SECTION 16

PREDESIGN STUDIES

Means of improving the efficiency/economics of all major portions of the complex were analyzed before establishing the final design configuration. In most cases, the economics of alternatives were compared; in the normal case, the differential product cost in dollars per million Btu was estimated. The economic prediction was one important input to the selection of the preferred process step.

Computer-assisted process design, fixed capital estimation, and profitability analyses were used in the development of the economic comparisons for these studies.

Technical and economic information contained in OCR R&D Report No. 82, Volumes I, II, and III, titled Demonstration Plant - Clean Boiler Fuels from Coal[14,15,16] was used as a basis of comparison. The process alternatives, their equipment, capital cost, and economic results were evaluated at the capacity of this plant, which processes 10,000 TPD using SRC-II technology. The alternatives were compared to updated capital cost estimates and economics of the report where applicable.

This section illustrates the results of studies that were completed.

16.1 ADDITIONAL SNG PRODUCTION BY LIGHT ENDS REFORMING

An objective of this design effort is production of significant quantities of SNG. Five alternatives were therefore investigated to increase the percentage of SNG produced. The results of this screening work showed that three of the methods were of marginal value or that high cost would eliminate them as candidate process alternatives.

The cases examined in detail were:

- Case A - Production of SNG and LPG from the Complex Offgases - Produce the plant fuel requirements in a low-Btu gasifier especially provided for the purpose. Figure 16-1 shows the block flow diagram required for this case.

- Case B - Production of SNG Only from the Complex Offgases - Reform and methanate the LPG components to utilize them as SNG. Figure 16-2 shows the block flow diagram required for this case.

Table 16-1 summarizes the economic comparison results for the two cases. The economic comparison shows a reduction of approximately 3% of the required selling price for the coproduction of both LPG and SNG as saleable products; these savings are within the accuracy of the estimate.

The alternative of separate LPG and SNG coproduction was selected for the design because of its simpler configuration.

16.2 HYDROGEN VIS-A-VIS SYNGAS AS DISSOLVER FEED

R&D Report No. 82 used syngas produced in a gasifier as a hydroliquefaction agent in the dissolvers. The gas subsequently produced in the dissolver section was used as inplant boiler fuel. This offgas is rich in carbon monoxide and relatively poor in hydrogen.

To meet the objective of producing significant quantities of SNG, the use of hydrogen-rich gas as a liquefaction agent was investigated. This method increases the concentration of hydrogen in the dissolver offgases, which facilitates the separation of the hydrogen from the carbon monoxide in a cryogenic separation unit.

Figure 16-3 shows a block flow diagram containing the units required to produce and use the hydrogen-rich gas. The total constructed cost of these facilities represents a 5.6% increase in total plant fixed capital investments over similar facilities in the R&D Report No. 82 plant. Table 16-2 shows the economic comparison of the two cases, including the influence of operating costs, catalyst and chemical cost, and coal consumption, and shows that the use of syngas requires a slightly lower product selling price (approximately 3%) as compared to the use of hydrogen.

Considering the range of accuracy of the estimate, the choice of the hydroliquefaction agent was therefore elective and subject to process and operations considerations. It was decided to use the higher hydrogen purity case.

16.3 USE OF RECYCLE SLURRY VIS-A-VIS FILTRATE AS COAL SLURRY AGENT

The R&D Report No. 82 design was based on using unfiltered dissolver product as the vehicle for slurrying the coal feed. This can be termed the slurry recycle method.

The hydrogen consumption for the case in which clear filtrate is used to slurry the feed coal can be in the range of 2 wt.% of the feed coal, but the slurry recycle mode can increase the hydrogen uptake to the range of 3 wt %. As a result of this increased hydrogen consumption, the product slate will tend to produce liquid fuels.

For the purposes of this comparison, the potential increase of the SNG production was not considered. The product slate was restricted to liquid products and all gases evolved are used in the plant as fuel. The difference of energy available as products is 1.8 x 10^9 Btu/D, which amounts to 1% of the total Btu value generated in the plant. This difference is well within the accuracy of the calculation of the total heat available. It was therefore concluded that the energy efficiencies for the two modes of operation are essentially equivalent and that the choice of slurry or nonslurry recycle depends primarily upon the overall product slate desired for the complex.

The recycle slurry method, SRC-II mode, was selected for the design.

16.4 REDUCTION OF DISSOLVER RESIDENCE TIME

Experimental runs in the Tacoma SRC Pilot Plant showed that high coal conversions can be obtained at relatively low liquid residence time in the dissolvers. The economic impact of reducing the nominal liquid space time in the dissolvers from 60 to 30 minutes was studied. The installed cost of the affected equipment for the R&D Report No. 82 design as escalated to 1975 was $41 million. The cost for the same equipment, considering the reduced dissolver residence time, was $32 million. The required annual revenue is reduced by $2.8 million per year, or approximately 1.7% of the total base required annual revenue.

A reduced dissolver residence time was used in the design.

16.5 ACID GAS REMOVAL

The R&D Report No. 82 design used a chemical absorption process to separate the hydrogen sulfide from the gas stream to produce an ecologically acceptable fuel gas. Considering the greater quantities to be treated for the Oil/Gas design, several physical solvent processes used for the same purpose were investigated. Quotations were obtained from two potential licensors of physical solvent separation processes, and it was found that the capital investment for both these processes is approximately $1.8 million lower for a 10,000-TPD coal feed plant than the chemical absorption process.

Considering catalyst usage, utilities, and other economic factors (see Table 16-3), the use of a physical solvent process reduces the annual revenue requirements.

A physical solvent process was used for hydrogen sulfide removal in the Oil/Gas plant design.

16.6 SOUR VIS-A-VIS SWEET SHIFT

A shift operation is required to increase the ratio of hydrogen to carbon monoxide in the gasifier gas product to make it suitable for production of high-purity hydrogen to be used in the dissolvers. The reaction used is:

$$CO + H_2O = H_2 + CO_2$$

The use of sour vis-a-vis sweet shift was studied to determine which of the two processes is the more economical for the Oil/Gas plant design. Figure 16-4 shows the two process configurations.

In the sweet shift configuration, gasifier product gas is cooled to 100°F for treating in an acid gas removal unit. Nearly all of the water present in the gas is condensed. Following acid gas removal, the gas is reheated to shift temperature (650 to 700°F) and steam is added. After the shift reaction, the gases are cooled again prior to removal of the carbon dioxide produced in shifting. Product gas is then available for use.

In the sour shift configuration, gasifier offgas is fed directly at 700°F to the shift unit with additional steam feed to adjust the steam-to-dry gas ratio. The shift product gas is then cooled for acid gas removal and process use.

The total acid gas removal burden is the same in both shift schemes. Sour shift offers the advantage of eliminating two sets of heat exchangers and one acid gas removal unit. Furthermore, it reduces the steam requirements by not condensing steam ahead of the shift unit. However, the sour shift unit must use a greater quantity of a more expensive catalyst. In addition, steam requirements in the sour shift unit are greater than in the sweet shift unit as a result of the presence of carbon dioxide in the feed. A single unit is used in the sour shift case, but two smaller units are used in the sweet shift case.

Fixed capital investment and operation costs for the two cases were estimated. Results indicate that the use of a sour shift procedure should reduce the fixed capital investment by approximately $2.2 million. Expected utility requirement reductions for sour vis-a-vis sweet shift are:

- Fuel gas: 763 MM Btu/yr
- Steam: 1,228.3 MM lb/yr
- Power: (36 M kWh/yr an increase)

Table 16-4 is a summary of the economic factors; it shows that the use of the sour shift procedure will reduce the required annual revenue by more than $6 million per year, equivalent to about 4% of the total required annual revenue.

The sour shift procedure was selected for the Oil/Gas plant design.

16.7 FILTER CAKE SOLVENT RECOVERY

In the R&D Report No. 82 design the wet filter cake, with projected 50 wt % solids and 50 wt % liquids, was sent directly to the gasifier with a resultant conversion of solvents or wash oils adhering to the filter cake to syngas. This was considered as one alternative for this study. As a potential improvement, an alternative was studied consisting of drying the filter cake and recovering the liquids adhering to the cake. The cake and coal would then be fed to the gasifier. Figure 16-5, block flow diagram, shows the major components of the system.

For the filter cake drying alternative, a second side-stripper was added to the main fractionator to recover a kerosene-range filter wash oil. This cut is light enough to be easily removed from the filter cake in a dryer but not so light as to vaporize in the filter and fail to wash off the adhering liquids. This wash oil is sent to the filter to wash through the filter cake and displace adhering filtrate. A volume of wash oil equal to twice the volume of the adhering liquid was used, and it was predicted that 97 wt % of the original adhering liquid would be displaced. The resultant wet filter cake is projected to be 50 wt % solids and 50 wt % liquids. Since the filtrate includes a large portion of the wash oil and additional recovered filtrate, the main fractionator and attendant equipment were sized to accommodate the larger flows.

Wet filter cake is dried in a rotary dryer with a circulating stream of heated gas used to provide heat and to remove vapors from the dryer. The wash oil is recovered as a liquid and the dried filter cake is mixed with raw coal and is fed to the gasifier. Table 16-5 shows the material and utility balance for the two alternative cases studied.

The addition of filter cake washing and drying increases the total capital investment cost of the plant by approximately $12 million, and the required annual revenue is also increased by about $14 million as shown in Table 16-6. The output of the complex is increased by approximately 37 x 10^9 Btu/day. The required selling price of the products is reduced by $0.435/MM Btu, a reduction of approximately 13%.

Filter cake drying was included in the design.

16.8 USE OF POWER RECOVERY TURBINES

The Clean Boiler Fuels from Coal design contains several streams that have to be depressurized. The incentives for using this energy to drive power-recovery turbines were investigated. Approximately 90% of the pressure drop of the streams with sufficient energy potential was utilized for control purposes. This procedure in turn reduces the duty requirements of the pressure letdown valves.

The analysis indicates that the economic impact of the use of power-recovery turbines is small. The reduction in required selling price is less than 0.5%. However, the technical advantage of reducing the duty imposed on the pressure-reducing valves was considered sufficient to use the power-recovery turbines in several applications in the Oil/Gas plant design.

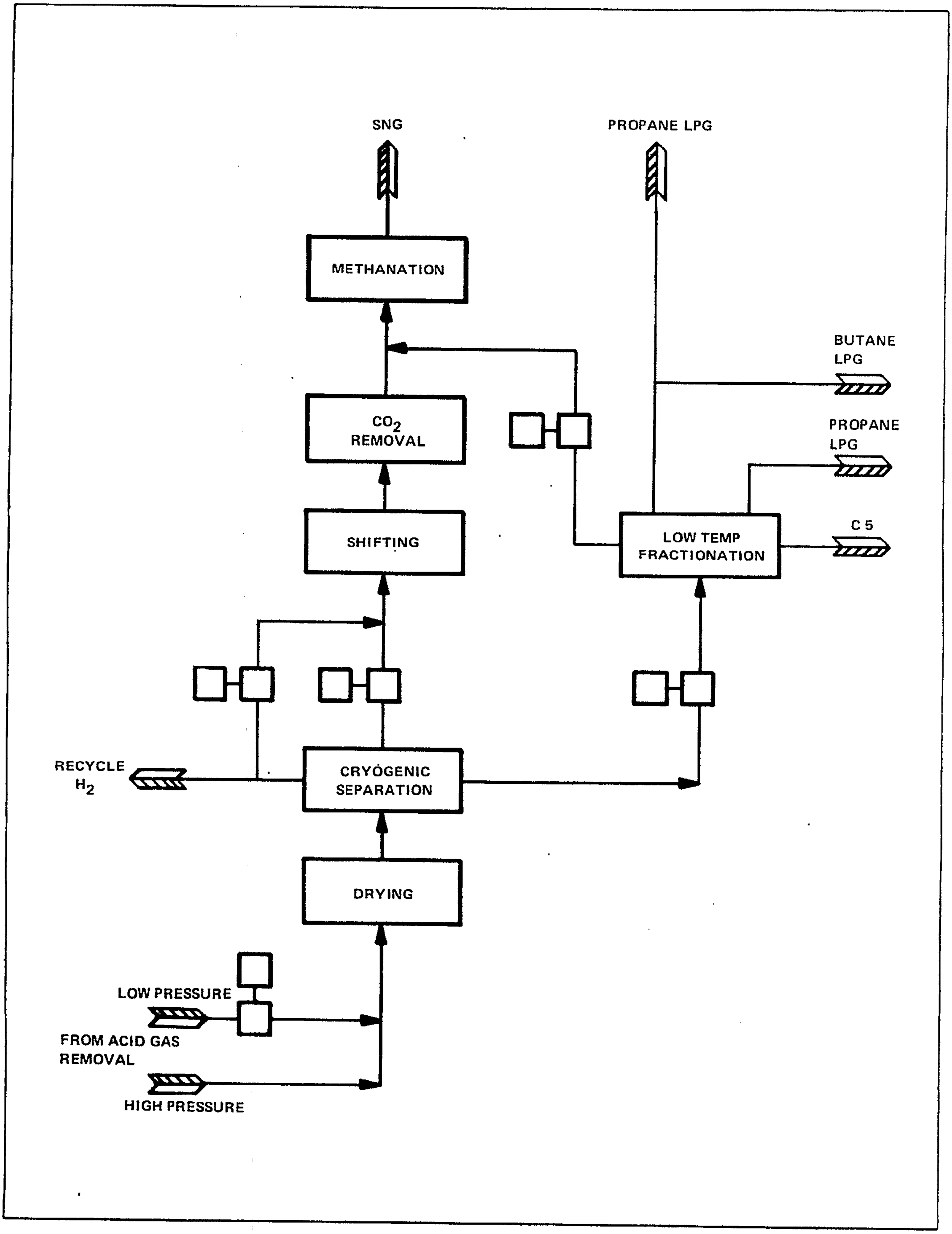

Figure 16-1 - SNG and LPG Production
Schematic Diagram

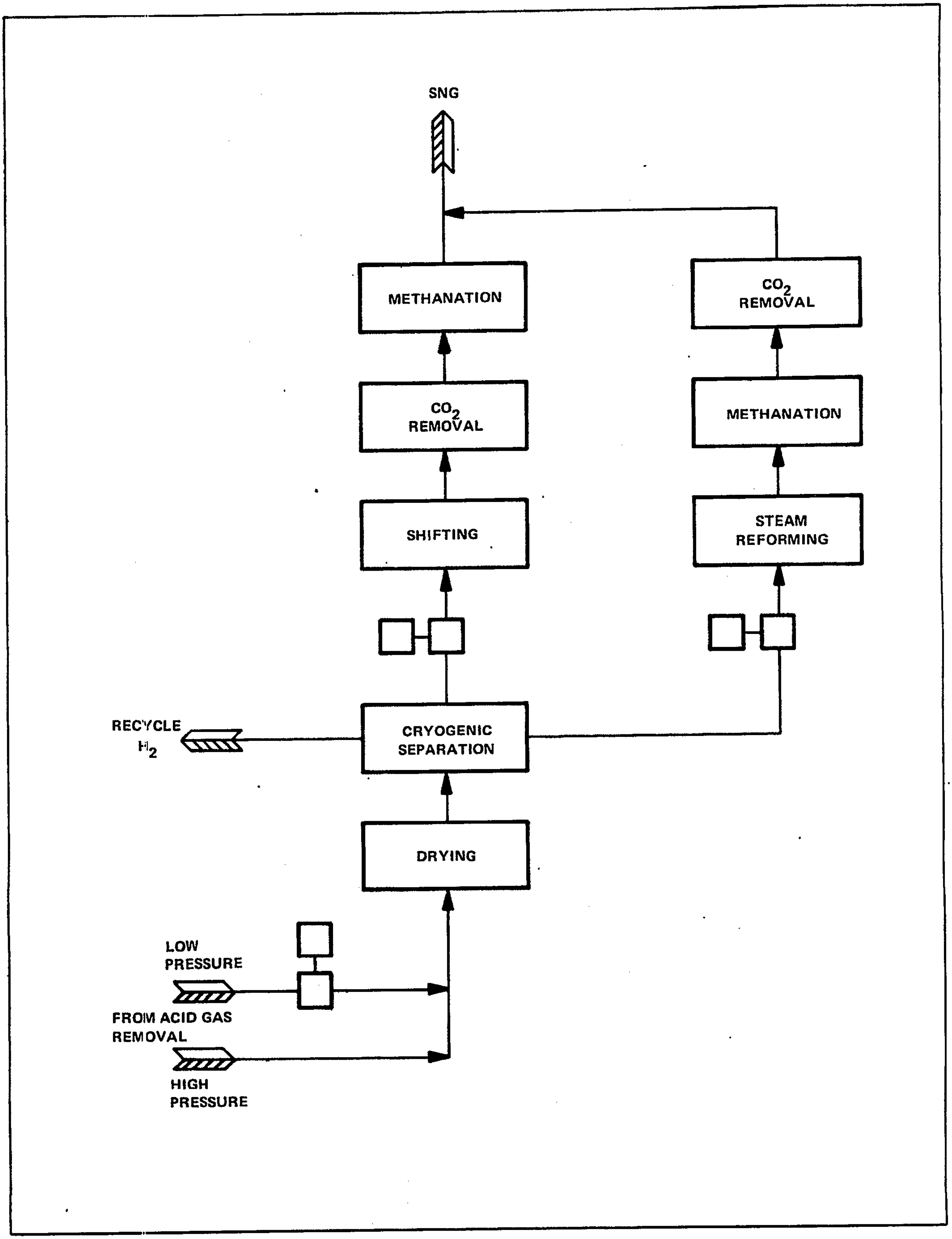

Figure 16-2 - SNG Production (No LPG) Product) Schematic Diagram

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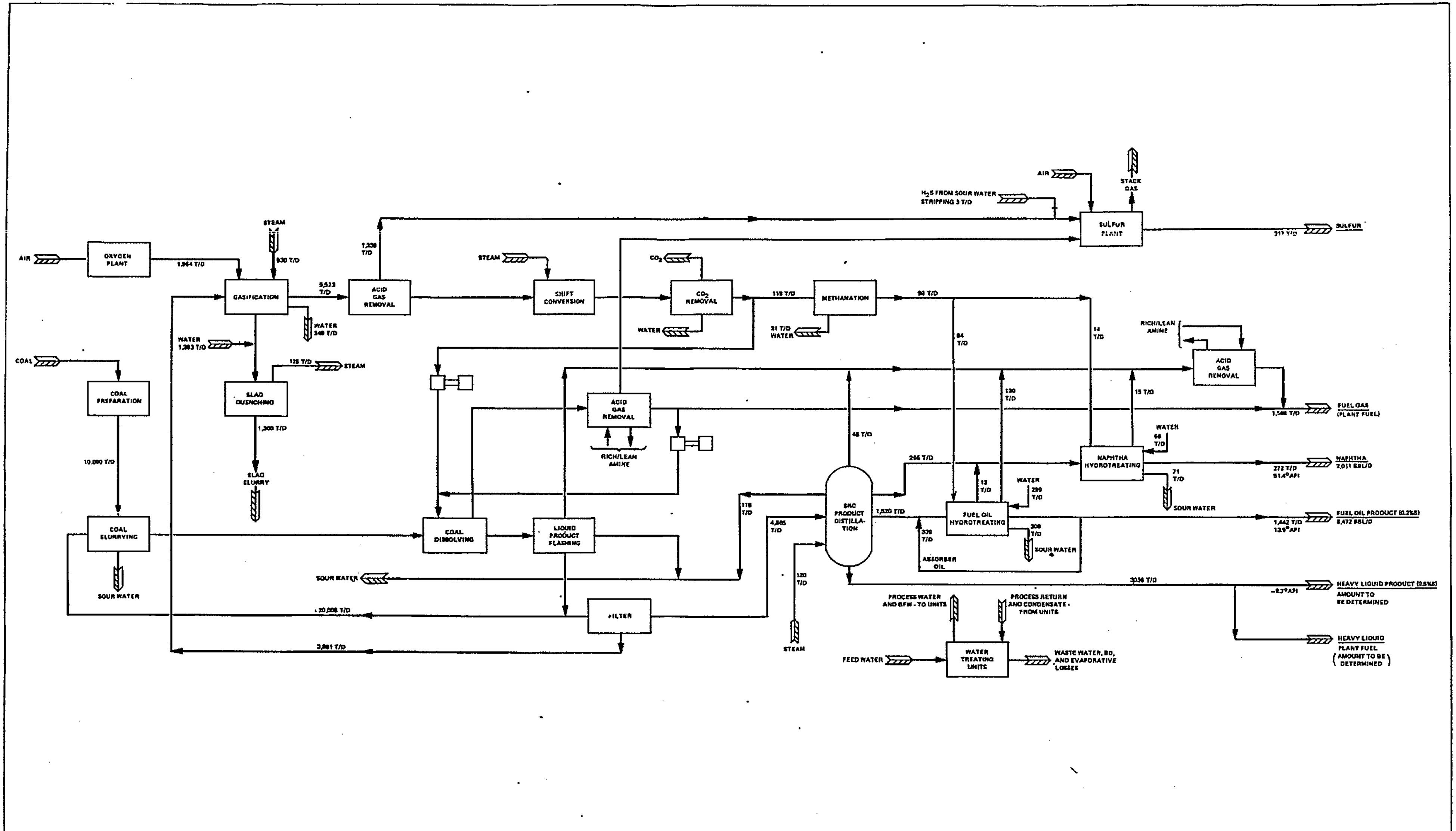

Figure 16-3 - H_2 Dissolver Feed

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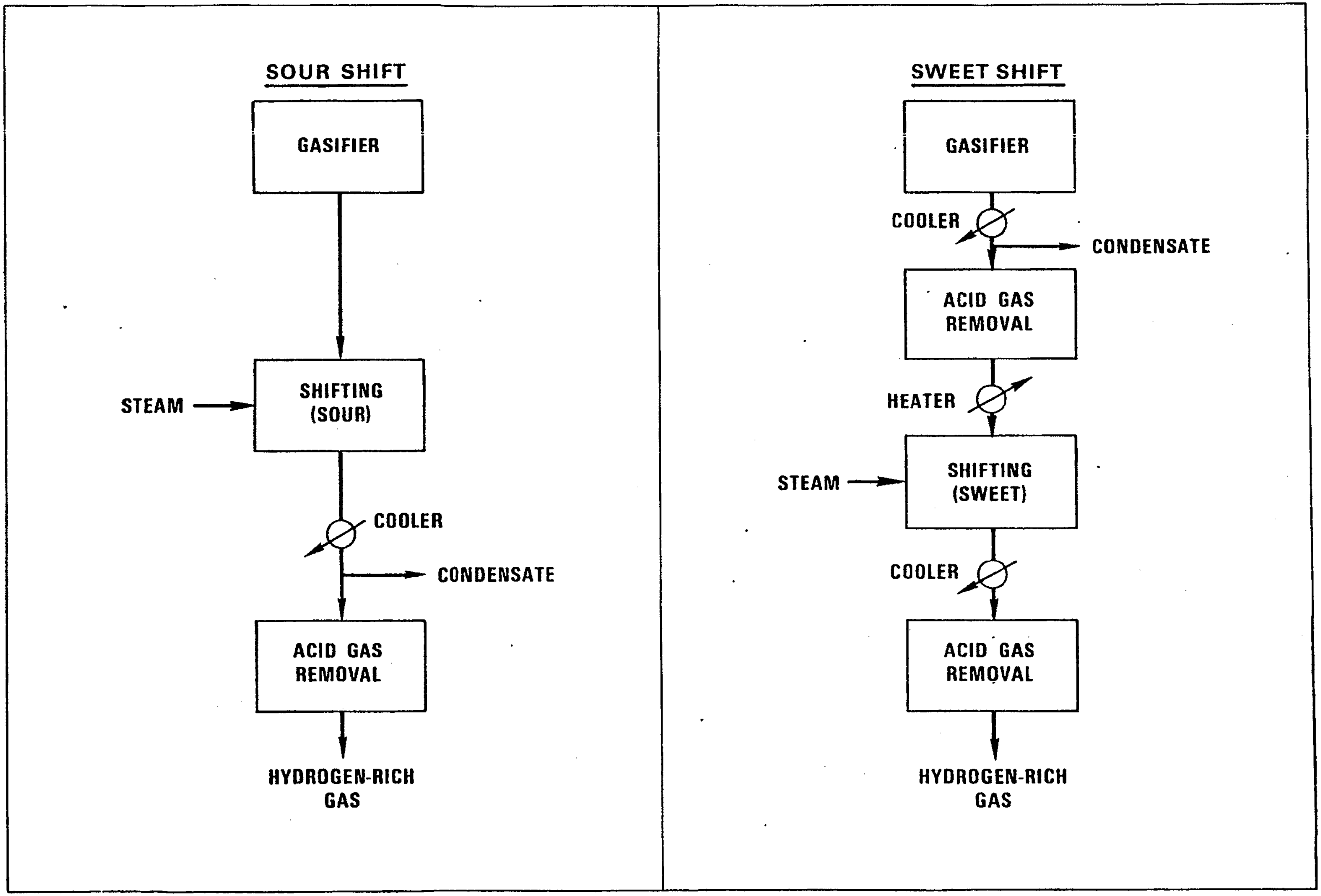

Figure 16-4 - SRC Process Sour Vis-a-Vis
Sweet Shift, Process Configuration

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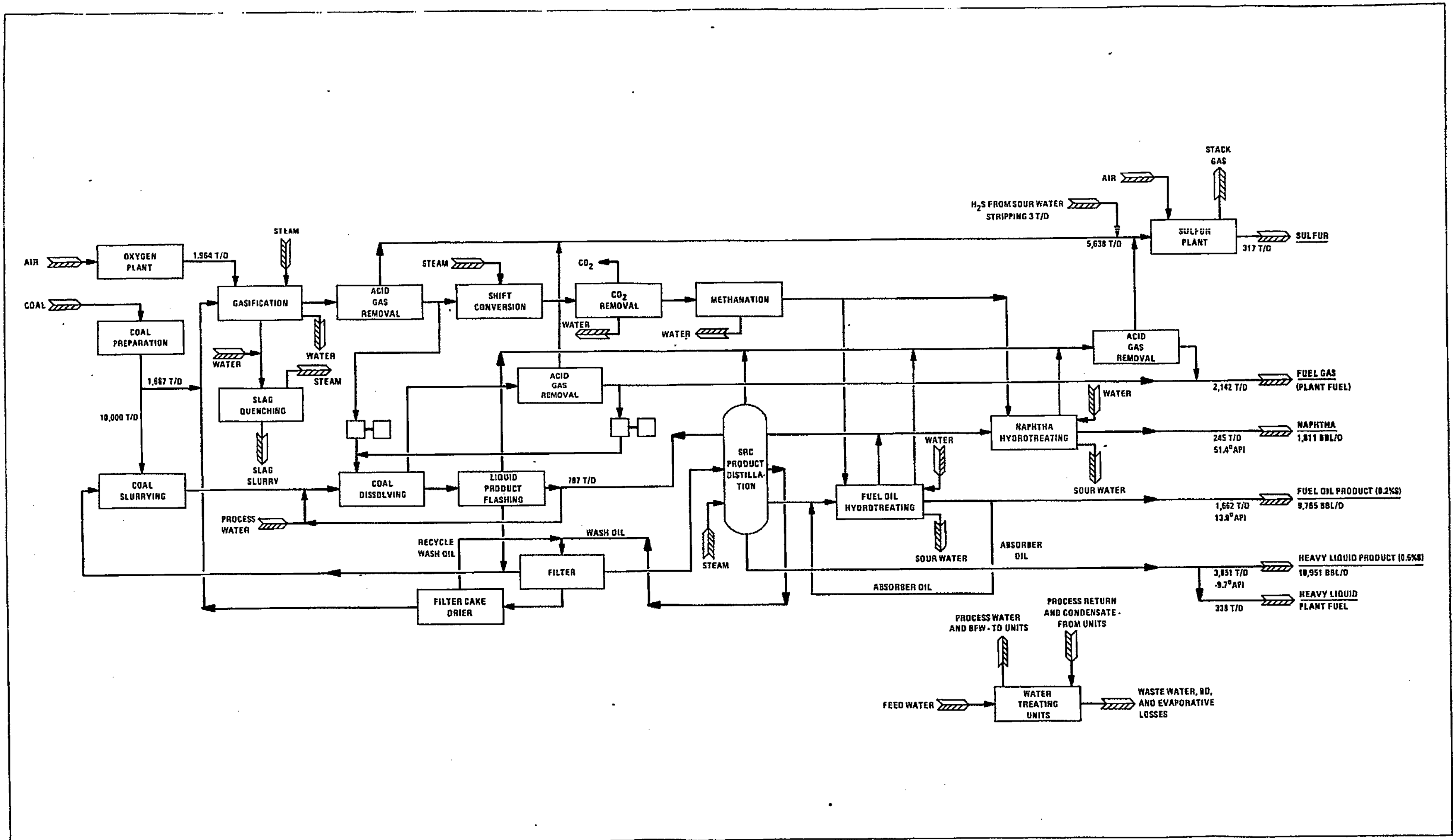

Figure 16-5 - Preliminary Block Flow Diagram Filter Cake Drying

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Table 16-1 - Additional SNG Production
EUAC at 12% DCF

	Updated Demonstration Plant		Case A (LPG+SNG)		Case B (SNG Only)	
		$MM/yr		$MM/yr		$MM/yr
Costs						
Coal	10,000 TPSD	39.6	10,000 TPSD	39.6	10,000 TPSD	39.6
Catalysts and Chemicals		4.4		4.7		5.0
Operating Labor		3.4		3.4		3.4
Maintenance Labor		3.7		3.8		3.9
Payroll Burden		2.5		2.5		2.7
Plant Overhead		5.7		5.8		6.3
Maintenance Materials		7.5		7.8		7.8
Utilities		Incl		Incl		Incl
Property Tax and Insurance		8.6		8.9		9.0
G&A Overhead		1.1		1.2		1.2
Total		76.5		77.7		78.9
Income Tax		39.4		41.0		41.3
Investment, $MM						
Fixed Capital	311	48.3	323	50.2	326	50.7
Initial Catalysts and Chemicals		0.3	2.591	0.3	2.541	0.3
Startup Costs	20	2.4	20	2.4	20	2.4
Total		51.0		52.9		53.4
Working Capital	20	2.1		2.1		2.1
Credit for Sulfur	320 T	(3.2)		(3.2)		(3.2)
Total Required Revenue		165.8		170.5		172.5
Production, MMBtu/d	156,720		126,828		124,350	
Required Selling Price $/MMBtu		3.20		4.07		4.2

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Table 16-2 - Syngas vs. Hydrogen as Dissolver Feed
Case A (LPG+SNG)
EUAC at 12% DCF

	Hydrogen		Syngas	
		$MM/yr		$MM/yr
Costs				
Coal	10,000 TPSD	39.6	10,825 TPSD	42.9
Catalysts and Chemicals		4.7		4.5
Operating Labor		3.4		3.4
Maintenance Labor		3.8		4.0
Payroll Burden		2.5		2.6
Plant Overhead		5.8		6.0
Maintenance Materials		7.8		8.2
Utilities		Incl		Incl
Property Tax and Insurance		8.9		9.4
G&A Overhead		1.2		1.2
Total		77.7		82.2
Income Tax		41.0		43.1
Investment, $MM				
Fixed Capital	323	50.2	341.3	53.1
Initial Catalysts and Chemicals	2.6	0.3	1.6	0.2
Startup Costs	20	2.4	20	2.4
Total		52.9		55.7
Working Capital	20	2.1	21	2.2
Credit for Sulfur	320	(3.2)	320	(3.2)
Required Revenue		170.5		180.0
Production, MMBtu/d	127,800		138,000	
Required Selling Price, $/MMBtu		4.1		3.96
Syngas Case		3.96		
Savings with Syngas		$ 0.14		

Table 16-3 - Alternate H_2S Removal Processes
EUAC to Achieve a 12% DCF After
Tax Return on Investment

			Physical Solvent					
			Process No. 1		Process No. 2		Chemical Absorption	
	Units	Unit Cost	Quantity per Year	EUAC $MM/yr	Quantity	EUAC $MM/yr	Quantity	EUAC $MM/yr
Catalysts and Chemicals								
Methanol	Mlb	$65.00			198	0.013		
Benfield Solution								0.004
Utilities								
Steam	Mlb	$ 3.20		-		-	573,400	1.835
Cooling Water	Mgal	$ 0.10	475,200	0.048	370,000	0.037	-	-
Power	kW	$ 0.03	19,100,000	0.57	6,400	0.192	1,300	0.037
Nitrogen $9/T	MSCF	$ 0.33	3.564×10^6	1.188				
Total				1.810		0.229		1.372
Capital Associated								
Fixed Capital Investment	$MM		2.1	0.281	2.36	0.316	3.9	0.523
Working Capital	$MM		0.43	0.046		0.017	0.99	0.104
Maintenance Labor				0.025		0.028		0.046
Payroll Burden				0.009		0.010		0.016
Maintenance Material				0.051		0.057		0.094
Plant Overhead				0.020		0.023		0.037
Property Tax and Insurance				0.058		0.065		0.107
Total				0.490		0.516		0.972
G&A Overhead				0.029		0.006		0.068
Income Tax				0.240		0.233		0.467
Total				2.569		0.997		3.383
Less Chemical Absorption Cost				3.383		3.383		
Savings				0.814		2.386		

Table 16-4 - Comparison of Sweet and Sour Shift Economics
Savings in EUAC With 12% DCF

Utilities		EUAC Savings $ Millions/yr
Fuel Gas	762,700 MMBtu/yr at 3.20/MMBtu	2.440
Power	36,000,000 kW/yr at $0.025/kWh	(1.080)
Steam	1,228,300 Mlb/yr at $3.20	3.930
Total		5.290
Capital Associated		
Fixed Capital Investment - $2.23 MM		0.347
Working Capital		0.065
Maintenance Material		0.054
Maintenance Labor		0.026
Payroll Burden		0.009
Plant Overhead		0.021
Property Tax and Insurance		0.061
G&A Overhead		0.050
Income Tax		0.325
Total Savings		$6.248
Sour Shift Savings in $/MMBtu based on 157,000 MMBtu/d		$0.121

Table 16-5 - Filter Cake Washing, Material and Utility Balance

	Without Filter[a] Cake Washing	With Filter Cake Washing
Coal Feed		
To Dissolvers	10,000 T/d	10,000 T/d
To Gasifier	-	1,667 T/d
Plant Products (after supplying plant fuel)		
Naphtha	270 T/d 2,000 B/d 10,600 MMBtu/d	245 T/d 1,800 B/d 9,600 MMBtu/d
Fuel Oil	1,440 T/d 8,500 B/d 48,800 MMBtu/d	1,660 T/d 9,800 B/d 56,300 MMBtu/d
Heavy Liquid	2,915 T/d 14,300 B/d 96,000 MMBty/d	3,850 T/d 19,000 B/d 126,800 MMBtu/d
Plant Fuel Required		
Fuel Gas	2,140 T/d	2,140 T/d
Heavy Liquid	120 T/d	340 T/d

[a]Refer to R&D Report No. 82

Table 16-6 - Filter Cake Solvent Recovery: Economic Evaluation
EUAC at 12% DCF

	EUAC $Million/Yr	EUAC $/MMBtu
Raw Material Coal at $12.00/T	6.600	
Capital Associated Costs		
Fixed Capital Investment ($11.93 million)	1.854	
Working Capital	0.122	
Maintenance (at 8% of FCI)		
Labor	0.283	
Payroll Burden	0.099	
Plant Overhead	0.229	
Materials	0.573	
Property Tax & Insurance	0.328	
	3.488	
G&A Overhead	0.086	
Income Tax	1.495	
Total Additional Revenue Required	$11.670	
Base Case at 156,700 MMBtu/d	165.844	3.207
With Solvent Recovery 194,000 MMBtu/d	177.514	2.773
Savings in $/MMBtu		0.434