applications. However, the high water vapor content in the treated gas (due to high temperature operation) make it somewhat less attractive for the integration of Benfield and cryogenic methane separation.

For the Exxon CCG process, the Benfield process generates a low H_2S concentration in the acid off-gas stream to the sulfur plant. It requires a more expensive sulfur recovery facility in order to meet the pollution emission standards.

3.1.4 SULFUR RECOVERY

Selexol and Rectisol produce only one acid off-gas stream containing concentrated H_2S (about 35%) which can readily be processed in a conventional Claus unit followed by a Beavon unit. On the other hand, Benfield Hi-Pure produces two acid off-gas streams, a dilute Claus gas (about 12% H_2S) and an H_2S tail gas. A more expensive sulfur recovery plant is required to treat the Benfield acid off-gas due to the low H_2S concentrations. R. M. Parsons Co. provided installed capital and utility requirements for the Benfield and Selexol (Rectisol) sulfur recovery units. Since sulfur recovery is outside the scope of this contract work, detailed balances and economics are not presented here; instead, a sulfur plant penalty is assigned to the Benfield process representing the additional capital and utilities required for the sulfur removal unit. This puts the three acid gas removal processes on an equitable basis.

3.1.5 ECONOMIC EVALUATION

3.1.5.1 TOTAL CAPITAL REQUIREMENT

Total capital investment estimates developed for each of the three process schemes are detailed in Table 3.1.5.1a. These estimates represent current day costs (March 1979) with no escalation included to account for the fact that the facility cannot be built instantaneously. Equipment designs and capital investments are based on four trains for the acid gas removal and adsorber/drier systems and cryogenic unit, and one train for product SNG and recycle compression. Four trains were selected so that equipment sizes would be reasonable, and fabrication and transportation problems would be minimized. Multiple trains also provide an added degree of reliability. No off-sites (cooling towers, electrical switchgear, etc.) were included in the estimates. On this basis, the integrated Benfield/cryogenic process results in the lowest initial capital outlay, $153,730,000. However, the integrated Selexol/cryogenic separation process, at $160,520,000 would be considered equivalent within the absolute accuracy of the estimate. The Rectisol/cryogenic separation process requires a substantially larger initial outlay, 20% greater than the Benfield/cryogenic separation process scheme.

Considering only the capital for the acid gas removal system alone, the Benfield process is, by far, the most economical process. Benfield system capital costs represent only two-thirds of the Selexol system and one-half of the Rectisol system initial investments.

However, in addition to investment for equipment required to treat the sour feed stream upstream of the cryogenic system, the Benfield acid gas removal section includes a sulfur plant capital penalty, as well as investment for an H_2S Trim Unit.

At $40,440,000 the cryogenic separation unit associated with the Selexol acid gas removal unit is the most economical. Feed temperature to the cryogenic unit is the main reason the cryogenic unit investments differ. Feed gas is delivered to the cryogenic separation section at 5°F, 5°F, and 132°F for the Selexol, Rectisol, and Benfield units respectively. As a result, the Benfield adsorber/drier systems require additional molecular sieve for CO_2 and H_2O removal (lower capacity) and a high temperature refrigeration unit to provide refrigeration for precooling the feed gas. In addition, the feed gas enters the cryogenic unit (cold box) at warmer temperatures, requiring additional heat exchange surface in the warm exchanger of the cold box. The warmer cold box feed streams result in warmer SNG and H_2 - CO recycle streams. These returning streams are used to precool the feed to the cold box to reduce the load on the high temperature refrigeration unit. Consequently, suction temperatures to the SNG and recycle compressors are warmer for Benfield resulting in larger power requirements for these compressors. The end result is the cryogenic unit associated with Benfield requires an investment of $47,430,000, compared to $40,440,000 for Selexol and $41,400,000 for Rectisol.

3.1.5.2 OPERATING COSTS

Table 3.1.5.2a summarizes the estimated operating costs for each of the integrated acid gas removal/cryogenic separation processes. Estimates of power, steam, cooling water requirements and solvent losses were provided by Lotepro, Benfield, and Air Products for the Rectisol, Benfield, Selexol acid gas removal units respectively. Utilities for the cryogenic units were determined from process designs developed by Air Products for each of the three acid gas removal processes. R. M. Parsons provided the utility estimates for the sulfur plants which were used to determine the operating cost penalty to be assessed to the Benfield process. A detailed utility summary is provided in Table 3.1.5.2b. No operating costs have been considered for pretreating the feed gas upstream of the acid gas unit and no credit has been given to the fuel stream produced in the cryogenic unit. Valuing the fuel stream at $2.50/MMBTU would reduce operating costs for each process by $12,500,000.

Operating costs, listed in Table 3.1.5.2a, are based on 7884 hours per year of plant operation, a 90% on-stream factor. From a reliability standpoint, Air Products believes that the design, technology and operating experience for the three integrated processes is such that any of the gas clean-up/cryogenic separation processes should be capable of operating at least 95% of the time. However, the nature of the upstream process is such that a 90% on-stream factor was chosen as being representative of a new technology coal facility.

On an operating cost basis, the integrated Selexol/cryogenic separation process is clearly the most economical. Total operating costs are $52,255,000, 16.3% less than the integrated Benfield process and 15.8% less than Rectisol. Considering the current energy situation and the fact that energy costs comprise approximately 75% of the operating costs, energy efficiency becomes extremely important.

3.1.5.3 DISCUSSION

The economic analysis shows that the Benfield process requires a lower initial capital investment than the Selexol process, $153,730,000 versus $160,520,000. However, annual operating costs for Selexol are substantially less than those for Benfield, $52,255,000 versus $62,085,000.

Capital charges for the Selexol, Rectisol, and Benfield integrated processes are $20,109,000, $23,073,000, and $19,272,000, respectively, or approximately 12.5% of total capital. Adding capital charges to operating costs yields total revenues required, which, given annual production, translates into a gas treating cost. Table 3.1.5.3a summarizes the results of the utility financing analysis. Twenty-year average gas costs are $0.856, $1.007, $0.967/MSCF for Selexol, Rectisol, and Benfield respectively. First year treating costs would be $0.987, $1.157, and $1.093/MSCF for the three processes. With utility financing, savings in operating costs for the Selexol/cryogenic separation process more than offset the additional initial capital expenditure required.

Table 3.1.5.3b shows the results of the CCU-A.G.A DCF analysis. The CCU-A.G.A. method, which does not account for escalation, results in significantly higher capital charges than those resulting from utility financing methcds. Capital charges are approximately 25% of total capital investment, twice the charges assessed under utility financing. Based on a 15% DCF return, gas treating costs are approximately 25% higher at $1.096, $1.282 and $1.195/MSCF for the Selexol, Rectisol, and Benfield processes, respectively.

The results of Air Products' DCF method are shown in Table 3.1.5.3c. Under Air Products' analysis, the required treating costs are very close to the utility financing method costs at $0.889/MSCF, $1.042/MSCF, and $0.971/MSCF for the integrated Selexol, Rectisol, and Benfield processes. Treating costs are shown in $/MSCF and $/MMBTU for each analysis method. The heating value for SNG produced by each process is the same and either cost unit yields similar overall results.

Utility financing and DCF analysis both indicate that the Selexol/ cryogenic separation process is the most promising process for the Exxon gasification scheme. Treating costs are minimized by employing this process.

3.1.5.4 SENSITIVITY ANALYSIS

In order to assess the degree of confidence in the selection of the most promising process, the operating and capital cost assumptions which most significantly affect the selection were analyzed. The most promising processes considered in the current study were the integrated Selexol and Benfield processes. The significant variables analyzed were investment, power, and steam costs. A sensitivity analysis was performed on each of these variables. In each case, the required change in the appropriate variable was determined such that treating costs for both process configurations were equivalent, all other things being equal.

Under the utility financing method described previously, Selexol investment must increase $74,680,000 or 48% before Selexol and Benfield treating costs are equal. The increase is only 20% using the CCU-A.G.A. method. Similarly, steam costs must decrease 58% and 52%, and power costs must increase 104% and 92% under utility and private investor financing methods, respectively. The implication of the sensitivity analysis is that the selection of the most promising process, the integrated Selexol/cryogenic system, is not very sensitive to variations in the most crucial assumptions. As a result, confidence in the selection is high.

3.1.6 CONCLUSION AND RECOMMENDATION

Results of the Task I screening study on the integration of commercially proven acid gas removal and cryogenic methane separation indicate that the Selexol/cryogenic separation alternative is the most economically attractive method for the Exxon CCG process to produce SNG from coal. Although the integrated Selexol/cryogenic process requires a slightly larger initial capital investment than the integrated Benfield process, operating cost savings are more than sufficient to overcome this penalty. On a 20-year average utility financing basis, the costs associated with purifying and separating the feed gas stream (treating costs) are $0.856/MSCF for the Selexol/cryogenic separation alternative; 11.5% lower than the costs for Benfield, and 14.9% lower than Rectisol. Evaluating the alternative processes using private investor financing methods produces similar results. Sensitivity analysis of investment, power, and steam costs indicates that the selection is relatively insensitive to the estimated values for these costs. Therefore, on the basis of requiring the lowest treating costs, the Selexol/cryogenic alternative is recommended as the best process to be used for the Exxon gasification scheme.

The Selexol/cryogenic separation process currently proposed represents significant capital and utility savings over the Selexol/cryogenic separation process presented by Exxon in its total facility estimate. A capital cost of $156,300,000 for the Air Products' design is approximately 75% of that proposed by Exxon. In addition, the annual utility requirement, $40,320,000/year, also represents approximately 75% of the corresponding utility cost presented by Exxon. Based on the work that has been completed for the screening study, capital and operating costs can potentially be reduced further through additional optimization steps in both the acid gas and cryogenic sections of the plant. These optimization steps may be considered in Task II.

With respect to the optimization and sensitivity study on the Selexol/cryogenic integration, the following parameters may be considered:

a. The intermediate flash pressure level of the rich Selexol solution - CH_4, CO, and H_2 losses vs gas recompression power.

b. Partial condensation of residual CO_2 in the cold box, to eliminate the use of adsorption beds for the removal of CO_2.

c. Modifications to the design of the CH_4 stripper column in the cold box to eliminate the use of vacuum methane refrigeration, with potential savings in horsepower.

APPENDIX B

FIG. 3.1.1.3a: SELEXOL AND CRYOGENIC INTEGRATION SCHEME

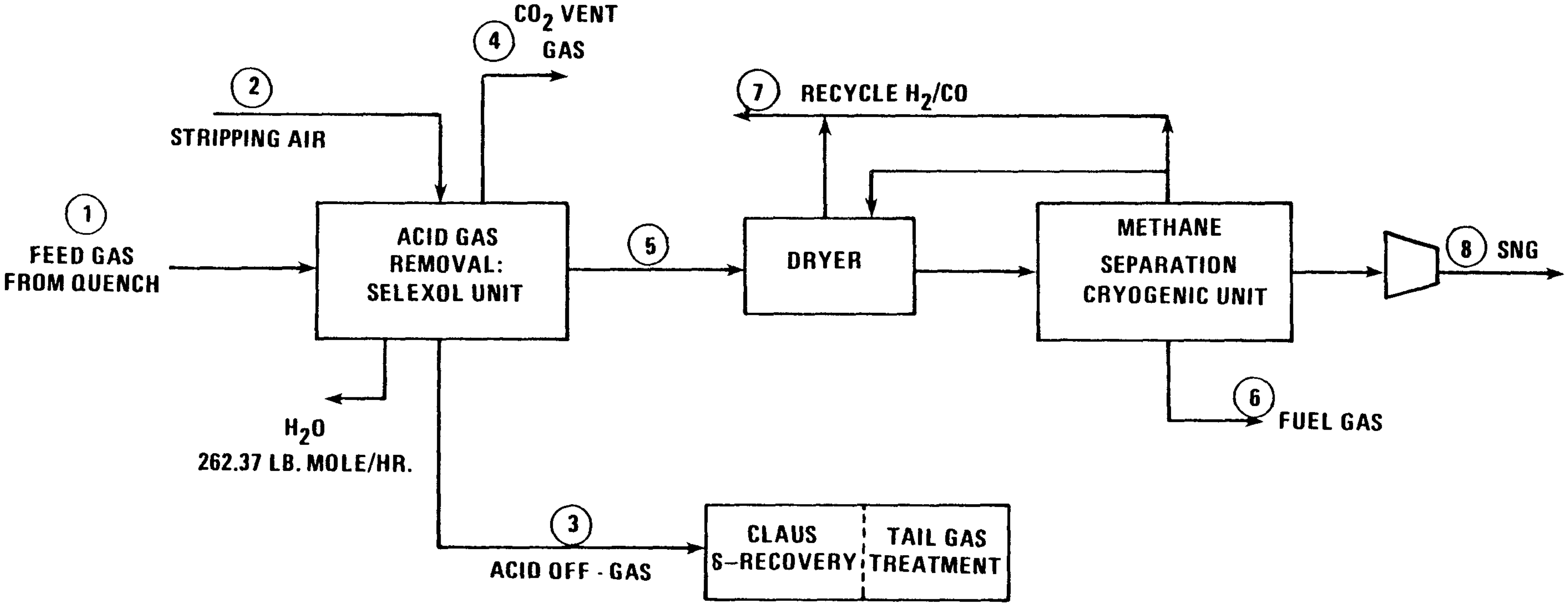

TABLE 3.1.1.3a
MATERIAL BALANCE FOR THE INTEGRATION OF SELEXOL AND CRYOGENIC SEPARATION

STREAM NO.		①	②	③	④	⑤	⑥	⑦	⑧
STREAM NAME		FEED GAS TO SELEXOL	STRIPPING AIR	ACID OFF-GAS TO CLAUS	CO_2 VENT	TREATED GAS	FUEL GAS	RECYCLE H_2/CO	SNG PRODUCT
PRESS., PSIA		450	14.7	32	15.7	431	45	600	1,015
TEMP., °F		120	80	110	4	4.8	105	157	120
LB–MOLE/HR	CO	10,510.0		0.28	5.92	10,503.80	1.61	10,473.96	28.23
	H_2	37,671.0		0.01	0.62	37,670.37		37,670.37	
	CH_4	35,657.0		4.12	69.70	35,583.18	1,611.13	5,743.92	28,228.13
	C_2H_6	5.0		0.03	0.44	4.53	0.25		4.28
	N_2	3,585.0	17,814.7		17,814.43	3,585.27	0.01	3,584.04	0.22
	O_2		4,735.5		4,735.50				
	CO_2	21,633.0	7.45	1,497.13	20,143.32	6.75		6.67	0.08
	COS	2.0		0.48	1.49	≤0.6ppm			≤1.7ppm
	H_2S	878.0		877.86	0.12	≤0.3ppm			≤0.8ppm
	H_2O	416.8		100.92	48.41	5.10		5.10	
	TOTAL	110,357.8	22,557.65	2,480.83	42,819.95	87,359.00	1,613.00	57,484.06	28,260.94
TOTAL, LBS/HR		2,032,572.0	650,909.00	97,731.3	1,539,343.4	1,041,975.2	25,900.3	562,264.3	453,792.7

Table 3.1.1.3b: Utility Summary
Selexol/Cryogenic Integration

1. Power

	KW
Refrigeration	
Selexol	19,860
Cryogenic	0
Compression	
Selexol	17,950
Cryogenic	70,580
Pumping	
Selexol	11,870
Cryogenic	50
Electric Heater (Cryo.)	5,180
Subtotal	125,490

2. Steam

165 psia, sat. steam	127,300 lbs/hr

3. Cooling Water, @ 85°F

	GPM
Selexol	30,800
Cryogenic	13,200
Subtotal	44,000

4. Chemicals

Selexol Solvent	450 lbs/day

FIG. 3.1.2.3a: RECTISOL AND CRYOGENIC INTEGRATION SCHEME

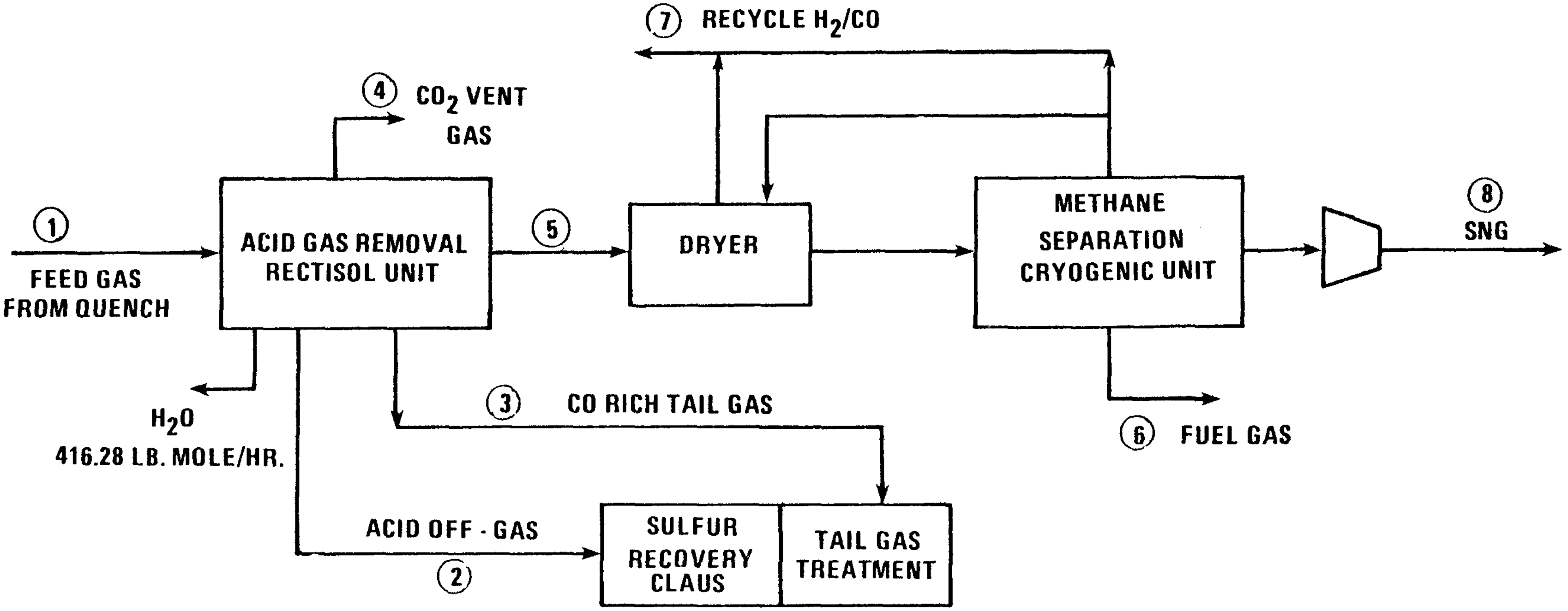

TABLE 3.1.2.3a
MATERIAL BALANCE FOR THE INTEGRATION OF RECTISOL AND CRYOGENIC SEPARATION

STREAM NO.		1	2	3	4	5	6	7	8
STREAM NAME		FEED GAS TO RECTISOL	ACID OFF-GAS TO CLAUS	CO RICH TAIL GAS TO TAIL GAS TREATMENT	CO_2 VENT	TREATED GAS	FUEL GAS	RECYCLE H_2/CO	SNG PRODUCT
PRESS., PSIA		450	26.2	21.8	17.4	432.1	45	600	1,015
TEMP., °F		120	95	89.3	89.3	5.0	105	220	120
LB-MOLE/HR	CO	10,510.0	1.18	53.84	3.21	10,451.77	1.62	10,421.88	28.27
	H_2	37,671.0	1.21	146.58	1.77	37,521.44	0	37,521.44	0
	CH_4	35,657.0	5.02	64.12	10.24	35,577.62	1,611.68	5,726.36	28,239.58
	C_2H_6	5.0	0.33	0.86	1.77	2.04	0.11		1.93
	N_2	3,585.0	5.25	0.95	0.33	3,578.47	0.02	3,578.23	0.22
	CO_2	21,633.0	1,446.16	3,656.68	16,523.95	6.21		6.15	0.06
	COS	2.0	1.97		0.01	<0.3ppm	<0.7ppm		<0.7ppm
	H_2S	878.0	877.00	0.04	0.93	<0.4ppm	<1.2ppm		<1.2ppm
	H_2O	416.8	0			0.52		0.51	
	TOTAL	110,357.8	2,338.12	3,923.07	16,542.22	87,138.11	1,613.43	57,254.57	28,270.06
TOTAL, LBS/HR		2,032,572.0	93,925.0	163,816.6	727,572.2	1,039,758.4	25,905.4	559,944.9	453,906.3

Table 3.1.2.3b: Utility Summary
Rectisol/Cryogenic Integration

1. Power

	KW
Refrigeration	
Rectisol	39,000
Cryogenic	0
Compression	
Rectisol	12,000
Cryogenic	74,800
Pumping	
Rectisol	8,000
Cryogenic	50
Electric Heater (Cryo.)	5,170
Subtotal	139,020

2. Steam

150 psig Sat. Steam	38,000 lbs/hr
65 psig Sat. Steam	260,000 lbs/hr

3. Cooling Water

	GPM
Rectisol	24,800
Cryogenic	22,260
Subtotal	47,060

4. Chemicals

Methanol	23,760 lbs/day

FIG. 3.1.3.4a: BENFIELD AND CRYOGENIC INTEGRATION SCHEME

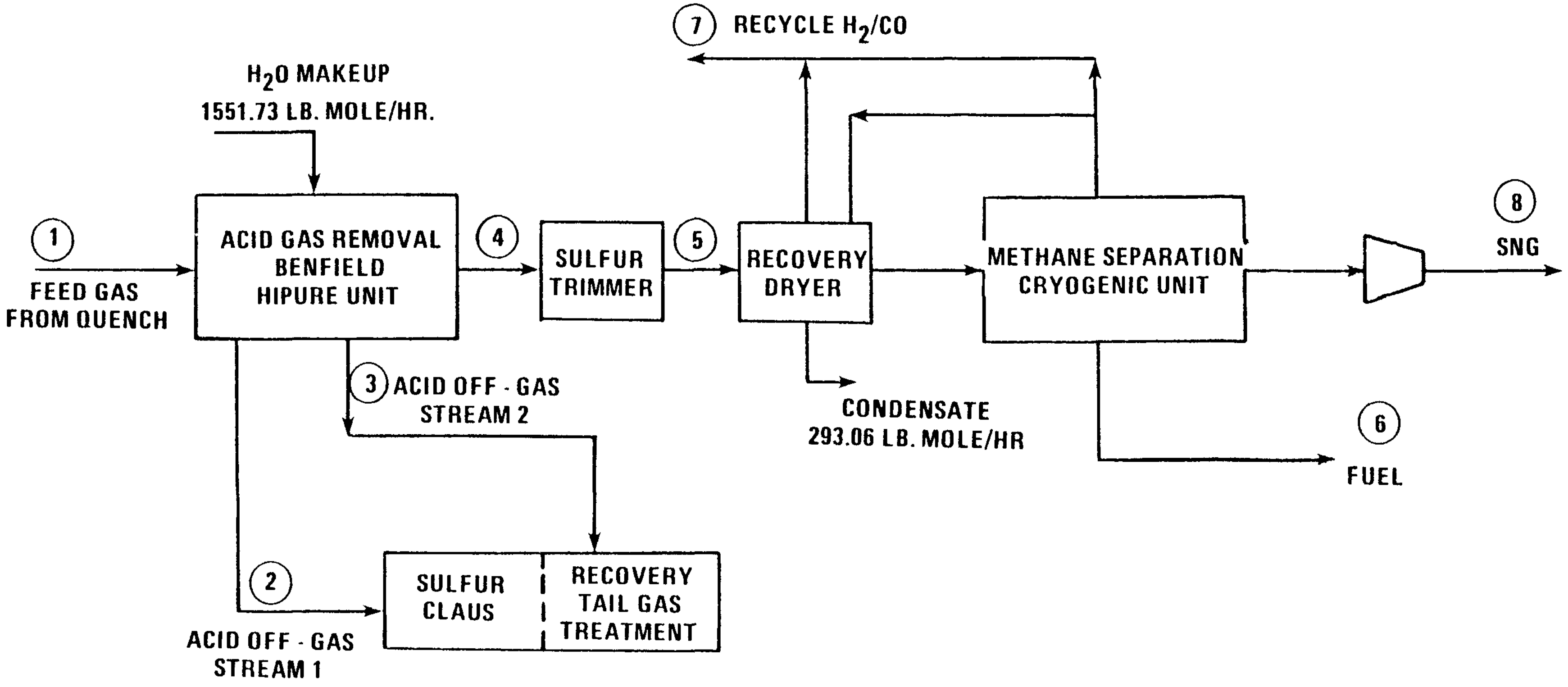

TABLE 3.1.3.4a
MATERIAL BALANCE FOR THE INTEGRATION OF BENFIELD AND CRYOGENIC SEPARATION

STREAM NO.		①	②	③	④	⑤	⑥	⑦	⑧
STREAM NAME		FEED GAS TO BENFIELD	ACID OFF-GAS TO CLAUS	ACID OFF-GAS TO TAIL GAS TREATMENT	TREATED GAS TO SULFUR TRIMMER	PURIFIED GAS TO CRYOGENIC	FUEL GAS	RECYCLE H_2/CO	SNG PRODUCT
PRESS., PSIA		450	24.7	25	444	440	45	600	1,015
TEMP., °F		120	120	120	122	122	105	230	120
LB–MOLE/HR	CO	10,510.0	4.84	22.78	10,482.38	10,482.38	1.61	10,452.56	28.21
	H_2	37,671.0	24.65	106.08	37,540.27	37,540.27		37,540.27	
	CH_4	35,657.0	18.79	88.55	35,549.66	35,549.66	1,611.19	5,729.30	28,209.17
	C_2H_6	5.0		0.02	4.98	4.98	0.26	0.01	4.71
	N_2	3,585.0	0.02	0.23	3,584.75	3,584.75	0.01	3,584.52	0.22
	CO_2	21,633.0	4,910.73	16,714.24	8.03	8.03	0.01	7.94	0.08
	COS	2.0			0.10	0.03	0.005		0.025
	H_2S	878.0	772.26	107.39	0.25	0.05	0.005		0.045
	H_2O	416.8	409.25	1,236.03	323.25	323.10		30.03	0.009
	TOTAL	110,357.8	6,140.54	18,275.32	87,493.67	87,493.25	1,613.09	57,344.63	28,242.47
TOTAL, LBS/HR		2,032,572.0	250,300.2	763,801.5	1,046,376.8	1,046,363.1	25,902.4	561,676.5	453,504.4

Table 3.1.3.4b: Utility Summary
Benfield/Cryogenic Integration

1. Power

	KW
Refrigeration	
Benfield	0
Cryogenic	1,770
Compression	
Benfield	0
Cryogenic	75,090
Pumping	
Benfield	11,050
Cryogenic	50
Electric Heater (Cryo.)	5,550
Subtotal	93,510

2. Steam

165 psia Sat. Steam	314,950 lbs/hr
65 psia Sat. Steam	495,620 lbs/hr

3. Cooling Water, @ 85°F

	GPM
Benfield	60,400
Cryogenic	17,800
	78,200

4. Chemicals

Benfield	$ 398/day
Sulfur trim	$ 40/day

TABLE 3.1.5.1a

TOTAL CAPITAL REQUIREMENT ($M) - (March 1979 $)

	Selexol	Rectisol	Benfield
Acid Gas Removal Section			
Acid Gas Removal System	57,890	77,100	38,900
Sulfur Plant Penalty	-	-	7,890
H_2S Trim Unit	-	-	2,600
Total	57,890	77,100	49,540
Cryogenic Separation Section			
Adsorber/Drier System	2,930	2,930	7,210
Cryogenic Unit	17,410	17,410	19,100
Compression	20,100	21,060	21,120
Total	40,440	41,400	47,430
Sub-total - Installed Equipment	98,330	118,500	96,970
Contractors Fee (3% Installed Equip. plus Engineering	3,100	3,710	3,070
Engineering (APCI estimate)	5,100	5,100	5,200
Sub-total	106,530	127,310	105,240
Contingency (15%)	15,980	19,100	15,790
Total Plant Investment (TPI)	122,510	146,410	121,030
Initial Charge of Chems./Cats.(1)	8,660	690	810
Paid-up Royalties(2)	580	3,700	1,500
Allowance for funds used during construction (S.T.I.)(3)	13,990	16,040	13,350
Start-up Costs (20% of gross operating Costs)	10,500	12,420	12,490
Sub-total - Total Investment	156,320	179,260	149,180
Working Capital(4) (WC)	4,200	4,900	4,550
Total Capital Requirement (TCR)	160,520	184,160	153,730

(1) Selexol - 963,360 gals. fo Selexol @ $8.75/gal + 118,800 lbs. of mole sieve @ $1.90/lb.

Rectisol - 832,700 ga. of Methanol @ $0.55/gal + 118,800 lbs. of mole sieve @ $1.90/lb.

Benfield - $413,700 (activated K_2CO_3) + 156,000 lbs. of mole sieve @ $1,90/lb + $104,000 of iron sponge.

(2) Lotepro, Benfield and Allied (Air Products) have provided estimates of royalty and licensing fees.

(3) STI (Short Term Interest) = 10% (average spending period) (TCR - WC - STI). The average spending period = 0.9825 yrs. for a 3 yr. construction period.

(4) WC = 0.9% (TPI) + 1/24 (annual treating costs).

TABLE 3.1.5.2a

OPERATING COSTS ($M/YR) 90% onstream = 7,884 hrs/yr

	Selexol	Rectisol	Benfield
Capital ($M)			
Investment	156,320	179,260	148,940
Working Capital	4,200	4,900	4,500
Total Capital	160,520	184,160	153,440
Operating Cost ($M/yr)			
Power @ $.035/kwh	35,318	39,137	26,316
Steam - 150 psig @ $3.50/M lbs	3,512	1,049	8,689
50 psig @ $2.80/M lbs	-	5,740	10,941
Cooling Water @ $.10/M gal circ.	2,081	1,963	3,699
Solvent Losses (1)	148	650	137
Adsorbent/Catalyst Replacement	113	113	218
Sulfur Plant Penalty	-	-	718
Hydrocarbon Losses (2)	739	1,332	1,505
Labor - Operating (20 men @ $30 M/yr	600	600	600
Maintenance (60% of Total Maint.) (3)	2,205	2,635	2,179
Supervision (20% of Oper. & Maint. Labor)	561	647	556
Supplies - Operating (30% of Oper. Labor)	180	180	180
Maintenance (40% of Tot. Maint.)	1,470	1,757	1,452
Taxes and Insurance (2.7% of TPI)	3,308	3,953	3,268
Administration & Gen. Overhead (60% of Total Labor)	2,020	2,329	2,001
Total Gross Operating Costs, $M/yr	52,255	62,085	62,459

(1) Solvent Costs - Selexol @ $8.75/gal.
Methanol @ $0.55/gal.
Activated K_2CO_3 @ $17.41/hr - Benfield Estimate

(2) Hydrocarbon Costs - Methane @ $3.50/MSCF
H_2 @ $2.50/MM BTU
CO @ $2.50/MM BTU

(3) Total Maintenance = 3% of Total Plant Investment

TABLE 3.1.5.2b

UTILITY SUMMARY

	Selexol	Rectisol	Benfield
Power, KW			
Acid Gas Section	49,670	59,030	11,050
Cryogenic Section	75,810	80,020	82,450
Miscellaneous (2%)	2,510	2,780	1,870
Total	127,990	141,830	95,370
Steam, lbs/hr			
Acid Gas Section			
150 psig	127,300	38,000	314,900
50 psig	-	260,000	495,600
Cooling Water, GPM			
Acid Gas Section	30,800	24,800	60,400
Cryogenic Section	13,200	16,700	17,800
Total	44,000	41,500	78,200
Solvent Losses, lbs/hr			
Acid Gas Section	18.8	990.0	\$17.41(1)/hr
Adsorbent & Catalyst Replacement, lbs/yr			
Acid Gas Section	-	-	13,850
Cryogenic Section	59,400	59,400	78,000
Methane, Hydrogen and Carbon Monoxide Losses, $\frac{\text{lb/moles}}{\text{hr}}$			
CH_4	70.7	79.4	107.3
H_2	0.6	149.5	130.7
CO	6.3	58.2	27.6

(1) \$17.41/hr estimate provided by Benfield.

TABLE 3.1.5.3a

TREATING COSTS

UTILITY FINANCING METHOD (20 YR. AVERAGE COST)

	Selexol	Rectisol	Benfield
Total Gross Operating Costs	52,255	62,085	62,459
Capital Charges			
Depreciation (20 yr. S.L.)	7,816	8,963	7,459
L.T. Interest (10% of Avg. Net Investment)	6,177	7,090	5,936
Taxes (49.5% - Fed., St., Loc.)	3,027	3,475	2,909
NPAT (15% Return on Equity Capital)	3,089	3,545	2,968
Total Capital Charges, \$M/yr	20,109	23,073	19,272
Total Required Revenues, \$M/yr	72,364	85,158	81,731
Production, MM SCF/yr	84,556	84,556	84,490
Treating Cost, \$/MSCF	0.856	1.007	0.967
Treating Cost, \$/MM Btu	0.848	0.998	0.959

TABLE 3.1.5.3b

TREATING COSTS

DCF - CCU - A.G.A. METHOD

	Selexol	Rectisol	Benfield
Total Gross Operating Costs	52,255	62,085	62,459
Capital Charges	40,390	46,327	38,508
Total Required Revenues, $M/yr	92,645	108,412	100,967
Production, MMSCF/yr	84,556	84,556	84,490
Treating Cost, $/MSCF	1.096	1.282	1.195
Treating Cost, $/MM Btu	1.086	1.271	1.185

TABLE 3.1.5.3c

TREATING COSTS

DCF - AIR PRODUCTS METHOD

	Selexol	Rectisol	Benfield
Total Gross Operating Costs	52,255	62,085	62,459
Capital Charges	22,915	25,991	19,581
Total Required Revenues, $M/yr	75,170	88,076	82,040
Production, MM SCF/yr	84,556	84,556	84,490
Treating Cost, $/MSCF	0.889	1.042	0.971
Treating Cost, $/MM Btu	0.881	1.033	0.962

DELETED

3.2 ROCKWELL HYDROGASIFICATION PROCESS

3.2.1 ACID GAS REMOVAL SYSTEM

For the Rockwell Hydrogasification process, the total acid gas partial pressure and the ratio of H_2S to CO_2 are 7.7 psia and 3.3, respectively. Because of the low combined H_2S/CO_2 ratio, chemical absorption processes were chosen for the screening study. The processes selected for acid gas removal were: MEA, DEA, improved DGA, and Sulfinol.

After a discussion with the Benfield Corporation, the Benfield HiPure process was eliminated from the list of consideration due to the low acid gas content which makes the Benfield process unattractive.

The MEA process was included for screening because it is an economically competitive process even with the low COS (about 10 ppmv, according to the NASA JANAF chemical equilibrium calculation) impurity present in the raw acid gas stream.

The specifications of the total sulfur and CO_2 leaving the acid gas removal system were set at less than 4 and 50 ppmv, respectively. The sulfur and CO_2 were subsequently further conditioned to the acceptable levels for the cryogenic or hypersorption units, and to meet the specifications for the pipeline gas.

All amine process designs were performed in-house, augmented by discussions with the process vendors. For the Sulfinol process, a quotation was obtained from the Shell Oil Co. A more detailed process design, equipment sizing and utility requirements were than prepared in-house according to the design parameters suggested by Shell. All of the processes were designed based upon a two train arrangement. For the high system pressure of the Rockwell case, two paralled trains give reasonable column and equipment sizes.

3.2.1.1 MEA ACID GAS REMOVAL

The MEA system employs aqueous monoethanolamine as the absorbent. The principle of this (and every) amine process rests on the reversible chemical reaction of a weak base with a weak acid gas to form a water soluble salt. The nature of the reversible reactions allows regeneration of the amine solution.

MEA is the strongest base among the different amines and therefore reacts most rapidly with the acid gases. Because of low molecular weight, resulting in higher acid gas pick up (on a weight basis), MEA requires a relatively low solvent rate and pumping power. Other advantages include its stability, low thermal degradation, and ease of reclaiming. On the other hand, the MEA process has some disadvantages. MEA reacts with COS and CS_2 to form nonregenerative compounds. The solvent loss is relatively high due to its high vapor pressure. MEA is not a selective acid gas removal process.

In the MEA system, the raw feed acid gas passes into an absorber and flows upward, countercurrent to the MEA solution which removes the acid gases. In this design the MEA solution strength was set at 20 weight percent; the rich acid gas loading was set at 0.394 moles of acid gas per mole of MEA. The rich amine solution is flashed to a low pressure level, and then is filtered to remove insoluble products. The solvent is then heated against the hot lean solvent to approximately 200°F before charging to the top of a stripper. In the stripper, the acid gases are stripped overhead by using low pressure steam. After cooling and removal of condensate, the acid off-gas is sent to a Claus unit for sulfur recovery. The hot lean solvent from the bottom of the stripper is cooled and pumped back to the absorber.

A slipstream of 1 to 3 percent of the total amine circulation rate is pumped through a reclaimer to restore amine usefulness by means of a semi-continuous batch distillation. Soda ash or caustic soda is added to neutralize the stable salts.

The process design, equipment sizing, and the utility estimate for the MEA system were performed in-house.

3.2.1.2 DEA ACID GAS REMOVAL

Aqueous diethanolamine solution is used as the absorbent in the DEA system. Because of the relatively high molecular weight, high DEA loading is generally used to reduce the solvent circulation rate. In this design, the DEA solution strength was set at 20 weight percent; the rich acid gas loading was set at 0.435 moles of acid gas per mole of DEA. Unlike MEA, DEA is not degraded by COS.

The process configuration and equipment arrangement are identical to that of MEA, with one exception, there is no reclaimer. DEA solution is more difficult to reclaim, and reclaiming is not economically justified.

The process design, equipment sizing, and utility estimate for the DEA system performed in-house.

3.2.1.3 IMPROVED DGA ACID GAS REMOVAL

The DGA system uses diglycolamine as the absorbent to remove acid gases as well as impurities, such as COS and CS_2. Savings in both capital and operating cost can be realized using a high DGA concentration, resulting in a low solvent circulation rate.

DGA has several important physical and chemical characteristics. Aqueous solutions have low freezing points (-40°F) and are thermally stable at temperatures in excess of 400°F. DGA is noncorrosive and has little tendency to foam. A relatively low vapor pressure results in low chemical losses. DGA reacts with COS and CS_2, and the reactions are thermally reversible at elevated temperature and regenerator pressure. However, DGA does have some disadvantages. The solvent is expensive and is a good extraction agent for aromatics, resulting in more expensive Claus sulfur recovery.

In 1965, the first commercial DGA plant was put on-stream by El Paso Natural Gas Company at Monument, New Mexico. Since then, over 30 plants using DGA in conventional amine plant configurations have been put in operation. Recently, an improved DGA process has been developed by Fluor Engineers and Constructors, Inc. Fluor has done extensive testing in Saudi Arabia to expand its DGA data base and to assure the reliability of its innovation.

The process configuration for improved DGA is similar to that of MEA. All of the process design, equipment sizing, and utilities estimate were performed in-house, augmented by discussions with Fluor and with the Jefferson Chemicals Co. In this design, the DGA solution strength was fixed at 65 weight percent; the rich acid gas loading was set at 0.450 moles of acid gas per mole of DGA.

3.2.1.4 SULFINOL ACID GAS REMOVAL

The Sulfinol system employs sulfolane (tetrahydrothiophene dioxide) and DIPA (diisopropanol amine) as solvents. Sulfinol combines both physical and chemical absorption capabilities in the absorption cycle. Generally, Sulfinol appears to have advantages, with regard to solvent circulation and utility requirements, with a feed H_2S/CO_2 ratio greater than 1 and/or with a feed acid gas partial pressure greater than 110 psia. Sulfinol also has another attractive feature: its capability to remove COS, CS_2, and mercaptans. Since 1963, Sulfinol has gained wide acceptance in the gas industry.

The Sulfinol process uses a conventional solvent absorption and regeneration cycle, as is usually employed in amine systems. The acid gases are removed from the feed gas by countercurrent contact with a lean solvent under pressure in the absorber. The treated gas is further washed with water to remove solvent carried over. The rich solvent is let down in pressure through a power recovery turbine. Then, the solvent is flashed before entering a heat exchanger to be heated by hot lean solvent from the bottom of the regenerator. Flash gas is compressed back to the absorber. The hot, rich solvent is regenerated to lean solvent by steam stripping. The hot lean solvent is cooled by air or water before returning to the absorber.

A quotation for the Sulfinol process was received from the Shell Oil Co. Shell furnished the overall detailed heat and material balance, major equipment sizes, major equipment sizes, the materials of construction, and the design parameters. Air Products consulted Shell to confirm and extend the design, equipment sizing, and utility estimates necessary for screening.

3.2.2 HYPERSORPTION FOR METHANE SEPARATION

The acid gas systems studied give essentially the same feed to the methane seperation unit. Therefore the description of the methane separation system is common to all of the acid gas system combinations.

3.2.2.1 HYPERSORPTION CYCLE DESCRIPTION

Process gas from the acid gas system is precooled to reduce the moisture content of the feed. This is required to prevent a buildup of water in the adsorption beds.

Feed gas enters the adsorption tower near the middle of the tower. As the gas contacts the moving carbon bed, the methane is preferentially adsorbed. Methane is then desorbed with a hot desorption gas (chiefly methane at about 600°F). The remaining portion of the feed gas not adsorbed by the carbon bed is withdrawn from the tower and sent to the CO treating section before being returned to the gasifier.

Hot desorption gas enters near the bottom of the tower and heats the moving carbon bed to desorb methane, while the vapors are being cooled. A portion of the vapor is withdrawn from the tower as an impure SNG stream to be further conditioned for the removal of sulfur contaminants. The remaining cool gas (desorption gas) is withdrawn from the tower, compressed, heated and re-enters the bottom of the tower. The desorption gas is preheated against a hydrogen rich recycle stream and the lift-purge gas stream. Final heating of the desorption gas is accomplished in a direct fired heater.

Hot carbon leaves the bottom of the tower and is combined with a hot lift gas (chiefly H_2, N_2, CO at 600°F). The carbon is lifted to a separator where it is separated from the lift gas before being fed back to the tower.

The hot lift gas combines with a purge stream from the top of the tower. This combined stream is used to preheat the desorption gas and the lift gas. The stream is then compressed and the purge portion of the stream is recycled to the tower feed stream. This recycle enables a high recovery of separation products. The remainder of the stream (lift gas) is preheated against itself and heated to 600°F against the methanator effluent, before combining with the hot carbon from the bottom of the tower.

Hot carbon entering the top of the tower requires cooling in order to facilitate the adsorption of methane. This is accomplished with a cool recycle stream (chiefly H_2, N_2, and CO). The stream enters the tower above the recycle H_2 offtake. This gas stream cools the moving carbon, while being warmed (550°F). The hot recycle stream is used to preheat the desorption gas before being compressed and returned to the tower.

The recycle H_2 which leaves the tower contains practicially all of the carbon monoxide contained in the feed gas. In order to meet the CO specifications required by Rockwell, the gas is compressed and methanated before being returned to the gasifier at 1615 psia.

SNG leaving the tower contains sulfur contaminants, because the carbon has an affinity for the sulfur compounds. The sulfur is removed by adsorption using activated charcoal before delivery to the pipeline.

3.2.2.2 DISCUSSION

The hypersorption system has the capability of operating at system pressure, thereby requiring only single stage machines. This results in lower power costs than the cryogenic system.

The main drawbacks to the system (6) are the size of the facility required and the uncertainty of the adsorber operation. The feed to the commercial SNG facility is about 50 times larger than the largest unit built to date. In addition, no new units have been built in the past twenty years, and none were ever built for high pressure operation. Furthermore, the hypersorption cycle requires roughly one half billion Btu/hr of fuel for the heater fuel requirement. This requirement is substantial, and is reflected in the reduced SNG production rate (i.e., a portion of the SNG is used for fuel).

The process design, equipment sizing, and the utilities estimate for the hypersorption system were performed in-house, based principally on the design information published by the Dravo Corporation (6). The uncertainty of this Air Products' estimate is inherently greater than the uncertainty in the cryogenic system estimate; the lack of key adsorption capacity data, detail on equipment internals, scale-up, and operating experience results in an estimate of reduced quality.

A pilot unit would have to be tested to determine the ruggedness and lift of the carbon bed, along with its adsorbent capacities. The replacement of the large amounts of carbon could lead to a high operating cost. Also, in using large diameter beds, engineering and testing would be required to insure proper distribution of the carbon and effective gas contact with minimal by-pass.

3.2.3 CRYOGENIC METHANE SEPARATION

3.2.3.1 CRYOGENIC SEPARATION SCHEME A

3.2.3.1.1 PROCESS CYCLE DESCRIPTION

This cryogenic cycle (labeled scheme A) was developed to fit the original overall flow scheme proposed by Rockwell (Fig. 2.1.2), with methanation of the CO in the SNG product stream. The cycle utilizes a methane wash stream to strip the CO from the recycle H_2 stream. The CO in the SNG stream is then methanated to reduce the concentration below the requirement of less than 0.5% CO in the SNG.

The feed from the acid gas system is precooled to 55°F against the cold box SNG effluent and a freon refrigeration system. Moisture and CO_2 are removed in a molecular sieve adsorption unit before the feed gas enters the cold box.

The cold box consists of four main units, a wound coiled heat exchanger, a CO stripper, a CO scrubber, and a closed loop N_2 refrigeration system. The system is not self-sufficient in refrigeration, therefore additional N_2 refrigeration is used.

Feed to the cold box is precooled against the SNG and H_2 recycle streams. The final stage of cooling and condensation is supplied as reboiler duty for the CO stripper. The vapor and liquid phases are then separated. The liquid (a major portion of the SNG) is flashed to approximately 1065 psia, warmed in the wound coil exchanger and exits the cold box. The vapor (which contains nearly all of the CO is flashed and fed to a CO scrubber where the CO is reduced to 10 ppm by washing with a pure CH_4 stream. The column overheads (the H_2 recycle stream) then exits the cold box via the wound coil exchanger.

Column bottoms, which contain the absorbed CO and absorbed methane, are then fed to a stripping column where CO and absorbed CH_4 are removed from the CH_4 wash stream. The CH_4 wash stream is recycled to the CO scrubber. The small CO/CH_4 stream refrigeration is utilized in the N_2 refrigeration system.

The hydrogen stream is compressed to 1615 psia before being returned to the gasifier.

The SNG stream from the cold box then undergoes impurity removal. Sulfur compounds are removed in an activated carbon system, CO is methanated, and moisture is removed in an alumina dryer system.

3.2.3.1.2 DISCUSSION

The methanation of the SNG as originally proposed by Rockwell contains numerous drawbacks. The design requires an unusually large number of equipment items and excessive power requirements. The cycle to separate the CO with the CH_4 was determined to be too complicated and expensive to pursue.

3.2.3.2 CRYOGENIC SEPARATION SCHEME B

3.2.3.2.1 PROCESS CYCLE DESCRIPTION

As an alternative to the originally proposed scheme, methanation of the cold box feed was considered (Fig. 3.2.4.5). The upstream elimination of CO results in a somewhat conventional cycle for the cryogenic separation of H_2 from CH_4.

This arrangement (labeled scheme B) utilizes three basic steps. Feed gas impurity removal, cryogenic separation, and compression.

Feed gas impurity removal consists of an activated carbon system for the removal of sulfur compounds, a methanator which reduces the CO and CO_2 contents, and an alumina dryer system for the removal of water.

The cold box contains two major equipment items, a wound coil heat exchanger and a separator. The feed gas is cooled and partially condensed in the wound coil heat exchanger. Vapor liquid phases are then separated. The liquid portion is the SNG product and the vapor is the H_2 to be recycled back to gasifier. Both streams are reduced in pressure to utilize optimum exchanger-power requirements, and supply refrigeration to cool the feed.

The cold box effluents are compressed and delivered to the gasifier and pipeline respectively.

3.2.3.2.2 DISCUSSION

Feed methanation provides several benefits for the cryogenic system. Along with a reduction in the CO content, CO_2 is reduced sufficiently to eliminate the use of molecular sieve type adsorbents. In addition, the separation required is essentially a two component (H_2/CH_4) separation versus a three component ($H_2/CO/CH_4$) separation. The net result is a lower power requirement.

Although the power required is more than the hypersorption cycle, the simplicity of the cycle and the minimal equipment items (all commercially proven) make it available for use with minimal testing and without pilot type operations.

3.2.3.3 COMPARISON OF CRYOGENIC SCHEME A AND SCHEME B

In addition to the complicated process cycle, cryogenic scheme A requires much higher utility requirements. Table 3.2.3.3 shows the utility comparison of cryogenic schemes A and scheme B. From these view points, the economic analysis of scheme A was not pursued.

3.2.4 INTEGRATION SCHEME OF ACID GAS REMOVAL AND METHANE SEPARATION

3.2.4.1 MEA AND HYPERSORPTION SCHEME

The process block diagram for the combination of MEA acid gas removal and hypersorption is shown in Figure 3.2.4.1. Table 3.2.4.1a and Table 3.2.4.1.b show the mass balance and the utility summary, respectively, for this scheme.

3.2.4.2 DEA AND HYPERSORPTION SCHEME

The process block diagram for the combination of DEA acid gas removal and hypersorption is shown in Figure 2.3.4.2. The mass balance and the utility summary are shown in Table 3.2.4.2a and Table 3.2.4.2b. respectively.

3.2.4.3 DGA AND HYPERSORPTION SCHEME

The process block diagram for the combination of improved DGA acid gas removal and hypersorption is shown in Figure 3.2.4.3. The mass balance and the utility summary are shown in Table 3.2.4.3a and Table 3.2.4.3b, respectively.

3.2.4.4 SULFINOL AND HYPERSORPTION SCHEME

Figure 3.2.4.4 shows the process block diagram for the combination of Sulfinol acid gas removal and hypersorption. The mass balance and the utility summary are shown in Table 3.2.4.4a and Table 3.2.4.4b, respectively.

3.2.4.5 MEA AND CRYOGENIC SEPARATION SCHEME B

Figure 3.2.4.5 shows the process block diagram for the combination of MEA acid gas removal and cryogenic separation (Scheme B). The mass balance and the utility summary are shown in Table 3.2.4.5a and Table 3.2.4.5b, respectively.

3.2.4.6 DEA AND CRYOGENIC SEPARATION SCHEME B

Figure 3.2.4.5 shows the process block diagram for the combination of DEA acid gas removal and cryogenic separation (Scheme B). The mass balance and the utility summary are shown in Table 3.2.4.5a and Table 3.2.4.6, respectively.

3.2.4.7 DGA AND CRYOGENIC SEPARATION SCHEME B

Figure 3.2.4.5 shows the process block diagram for the combination of DGA acid gas removal and cryogenic separation (Scheme B). The mass balance and the utility summary are shown in Table 3.2.4.5a and Table 3.2.4.7, respectively.

3.2.4.8 SULFINOL AND CRYOGENIC SEPARATION SCHEME B

The process block diagram for the combination of Sulfinol acid gas removal and cryogenic separation (Scheme B) is shown in Figure 3.2.4.8. The mass balance and the utility summary are shown in Table 3.2.4.8a and Table 3.2.4.8.b, respectively.

3.2.5 SULFUR RECOVERY

The acid off gas stream is very concentrated in H_2S (about 69%) and is suitable for conventional Claus sulfur recovery. Since sulfur recovery is outside the scope and independent of the work for the present contract, no cost is included in the overall economic analysis.

3.2.6 ECONOMIC EVALUATION

3.2.6.1 TOTAL CAPITAL REQUIREMENT

Total capital investment estimates developed for each of the four acid gas schemes integrated with both cryogenic separation and moving bed adsorption (hypersorption) units are detailed in Table 3.2.6.1. These estimates represent current day costs (March 1979) with no escalation. Equipment designs and capital investments are based on two trains for the acid gas removal, adsorber/drier, cryogenic and hypersorption units, and one train for product SNG and recycle compression. No offsites (cooling towers, electrical switchgear, etc.) were included in the estimates. On this basis, the integrated DGA/hypersorption process results in the lowest initial capital outlay, $93,035,000. The remaining seven process schemes would be considered equivalent within the accuracy of the estimate with the integrated Sulfinol/cryogenic process requiring the highest capital outlay, $103,211,000.

Considering only the capital requirement for the acid gas removal system, DGA is the most economical process at $6,008,000; DGA represents 61%, 71% and 84% of the capital costs for the Sulfinol, DEA, and MEA processes respectively.

All four acid gas schemes require methanation and sulfur trim units to meet SNG product CO and H_2S (total sulfur) specifications. The methanation unit capital is essentially the same for the integrated cryogenic and hypersorption schemes. However, the activated carbon sulfur trim unit capital for the four integrated/cryogenic schemes is significantly greater than that required for the four integrated/hypersorption schemes. This difference is a direct function of the pressure conditions for each sulfur trim unit. The lower H_2S partial pressure in the cryogenic process sulfur trim unit plus approximately four times greater flow results in larger carbon requirements.

The total installed equipment cost for the cryogenic separation unit, $33,350,000, is equivalent for all four integrated schemes and is slightly greater than the hypersorption unit capital of $32,000,000, for the four integrated schemes.

3.2.6.2 OPERATING COSTS

Table 3.2.6.2a summarizes the estimated operating costs for each of the eight integrated process schemes. All costs indicated were based on Air Products' estimates. A detailed utility summary is provided in Table 3.2.6.2b. As with the Exxon study, no operating costs have been considered for pretreating the feed gas upstream of the acid gas unit.

Operating costs listed in Table 3.2.6.2a are based on 7884 hours per year of plant operation, a 90% on-stream factor. As with the Exxon case, the 90% factor was chosen based on the coal handling and gasification processes.

On an operating cost basis, the combined DGA/hypersorption process is the most economical with gross operating costs of $43,942,000. The combined DGA/cryogenic combined process is the most economical of the integrated cryogenic schemes with costs at $44,321,000. All schemes were reasonably equivalent in total operating costs with the integrated Sulfinol/cryogenic process the least economical at $46,550,000. Energy costs, power and fuel, represent 60% of the total operating costs, indicating that energy costs are very significant for all eight process schemes.

3.2.6.3 DISCUSSION

In order to properly evaluate the treating costs of the processes an equivalent cost basis had to be established. The basis chosen here was $/MMBTU which took into account,

(a) different SNG product flow rates for the cryogenic unit and hypersorption unit, 85,547 MMSCF versus 80,522 MMSCF per year,

(b) different heating values (HHV) for the cryogenic and hypersorption-produced SNG, 971.23 BTU/SCF versus 1007.46 BTU/SCF.

Based simply on lowest capital and operating costs, the combined DGA/hypersorption process is the most economical, $93,035,000 and $43,942,000.

However, an evaluation based on the adjusted SNG production, considering treating costs in $/MMBTU, shows the integrated DGA/cryogenic process is more economical.

Under the utility financing method, capital charges at about 12.5% of the total capital are added to total operating costs to determine the total required revenues. Given annual production, total required revenues translate into a gas treating cost. Table 3.2.6.3a summarizes the results of the utility financing analysis. Twenty-year average gas costs are $0.679/MMBTU and $0.686/MMBTU for integrated DGA/cryogenic and DGA/hypersorption processes, respectively. First year treating costs would be $0.758/MMBTU and $0.764/MMBTU, respectively.

Table 3.2.6.3b summarizes the results of the CCU-A.G.A. DCF analysis. As previously noted, this method results in capital charges of approximately 25% of total capital. Based on a 15% DCF return, gas treating costs are $0.823/MMBTU and $0.828/MMBTU for integrated DGA/cryogenic and DGA/hypersorption processes, respectively.

The results of Air Products DCF method are shown in Table 3.6.2.3c with treating costs comparing closely to the utility financing method costs.

3.2.6.4 SENSITIVITY ANALYSIS

All eight process schemes are reasonably close based on treating costs. A sensitivity analysis considering the two most promising processes, both integrated DGA schemes, was performed. This analysis considered power, fuel, and to a lesser extent, capital variations.

Under the utility financing method, only a 3.0% increase in power costs would make both integrated DGA process treating costs equal. The increase is approximately 2.5% using the CCU-A.G.A. method. Similarly, fuel costs must decrease 3.0% and 2.0% under utility and private investor financing methods, respectively. As a result, the DGA/cryogenic and DGA/hypersorption alternatives are considered equivalent from an economic standpoint.

The hypersorption process requires substantial quantities of fuel for adsorber reactivation, where as the cryogenic unit requires no fuel (but requires power). The fuel costs have a significant impact on the annual operating costs for the integrated hypersorption schemes. Consequently, the selection of the most promising process scheme is very sensitive to the value assigned for fuel as evidenced by the above analysis.

The following considerations were used as a basis for assigning a value of $2.50/MMBTU for fuel:

(a) $2.50/MMBTU reflects current day alternate fuel costs.

(b) This figure approximates the cost of low BTU fuel which could be obtained from the upstream char gasification.

(c) The figure is representative of the incremental cost for capital, coal, and utilities required to produce additional SNG for internal plant consumption as fuel.

If the fuel is considered at $2.50/MMBTU, treating costs for the integrated DGA/hypersorption process are $0.583, $0.724 and $0.573/MMBTU under CCU-AGA utility and private investor financing methods and Air Products DCF method, respectively. These treating costs represent 12-14% savings over the best cryogenic alternate, the integrated DGA/cryogenic scheme. Based on economic considerations alone, the integrated DGA/hypersorption process would then be the most promising scheme.

However, because the hypersorption process requires substantial quantities of fuel (approximately 5% of the total SNG production), product SNG was considered the most appropriate source of fuel. A fuel value of $4.50/MMBTU was assumed because this value approximates the cost of product SNG produced by the Rockwell Hydro gasification process. $4.50/MMBTU represents the opportunity cost of the fuel, that is, the cost associated with the opportunity foregone by not valuing the SNG fuel at its best alternate use as product. This analysis assumed that a market is available for SNG at approximately $4.50/MMBTU which would have to be the case before a coal facility of this type would be constructed.

It should be further stated, however, that even with a favorable fuel value, the hypersorption technology is judged to represent a significant technical risk in comparison to cryogenic separation technology. It is felt the potential economic benefits of hypersorption would have to be quite substantial in order to justify such technical exposure.

3.2.7 CONCLUSION AND RECOMMENDATION

The economics of all of the integrated processes are greatly affected by the values assigned to the utilities.

Though the overall economic analysis did not indicate any process to be clearly the most promising, the integrated DGA/cryogenic process was selected as the most promising process. The following considerations formed the basis of this selection:

(a) Cryogenic separation processes similar to that proposed for the Rockwell gasifier scheme have a proven track record, whereas the hypersorption process has not been proven commercially. Significant R&D and demonstration effort would be required to substantiate the hypersorption technology.

(b) The actual mechanical operation of the moving carbon bed would appear to be far more complex and problematic than cryogenic separation.

(c) The integrated DGA/cryogenic process is the most economical combination on a gas treating cost basis.

APPENDIX C

APPENDIX C (cont'd)

Table 3.2.3.3: Utility Comparison of Cryogenic Schemes A and B

		Scheme A - Methanator on SNG	Scheme B - Methanator on Feed
1.	Power, KW	205,390	97,820
2.	Steam, MM lbs/day 250 psia, 750°F	2.65	2.45
3.	Cooling Water, GPM	83,010	53,820
4.	Chemicals:		
	Active Carbon Replacement for Sulfur Trim, lbs/day	570	2,240
	Methanation Catalyst Replacement, ft^3/day	4	25
	Alumina Replacement for Dryer, lbs/day	140	215
	Mole Sieve Replacement, lbs/day	250	-

FIGURE 3.2.4.1
PROCESS BLOCK DIAGRAM FOR MEA ACID GAS REMOVAL AND HYPERSORPTION SNG SEPARATION

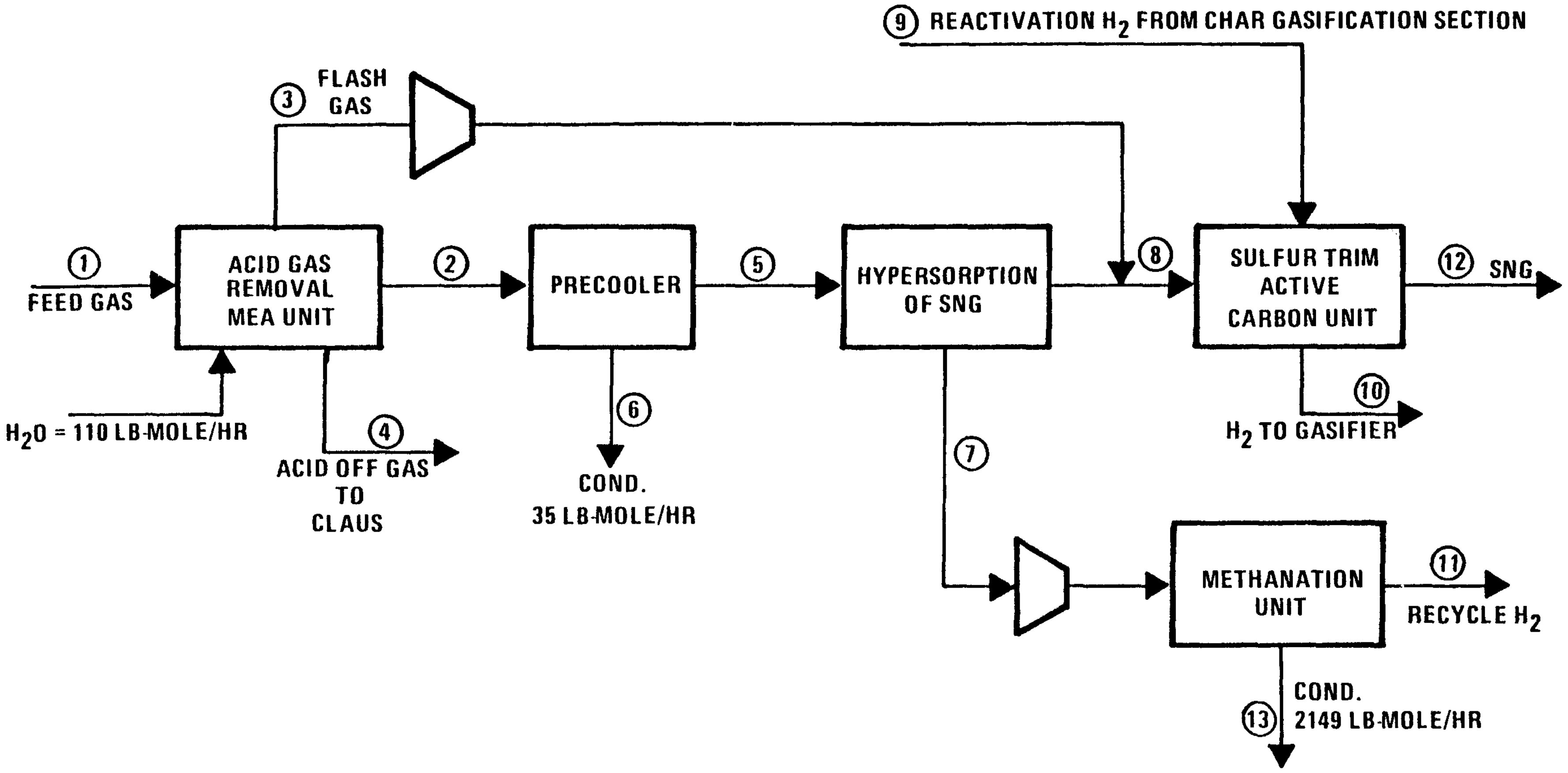

TABLE 3.2.4.1a
MATERIAL BALANCE OF MEA–HYPERSORPTION INTEGRATION

STREAM NO.		①	②	③	④	⑦	⑧	⑨ *	⑩ *	⑪	⑫
STREAM NAME		RAW ACID GAS FEED	TREATED GAS	FLASH GAS	ACID OFF-GAS TO CLAUS	HYPER-SORPTION OFF-GAS	SULFUR TRIM INLET	REACTIVATION H_2 FROM CHAR GASIFICATION	REACTIVATED H_2 TO GASIFIER	RECYCLE H_2	PRODUCT SNG
PRESS., PSIA TEMP., °F		1420 110	1410 95	200 120	25 120	1395 100	1395 135	1625 100	1615 105	1615 105	1015 105
LB–MOLE/HR	CH_4	26,872.00	26,782.00	78.00	12.00		26,860.00			2,200.68	26,860.00
	H_2	63,493.00	63,421.00	62.00	10.00	63,421.00	62.00	10,247.00	10,247.00	56,821.57	62.00
	CO	2,230.00	2,227.00	3.00		2,204.00	26.00			0.71	26.00
	N_2	281.00	281.00			281.00				281.00	
	CO_2	117.00	4.64		112.60		4.64			2.61	4.64
	H_2S	392.00	0.37		391.87		0.37				
	COS	1.00	0.46		0.30		0.46				⩽(4ppm)
	CS_2	(0.13ppm)	(0.13ppm)				0.01				
	C_2H_2	(0.88ppm)	(0.88ppm)				0.08				0.08
	NH_3	3.00			3.00						
	HCN	2.00			2.00						
	H_2O	4.00	76.00		38.00	41.00				49.07	
	TOTAL	93,395.00	92,792.47	143.00	569.77	65,947.00	26,953.56	10,247.00	10,247.00	59,355.64	26,952.72
TOTAL, LBS/HR		648,189.0	629,384.0	1,460.4	19,331.0	198,201	432,016	20,658	20,658	158,749	431,975

NOTE = * ON STREAM EVERY TWO HOURS FOR ONE-HOUR PERIOD

Table 3.2.4.1.b: Utility Summary for MEA - Hypersorption Integration

1. Power	
	KW
MEA	1,640
Hypersorption	32,820
Misc.	690
Subtotal	35,150
2. Steam	
MEA	lbs/hr
60 psig steam	59,890
100 psig steam	15,790
Hypersorption	
250 psia, 750°F steam	176,670 (for 12 hrs/day)
3. Fuel	
Hypersorption	
Heater	517.1 MM Btu/hr
4. Cooling Water	GPM
MEA	2,610
Hypersorption	49,110
Subtotal	51,720
5. Chemicals	
MEA	
MEA	2,590 lbs/day
Active Carbon	50 lbs/day
Soda Ash	2,500 lbs/day
Hypersorption	
Active Carbon Loss	1,900 lbs/day
Active Carbon Replacement for Moving Bed	2,750 lbs/day
Active Carbon Replacement for Sulfur Removal	80 lbs/day
Methanator Catalyst Replacement	10 ft^3/day

FIGURE 3.2.4.2
PROCESS BLOCK DIAGRAM FOR DEA ACID GAS REMOVAL AND HYPERSORPTION SNG SEPARATION

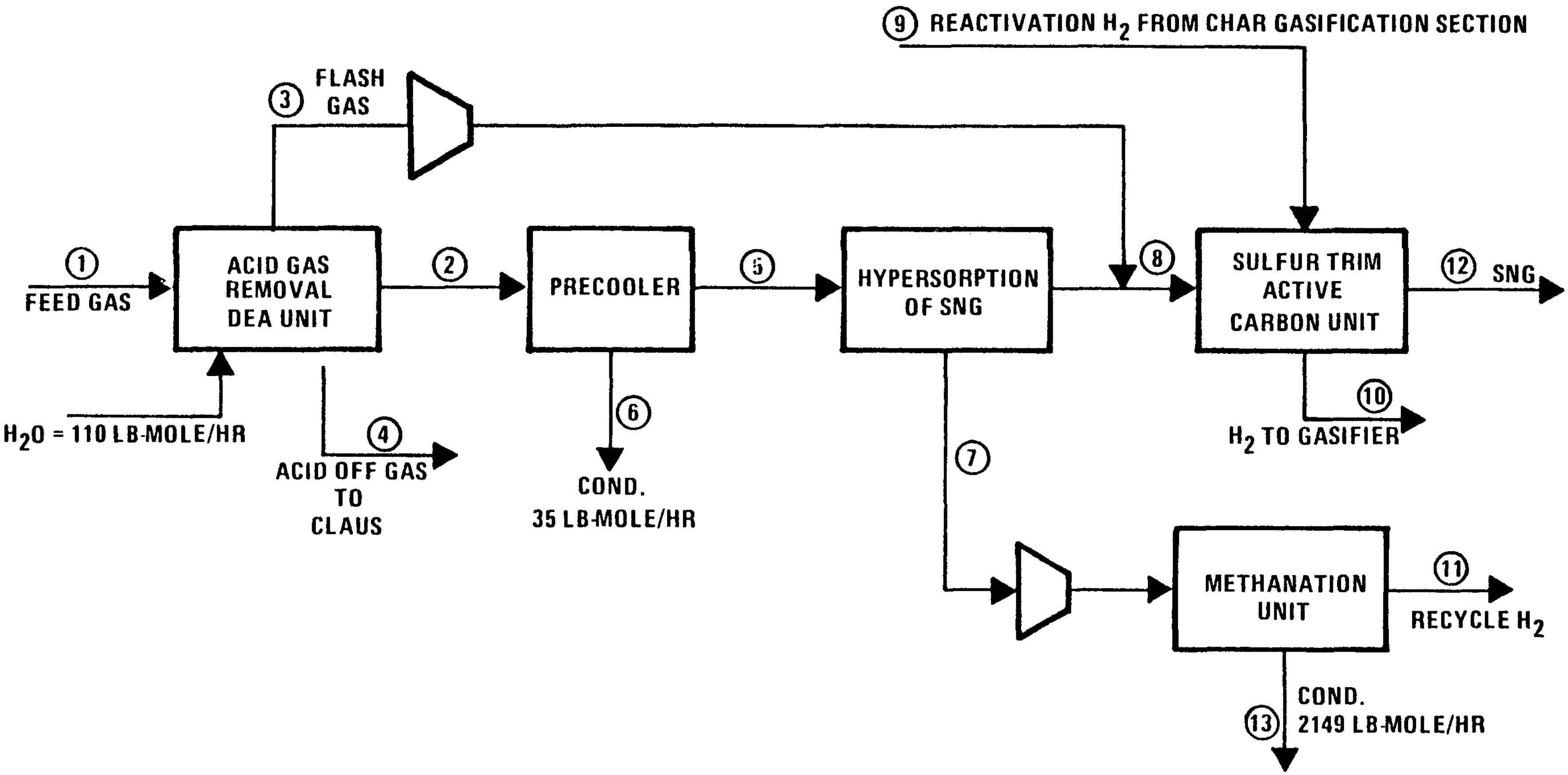

TABLE 3.2.4.2a
MATERIAL BALANCE OF DEA–HYPERSORPTION INTEGRATION

	STREAM NO.	①	②	③	④	⑦	⑧	⑨ *	⑩ *	⑪	⑫
	STREAM NAME	RAW ACID GAS FEED	TREATED GAS	FLASH GAS	ACID OFF-GAS TO CLAUS	HYPER-SORPTION OFF-GAS	SULFUR TRIM INLET	REACTIVATION H_2 FROM CHAR GASIFICATION	REACTIVATED H_2 TO GASIFIER	RECYCLE H_2	PRODUCT SNG
	PRESS., PSIA	1420	1410	200	25	1395	1395	1625	1615	1615	1015
	TEMP., °F	110	95	120	120	100	135	100	105	105	105
LB–MOLE/HR	CH_4	26,872.00	26,782.00	78.00	12.00		26,860.00			2,200.68	26,860.00
	H_2	63,493.00	63,421.00	62.00	10.00	63,421.00	62.00	10,247.00	10,247.00	56,821.57	62.00
	CO	2,230.00	2,227.00	3.00		2,204.00	26.00			0.71	26.00
	N_2	281.00	281.00			281.00				281.00	
	CO_2	117.00	4.64		112.60		4.64			2.61	4.64
	H_2S	392.00	0.37		391.87		0.37				(≤4 ppmv)
	COS	1.00	1.00				1.00				
	CS_2	(0.13 ppm)	(0.13 ppm)				0.01				
	C_2H_2	(0.88 ppm)	(0.88 ppm)				0.08				0.08
	NH_3	3.00			3.00						
	HCN	2.00			2.00						
	H_2O	4.00	76.00		38.00	41.00				49.07	
	TOTAL	93,395.00	92,717.01	143.00	569.77	65,947.00	26,954.10	10,247.00	10,247.00	59,355.64	26,952.72
	TOTAL, LBS/HR	648,189.0	629,416.5	1,460.4	19,331.0	198,201.0	432,048.	20,658.0	20,658.0	158,749.	431,975.

NOTE = * *ON STREAM EVERY TWO HOURS FOR ONE HOUR PERIOD.*

Table 3.2.4.2.b: Utility Summary for DEA - Hypersorption Integration

1. Power	
	KW
DEA	2,220
Hypersorption	32,820
Misc.	700
Subtotal	35,740
2. Steam	
DEA	lbs/hr
60 psig steam	78,360
Hypersorption	
250 psia, 750°F steam	186,670 (for 12 hrs/day)
3. Fuel	
Hypersorption	
Heater	517.1 MM Btu/hr
4. Cooling Water	GPM
DEA	2,610
Hypersorption	49,110
Subtotal	51,720
5. Chemicals	
DEA	
DEA	2,200 lbs/day
Active Carbon	50 lbs/day
Soda Ash	2,500 lbs/day
Hypersorption	
Active Carbon	1,900 lbs/day
Active Carbon Replacement for Moving Bed	2,750 lbs/day
Active Carbon Replacement for Sulfur Removal	90 lbs/day
Methanator Catalyst Replacement	10 ft^3/day

FIGURE 3.2.4.3
PROCESS BLOCK DIAGRAM FOR DGA ACID GAS REMOVAL AND HYPERSORPTION SNG SEPARATION

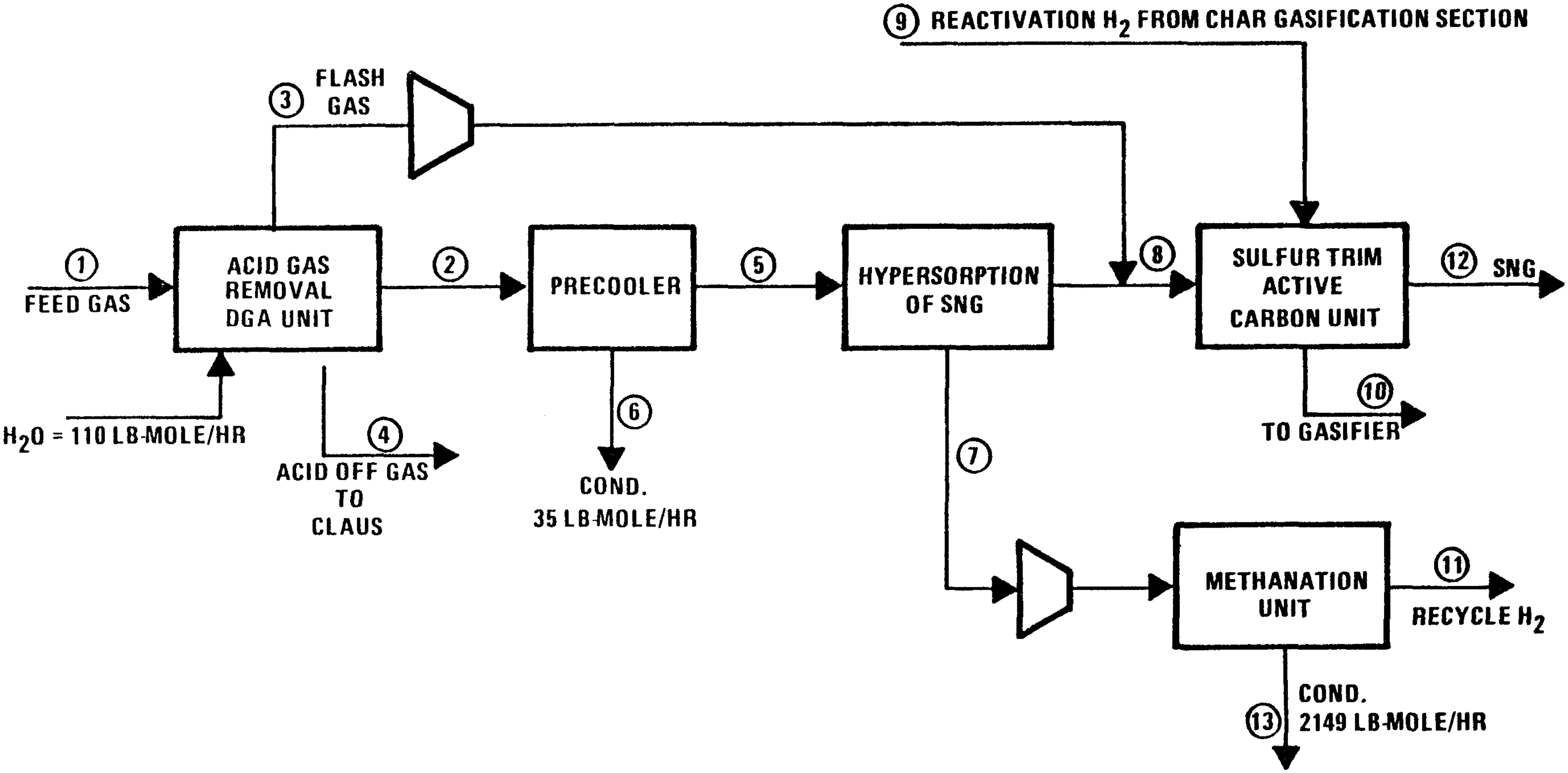

TABLE 3.2.4.3a
MATERIAL BALANCE OF DGA–HYPERSORPTION INTEGRATION

STREAM NO.		①	②	③	④	⑦	⑧	⑨ *	⑩ *	⑪	⑫
STREAM NAME		RAW ACID GAS FEED	TREATED GAS	FLASH FUEL GAS	ACID OFF-GAS TO CLAUS	HYPER-SORPTION OFF-GAS	SULFUR TRIM INLET	REACTIVATION H_2 FROM CHAR GASIFICATION	REACTIVATED H_2 TO GASIFIER	RECYCLE H_2	PRODUCT SNG
PRESS., PSIA		1420	1410	200	25	1395	1395	1625	1625	1615	1015
TEMP., °F		110	95	120	120	100	135	100	105	105	105
LB–MOLE/HR	CH_4	26,872.00	26,782.00	78.00	12.00		26,860.00			2,200.68	26,860.00
	H_2	63,493.00	63,421.00	62.00	10.00	63,421.00	62.00	10,247.00	10,247.00	56,821.57	62.00
	CO	2,230.00	2,227.00	3.00		2,204.00	26.00			0.71	26.00
	N_2	281.00	281.00		281.00					281.00	
	CO_2	117.00	4.64		112.60		4.64			2.61	4.64
	H_2S	392.00	0.37		391.87		0.37				
	COS	1.00	0.37				0.37				(≤4 ppm)
	CS_2	(0.13 ppm)	(0.13 ppm)				0.01				
	C_2H_2	(0.88 ppm)	(0.88 ppm)				0.08				0.08
	NH_3	3.00			3.00						
	HCN	2.00			2.00						
	H_2O	4.00	76.00		38.00	41.00				49.07	
	TOTAL	93,395.00	92,792.38	143.00	569.47	65,947.00	26,953.47	10,247.00	10,247.00	59,355.64	26,952.72
TOTAL, LBS/HR		648,189.0	629,380.0	1,460.4	19,313.0	65,947.00	432,011	20,658.0	20,658.0	158,749.	431,975.0

NOTE = * ON STREAM EVERY TWO HOURS FOR ONE HOUR PERIOD

Table 3.2.4.3.b: Utility Summary for DGA - Hypersorption Integration

1. Power

	KW
DGA	2,480
Hypersorption	32,820
Misc.	710
Subtotal	35,010

2. Steam

DGA	lbs/hr
60 psig steam	38,200
Hypersorption	
250 psia, 750°F steam	186,670 (for 12 hrs/day)

3. Fuel

Hypersorption	
Heater	517.1 MM Btu/hr

4. Cooling Water

	GPM
DGA	3,660
Hypersorption	49,110
Subtotal	52,770

5. Chemicals

DGA	
DGA	150 lbs/day
Active Carbon	22 lbs/day
Soda Ash	2,500 lbs/day
Hypersorption	
Active Carbon Loss	1,900 lbs/day
Active Carbon Replacement for Moving Bed	2,750 lbs/day
Active Carbon Replacement for Sulfur Removal	90 lbs/day
Methanator Catalyst Replacement	10 ft^3/day

FIGURE 3.2.4.4
PROCESS BLOCK DIAGRAM FOR SULFINOL ACID GAS REMOVAL AND HYPERSORPTION SNG SEPARATION

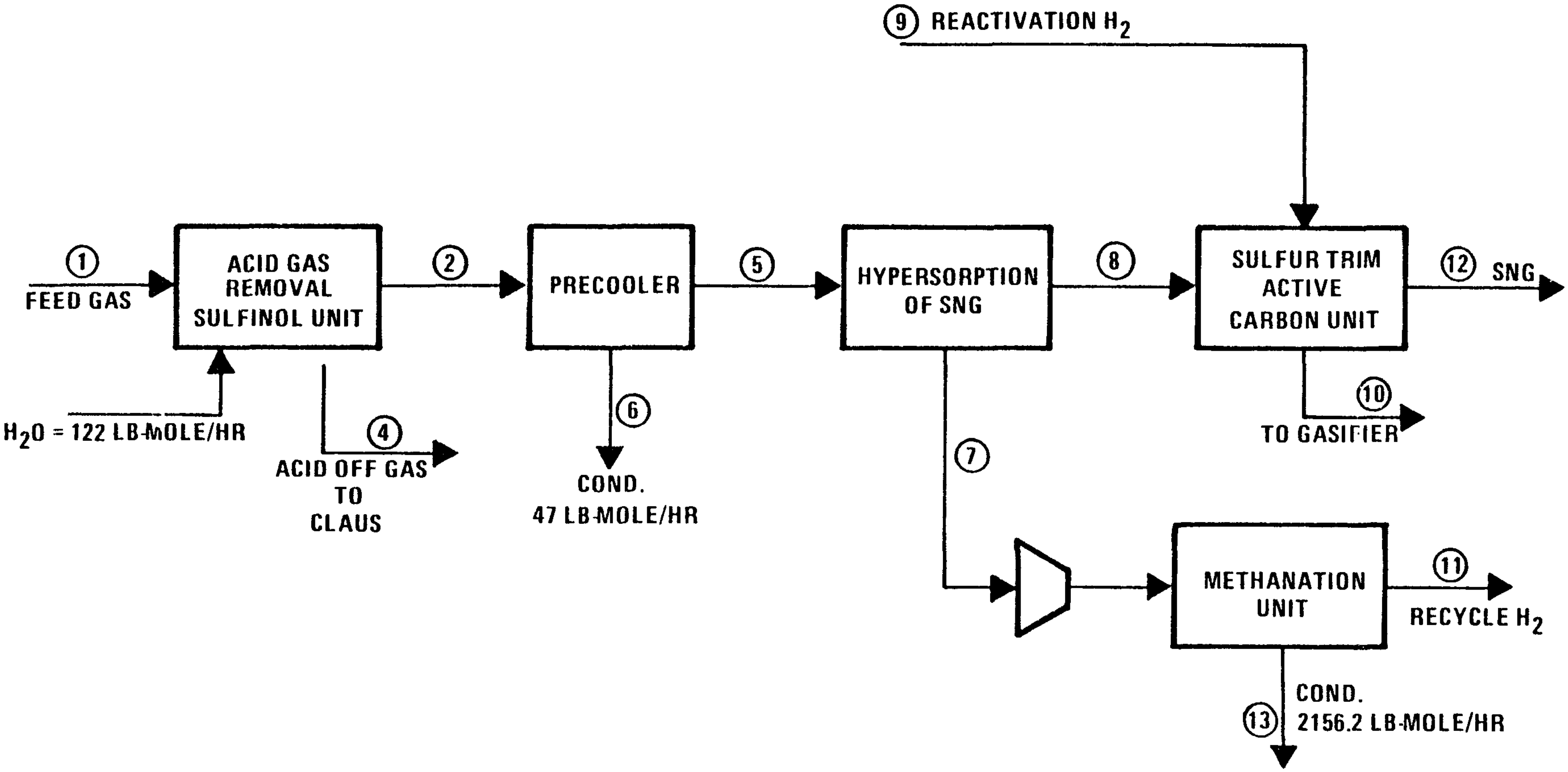

TABLE 3.2.4.4a
MATERIAL BALANCE OF SULFINOL–HYPERSORPTION INTEGRATION

STREAM NO.		①	②	④	⑦	⑧	⑨ *	⑩ *	⑪	⑫	
STREAM NAME		RAW ACID GAS FEED	TREATED GAS	ACID OFF-GAS TO CLAUS	HYPER-SORPTION OFF-GAS	SULFUR TRIM INLET	REACTIVATION H_2 FROM CHAR GASIFICATION	REACTIVATED H_2 TO GASIFIER	RECYCLE H_2	PRODUCT SNG	
PRESS., PSIA		1420	1410	25	1395	1395	1625	1625	1615	1015	
TEMP., °F		110	100	120	100	135	100	105	105	105	
LB–MOLE/HR	CH_4	26,872.00	26,863.00	9.00		26,863.00			2,202.68	26,863.00	
	H_2	63,493.00	63,478.00	15.00	63,478.00		10,069.00	10,069.00	56,872.57		
	CO	2,230.00	2,229.00	1.00	2,206.00	23.00			0.71	23.00	
	N_2	281.00	281.00		281.00				281.00		
	CO_2	117.00	5.85	111.15		5.85			2.61	5.85	
	H_2S	392.00	0.37	391.63		0.37				(≤ 4 ppm)	
	COS	1.00	0.09	0.91		0.09					
	CS_2	(0.13 ppm)	(0.10 ppm)			0.01					
	C_2H_2	(0.88 ppm)	(0.88 ppm)			0.08				0.08	
	NH_3	3.00		3.00							
	HCN	2.00		2.00							
	H_2O	4.00	88.00	38.00	41.00				49.09		
TOTAL		93,395.00	92,945.31	571.69	66,006.00	26,892.4	10,069.00	10,069.00	59,408.66	26,891.93	
TOTAL, LBS/HR		648,189.0	631,109.0	19,285.5	198,372.0	431,886.0	20,299.0	20,299.0	158,884.0	431,867.0	

NOTE: * ON STREAM EVERY TWO HOURS FOR ONE HOUR PERIOD

Table 3.2.4.4.b: Utility Summary for Sulfinol - Hypersorption Integration

1. Power

	KW
Sulfinol	2,070
Hypersorption	32,700
Misc.	700
Subtotal	35.470

2. Steam

Sulfinol	lbs/hr
100 psig steam	62,940
500 psig steam	32,300
Hypersorption	
250 psia, 750°F steam	173,340 (for 12 hrs/day)

3. Fuel

Hypersorption	
Heater	517.1 MM Btu/hr

4. Cooling Water

	GPM
Sulfinol	40
Hypersorption	49,090
Subtotal	49,130

5. Chemicals

Sulfinol	
DIPA	1,120 lbs/day
Sulfolane	30 lbs/day
Hypersorption	
Active Carbon	1,900 lbs/day
Active Carbon Replacement for Moving Bed	2,750 lbs/day
Active Carbon Replacement for Sulfur Removal	80 lbs/day
Methanator Catalyst Replacement	10 ft^3/day

FIGURE 3.2.4.5
AMINES (MEA, DEA, DGA) AND
CRYOGENIC INTEGRATION–SCHEME B

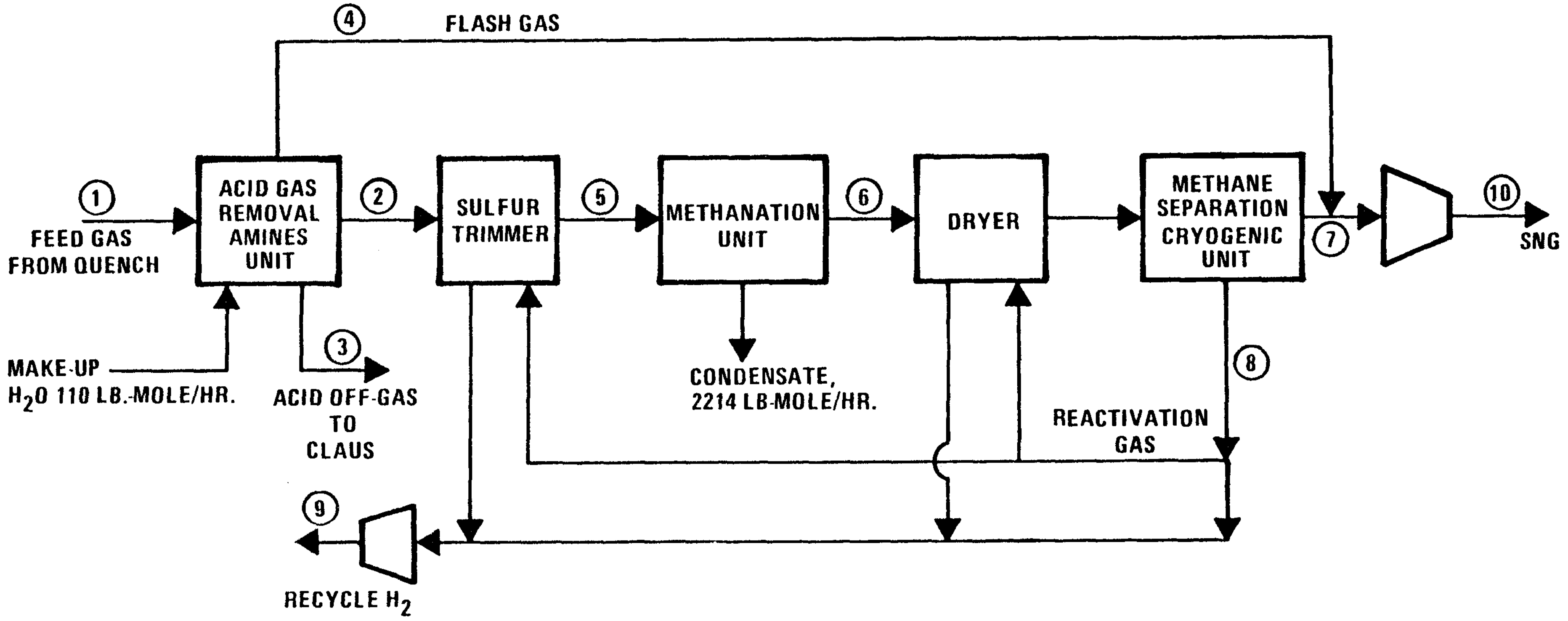

TABLE 3.2.4.5a
MATERIAL BALANCE FOR THE INTEGRATION OF AMINES (MEA, DEA, DGA) AND CRYOGENIC–SCHEME B

	STREAM NO.	①	②	③	④	⑤	⑥	⑦	⑧	⑨	⑩
	STREAM NAME	FEED GAS TO AMINE UNIT	TREATED GAS TO SULFUR TRIM	ACID OFF-GAS TO CLAUS UNIT	FLASH GAS	SULFUR TRIM OUTLET	DRYER INLET	CRYOGENIC OUTLET CH_4 STREAM	CRYOGENIC OUTLET H_2 STREAM	RECYCLE H_2	SNG PRODUCT
	PRESS, PSIA	1420	1410	25	200	1405	1385	18	388	1615	1015
	TEMP., °F	110	95	120	120	95	105	94	96	105	105
LB–MOLE/HR	CH_4	26,872.0	26,782.0	12.0	78.0	26,782.0	29,013.7	27,017.7	1,996.0	1,996.0	27,095.7
	H_2	63,493.0	63,421.0	10.0	62.0	63,421.0	56,722.0	1,320.0	55,402.0	55,402.0	1,382.0
	CO	2,230.0	2,227.0		3.0	2,227.0	≤(1 ppm)	≤(1 ppm)	≤(1 ppm)	≤(1 ppm)	3.0
	N_2	281.0	281.0			281.0	281.0	151.0	130.0	130.0	151.0
	CO_2	117.0	4.6	112.4		4.6	≤(1 ppm)				
	H_2S	392.0	0.4	391.6		≤(0.1 ppm)	≤(0.1 ppm)	≤(0.3 ppm)			≤(0.3 ppm)
	COS	1.0	≤(11 ppm)								
	CS_2	(0.13 ppm)									
	C_2H_2	(0.88 ppm)	(0.88 ppm)			(0.88 ppm)					
	NH_3	3.0		3.0							
	HCN	2.0		2.0							
	H_2O	4.0	76.00	38.0		76.0	98.0			98.0	
	TOTAL	93,395.0	92,792.0	569.0	143.0	92,791.6	86,114.7	28,488.7	57,528.0	57,626.0	28,631.7
	TOTAL, LBS/HR	648,189.0	629,416.0	19,295.0	1460.	629,342.0	589,456.0	440,336.0	147,354.0	149,120.0	441,796.0

Table 3.2.4.5.b: Utility Summary for MEA - Cryogenic Scheme B Integration

1. Power

	KW
MEA	1,640
Cryogenic Scheme B	97,820
Misc	690
Subtotal	100,150

2. Steam

	lbs/hr
MEA	
60 psig steam	59,890
150 psig steam	15,790
Cryogenic Scheme B	
250 psia, 750°F steam	204,170 (for 12 hrs/day)

3. Cooling Water

	GPM
MEA	2,610
Cryogenic Scheme B	53,820
Subtotal	56,430

4. Chemicals

MEA	
MEA	2,590 lbs/day
Active Carbon	50 lbs/day
Soda Ash	2,500 lbs/day
Cryogenic Scheme B	
Active Carbon Replacement for Sulfur Removal	2,240 lbs/day
Alumina Replacement for Dryer	215 lbs/day
Methanator Catalyst Replacement	25 ft^3/day

Table 3.2.4.6: Utility Summary for DEA - Cryogenic Scheme B Integration

1. Power

	KW
DEA	2,220
Cryogenic Scheme B	97,820
Misc	700
Subtotal	100,740

2. Steam

	lbs/hr
DEA	
60 psig steam	78,360
Cryogenic Scheme B	
250 psia, 750°F steam	204,170 (for 12 hrs/day)

3. Cooling Water

	GPM
DEA	2,610
Cryogenic Scheme B	53,820
Subtotal	56,430

4. Chemicals

DEA	
DEA	2,220 lbs/day
Active Carbon	16 lbs/day
Soda Ash	2,500 lbs/day
Cryogenic Scheme B	
Active Carbon Replacement for Sulfur Removal	2,240 lbs/day
Alumina Replacement for Dryer	215 lbs/day
Methanator Catalyst Replacement	25 ft^3/day

Table 3.2.4.7: Utility Summary for DGA - Cryogenic Scheme B Integration

1. Power

	KW
DGA	2,480
Cryogenic Scheme B	97,820
Misc	710
Subtotal	101,010

2. Steam

	lbs/hr
DGA	
60 psig steam	38,200
Cryogenic Scheme B	
250 psia, 750°F steam	204,170 (for 12 hrs/day)

3. Cooling Water

	GPM
DGA	3,660
Cryogenic Scheme B	53,820
Subtotal	57,480

4. Chemicals

DGA	
DGA	150 lbs/day
Active Carbon	22 lbs/day
Soda Ash	2,500 lbs/day
Cryogenic Scheme B	
Active Carbon Replacement for Sulfur Removal	2,240 lbs/day
Alumina Replacement	215 lbs/day
Methanator Catalyst Replacement	25 ft^3/day

FIGURE 3.2.4.8
SULFINOL AND CRYOGENIC INTEGRATION–
SCHEME B

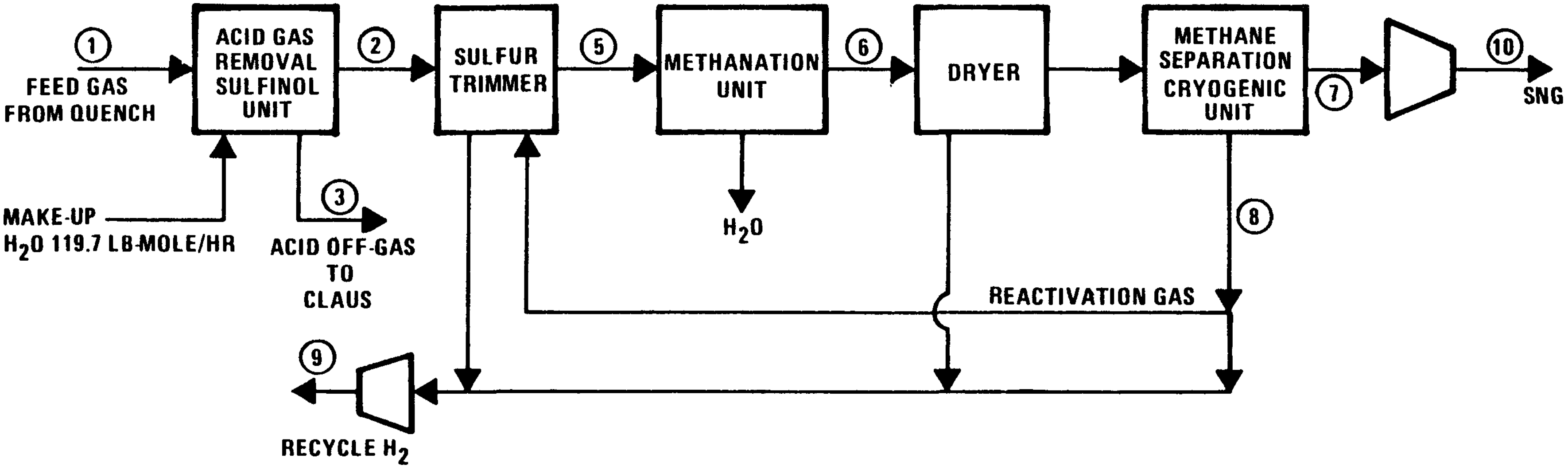

TABLE 3.2.4.8a
MATERIAL BALANCE FOR THE INTEGRATION OF SULFINOL AND CRYOGENIC–SCHEME B

STREAM NO.		①	②	③	⑤	⑥	⑦	⑧	⑨	⑩
STREAM NAME		FEED GAS TO SULFINOL	TREATED GAS TO SULFUR TRIM	ACID OFF-GAS TO CLAUS UNITS	SULFUR TRIM OUTLET	DRYER INLET	CRYOGENIC OUTLET–CH_4 STREAM	CRYOGENIC OUTLET–H_2 STREAM	RECYCLE H_2	SNG PRODUCT
PRESS, PSIA		1420	1410	27	1405	1385	18	388	1615	1015
TEMP., °F		110	100	120	100	105	94	96	105	105
LB–MOLE/HR	CH_4	26,872.0	26,863.0	9.0	26,863.0	29,098.0	27,102.0	1,9960	1,996.0	27,102.0
	H_2	63,493.0	63,478.0	15.0	63,478.0	56,767.4	1,365.4	55,402.0	55,402.0	1,365.4
	CO	2,230.0	2,229.0	1.0	2,229.0	≤(1 ppm)	≤(1 ppm)	≤(1 ppm)	≤(1 ppm)	≤(1 ppm)
	N_2	281.0	281.0		281.0	281.0	151.0	130.0	130.0	151.0
	CO_2	117.0	5.9	111.10	5.9	≤(1 ppm)				
	H_2S	392.0	0.4	391.6	≤(0.1 ppm)	≤(0.1 ppm)	≤(0.3 ppm)			≤(0.3 ppm)
	COS	1.0	0.1	0.9						
	CS_2	(0.13 ppm)								
	C_2H_2	(0.88 ppm)	(0.88 ppm)		(0.88 ppm)					
	NH_3	3.0		3.0						
	HCN	2.0		2.0						
	H_2O	4.0	88.0	35.7	88.0	98.0			98.0	
TOTAL		93,395.0	92,945.4	569.3	92,944.9	86,244.4	28,618.4	57,528.0	57,626.0	28,618.4
TOTAL, LBS/HR		648,189.0	631,105.0	19,240.0	631,042.0	590,900.0	441,780.0	147,354.0	149,120.0	441,780.0

Table 3.2.4.8.b: Utility Summary for Sulfinol - Cryogenic Scheme B Integration

1. Power

	KW
Sulfinol	2,070
Cryogenic Scheme B	97,820
Misc	700
Subtotal	100,590

2. Steam

Sulfinol	lbs/hr
150 psig steam	62,940
500 psig steam	32,300
Cryogenic Scheme B	
250 psia, 750°F steam	204,170 (for 12 hrs/day)

3. Cooling Water

	GPM
Sulfinol	40
Cryogenic Scheme B	53,820
Subtotal	53,860

4. Chemicals

Sulfinol	
DIPA	1,120 lbs/day
Sulfolane	30 lbs/day
Cryogenic Scheme B	
Active Carbon Replacement for Sulfur Removal	2,240 lbs/day
Alumina Replacement for Dryer	215 lbs/day
Methanator Catalyst Replacement	25 ft^3/day

TABLE 3.2.6.1

TOTAL CAPITAL REQUIREMENT ($M)

	WITH CRYOGENIC UNIT				WITH HYPERSORPTION UNIT			
	DGA	SULFINOL	DEA	MEA	DGA	SULFINOL	DEA	MEA
Acid Gas Removal Section	6,008	9,870	8,410	7,172	6,008	9,870	8,410	7,172
Methanation	6,950	6,950	6,950	6,950	6,717	6,717	6,717	6,717
Sulfur Trim	10,400	10,400	10,400	10,400	4,320	3,945	4,320	4,046
Hypersorption	--	--	--	--	32,000	31,700	32,000	32,000
Cryogenic Separation Section								
Adsorber/Drier System	1,646	1,646	1,646	1,646				
Cryogenic Unit	5,224	5,224	5,224	5,224				
Compression	26,480	26,480	26,480	26,480				
Total	33,350	33,350	33,350	33,350				
Subtotal - Installed Equipment	56,708	60,570	59,110	57,872	49,045	52,232	51,447	49,935
Contractors Fee (3% Installed Equip. plus Engineering)	1,800	1,940	1,890	1,840	1,600	1,700	1,700	1,640
Engineering (APCI estimate)	3,350	3,973	3,782	3,555	4,770	4,930	4,930	4,818
Subtotal	61,858	66,483	64,782	63,267	55,415	58,862	58,077	56,393
Contingency(1)	9,280	9,970	9,717	9,490	13,850	14,715	14,500	14,100
Total Plant Investment (TPI)	71,138	76,453	74,499	72,757	69,265	73,577	72,577	70,493
Initial Charge of Chems./Cats.	5,110	5,330	5,110	5,081	3,990	3,950	3,990	3,890
Allowance for funds used during construction (STI)	8,360	8,950	8,720	8,540	8,060	8,520	8,410	8,190
Start-up Costs (20% of gross operating Costs)	8,864	9,311	9,144	9,110	8,790	9,150	9,030	8,840
Subtotal - Total Investment	93,472	100,044	97,473	95,488	90,105	95,197	94,007	92,513
Working Capital (WC)	2,990	3,167	3,100	3,060	2,930	3,075	3,035	2,990
Total Capital Requirement (TCR)	96,462	103,211	100,573	98,548	93,035	98,272	97,042	95,503

(1) 15% for cryogenic; 25% for hypersorption

TABLE 3.2.6.2a

OPERATING COSTS ($M/yr) 90% onstream = 7,884 hrs/yr

	WITH CRYOGENIC UNIT				WITH HYPERSORPTION UNIT			
	DGA	SULFINOL	DEA	MEA	DGA	SULFINOL	DEA	MEA
Capital ($M)								
Investment	93,341	99,901	97,325	95,356	90,105	95,197	94,007	91,454
Working Capital	2,976	3,152	3,086	3,052	2,930	3,075	3,035	2,960
Total Capital	96,317	103,053	100,411	98,408	93,035	98,272	97,042	94,414
Operating Cost ($M/yr)								
Power @ $.035/KWH	27,694	27,576	27,617	27,454	9,937	9,786	9,861	9,700
Steam - 250 psig, @ $4.78/m-lb	3,800	3,300	3,800	3,800	3,520	3,270	3,520	3,330
500 psig, @ $4.37/m-lb	--	1,117	--	--	--	1,117	--	--
150 psig, @ $3.50/m-lb	--	1,737	--	436	--	1,737	--	436
50 psig, @ $2.80/m-lb	845	--	1,730	1,300	845	--	1,730	1,300
Cooling water, @ $.10/m-gal. circ.	2,573	2,502	2,624	2,624	2,496	2,324	2,446	2,446
Solvent Losses	99	180	304	347	99	180	304	349
Absorbent/Catalyst Replacement	2,650	2,700	2,850	2,906	1,582	1,551	1,580	1,589
Carbon Losses	--	--	--	--	687	687	687	687
Fuel, $4.50/MMBTU	--	--	--	--	18,346	18,346	18,346	18,346
Labor-Operating (20 men @ $30M/yr)	600	600	600	600	600	600	600	600
Maintenance (60% of Tot. Maintenance)	1,280	1,376	1,340	1,310	1,250	1,324	1,306	1,270
Supervision (20% of Operating & Maint. Labor)	376	395	388	382	370	385	380	375
Supplies - Operating (30% of Operating Labor)	180	180	180	180	180	180	180	180
Maintenance (40% of Total Maintenance)	854	918	894	874	830	983	870	850
Taxes & Insurance (2.7% of TPI)	1,920	2,063	2,011	1,963	1,870	1,987	1,960	1,903
Administration & Gen. Overhead (60% of Total Labor)	1,350	1,411	1,380	1,370	1,330	1,385	1,370	1,350
Total Gross Operating Costs, $M/yr	44,321	46,555	45,718	45,546	43,942	45,742	45,140	44,711

TABLE 3.2.6.2b

UTILITY SUMMARY

	WITH CRYOGENIC UNIT				WITH HYPERSORPTION UNIT			
	DGA	SULFINOL	DEA	MEA	DGA	SULFINOL	DEA	MEA
Power, KW								
Acid Gas Section	2,488	2,070	2,216	1,640	2,488	2,070	2,216	1,640
Methane Separation Section	95,904	95,904	95,904	95,904	32,819	32,700	32,819	32,819
Miscellaneous (2%)	1,970	1,960	1,962	1,950	706	695	701	690
Total	100,362	99,934	100,082	99,494	36,013	35,465	35,736	35,149
Steam, lbs/hr								
Acid Gas Section								
500 psig	--	32,200	--	--	--	32,200	--	--
150 psig	--	62,940	--	15,788	--	62,940	--	15,788
50 psig	38,200	--	78,360	58,890	38,200	--	78,360	58,890
Methane Separation Section 250 psig (750°F)	102,080	102,080	102,080	102,080	93,333	86,666	93,333	88,333
Cooling water, GPM								
Acid Gas Section	3,660	44	2,610	2,610	3,660	44	2,610	2,610
Methane Separation Section	53,816	53,816	53,816	53,816	49,108	49,084	49,108	49,108
Total	57,476	53,860	56,426	56,426	52,768	49,128	51,718	51,718
Solvent Losses, lbs/hr								
Acid Gas Section	6	520	72	108	6	520	92	108
Adsorbent & Catalyst Replacement, lbs/yr								
Acid Gas Section	631,430	--	629,350	640,430	631,430	--	629,350	640,430
Methane Separation Section								
Carbon, Sulfur Bed	736,100	736,100	736,100	736,100	27,130	24,417	27,130	25,095
Alumina (A-201)	70,550	70,550	70,550	70,550	--	--	--	--
Methanator Catalyst (ft^3/yr)	8,100	8,100	8,100	8,100	3,422	3,422	3,422	3,422
Carbon, Activated	--	--	--	--	904,600	904,600	904,600	904,600
Fuel, MMBTU/hr	--	--	--	--	517.1	517.1	517.1	517.1

TABLE 3.2.6.3a

TREATING COSTS

UTILITY FINANCING METHOD (20 YR. AVERAGE COST)

	WITH CRYOGENIC UNIT				WITH HYPERSORPTION UNIT			
	DGA	SULFINOL	DEA	MEA	DGA	SULFINOL	DEA	MEA
Total Gross Operating Costs	44,321	46,555	45,718	45,546	43,942	45,742	45,140	44,711
Capital Charges, $M/yr								
Depreciation (20 yr S.L.)	4,674	5,000	4,874	4,774	4,500	4,750	4,700	4,625
L.T. Interest (10% Ave. Net Investment)	3,730	3,990	3,887	3,810	3,600	3,300	3,750	8,690
Taxes (49.5% = Fed., State, Local)	1,828	1,955	1,906	1,870	1,765	1,862	1,843	1,310
NPAT (15% Return on Equity Capital)	1,865	1,995	1,944	1,906	1,800	1,900	1,880	1,850
Total Capital Charges, $M/yr	12,097	12,940	12,611	12,360	11,665	12,322	12,173	11,975
Total Required Revenues, $M/yr	56,418	59,494	58,329	57,906	55,607	58,064	57,313	56,686
Production, MMSCF/yr	85,547	85,547	85,547	85,547	80,522	80,336	80,522	80,522
Treating Cost, $/MSCF	0.660	0.696	0.682	0.677	0.691	0.723	0.712	0.704
Treating Cost, $/MMBTU	0.679	0.716	0.702	0.697	0.686	0.716	0.707	0.699

TABLE 3.2.6.3b

TREATING COSTS

DCF - CCU - A.G.A METHOD

	WITH CRYOGENIC UNIT				WITH HYPERSORPTION UNIT			
	DGA	SULFINOL	DEA	MEA	DGA	SULFINOL	DEA	MEA
Total Gross Operating Costs	44,321	46,550	45,718	45,546	43,942	45,574	44,975	44,534
Capital Charges	24,086	25,783	25,115	24,504	23,212	24,527	24,221	23,858
Total Required Revenues, $M/yr	68,407	72,333	70,833	70,150	67,154	70,101	69,196	68,392
Production, MMSCF/yr	85,547	85,547	85.547	85,547	80,522	80,336	80.522	80.522
Treating Cost, $/MSCF	0.800	0.846	0.828	0.820	0.834	0.873	0.859	0.849
Treating Cost, $/MMBTU	0.823	0.871	0.853	0.844	0.828	0.865	0.853	0.843

TABLE 3.2.6.3c

TREATING COSTS

DCF - AIR PRODUCTS METHOD

	WITH CRYOGENIC UNIT				WITH HYPERSORPTION UNIT			
	DGA	SULFINOL	DEA	MEA	DGA	SULFINOL	DEA	MEA
Total Gross Operating Costs	44,321	46,550	45,718	45,546	43,942	45,574	44,975	44,534
Capital Charges	10,166	11,350	11,847	10,476	10.732	12,107	11,390	11,429
Total Required Revenues, $M/yr	55,606	58,685	57,488	57,060	54,674	57,681	56,365	55,963
Production, MMSCF/yr	85,547	85,547	85.547	85.547	80.522	80.336	80,522	80,522
Treating Cost, $/MSCF	0.650	0.686	0.672	0.667	0.679	0.718	0.700	0.695
Treating Cost, $/MMBTU	0.669	0.706	0.709	0.687	0.674	0.713	0.695	0.690

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