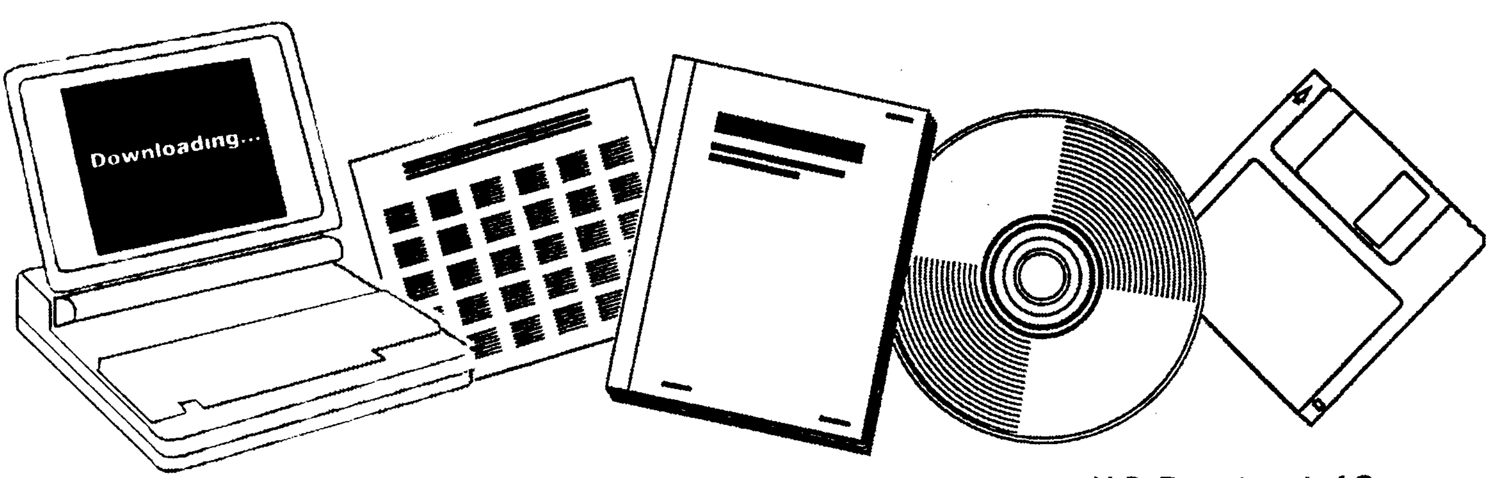
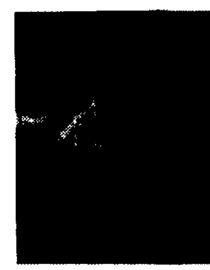
BASELINE DESIGN/ECONOMICS FOR ADVANCED FISCHER-TROPSCH TECHNOLOGY. AUARTERLY REPORT, JULY--SEPTEMBER 1992

BECHTEL NATIONAL, INC.
RICHLAND, WA

1992





DE95011372

U.S. Department of Energy
Pittsburgh Energy Technology Center

Baseline Design/Economics for Advanced Fischer-Tropsch Technology

Contract No. DE-AC22-91PC90027

Quarterly Report

July – September 1992

AMOCO

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DOE/PC/90027--T5

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Introduction and Summary

Section 1

Introduction and Summary

This report is Bechtel's fourth quarterly technical progress report and covers the period of July through September, 1992.

1.1 Introduction

Bechtel, with Amoco as the main subcontractor, initiated a study on September 26, 1991, for the U.S. Department of Energy's (DOE's) Pittsburgh Energy Technology Center (PETC) to develop a computer model and baseline design for advanced Fischer-Tropsch (F-T) technology. This 24-month study, with an approved budget of $2.3 million, is being performed under DOE Contract Number DE-AC22-91PC90027.

The objectives of the study are to:

- Develop a baseline design for indirect liquefaction using advanced F-T technology.
- Prepare the capital and operating costs for the baseline design.
- Develop a process flowsheet simulation (PFS) model.

The baseline design, the economic analysis and computer model will be major research planning tools that PETC will use to plan, guide and evaluate its ongoing and future research and commercialisation programs relating to indirect coal liquefaction for the manufacture of synthetic liquid fuels from coal.

The study has been divided into seven major tasks:

- Task 1: Establish the baseline design and alternatives.
- Task 2: Evaluate baseline economics.
- Task 3: Develop engineering design criteria.
- Task 4: Develop a process flowsheet simulation (PFS) model.
- Task 5: Perform sensitivity studies using the PFS model.
- Task 6: Document the PFS model and develop a DOE training session on its use.
- Task 7: Perform project management, technical coordination and other miscellaneous support functions.

1.2 Summary

During the reporting period work progressed on Tasks 1, 4 and 7. This report covers work done during the period and consists of five sections:

- Introduction and Summary.
- Preliminary Design for Syngas Production (Section 100) - Task 1.
- Preliminary F-T Reaction Loop Design (Section 200) - Task 1.
- Development of a Process Simulation Model - Task 4.
- Key Personnel Staffing Report - Task 7.

Under Task 1, preliminary process design information was reported in the Third Quarterly Report for the following plants (the number of on-stream trains is shown for each plant):

Plant 101	Coal Receiving and Storage (1 train).
Plant 102	Coal Drying and Grinding (5 trains).
Plant 103	Shell Coal Gasification (8 trains).
Plant 104	COS/HCN Hydrolysis (8 trains).
Plant 106	Acid Gas Removal (4 trains).
Plant 108	Sulfur Polishing (8 trains).
Plant 109	Syngas Wet Scrubbing (8 trains).

Designs are now provided for three additional plants in Area 100 - Syngas Production - and two plants in Area 200 - F-T Reaction Loop. These are as follows:

Plant 105	Sour Water Stripping (1 train).
Plant 107	Sulfur Plant (Claus/TGT) (2 trains).
Plant 110	Air Separation (8 trains).
Plant 202	CO_2 Removal (8 trains).

Plant 206 Autothermal Reforming (4 trains).

For Plants 105 and 107, material balances and preliminary information on utilities are provided. For Plant 110, a balance for eight 2100 tpd oxygen plants is given. This is about the right capacity but the autothermal reforming requirement still needs to be firmly established. For the Plants in Area 200, the material balance information will be finalized in the next quarterly report.

Under Task 4, preliminary block simulation models have been developed and are discussed for all plants in Area 100 except for the sulfur plant (Plant 107). An overall Area 100 simulation in ASPEN/SP also has been developed and a block diagram is provided showing how the different blocks interact.

Under Task 7, a considerable amount of effort went into preparations for a Progress Review Meeting at PETC headquarters on September 21, for the PETC Workshop on Slurry Reactor Hydrodynamics on September 22, where Joe Fox and Sam Tam made a slide presentation, and for the PETC Contractor's Meeting on September 23 and 24th, where Sam Tam presented a paper summarizing progress on the baseline study.

Preliminary Design for Syngas Production

Section 2

Preliminary Design for Syngas Production

The process units within the syngas production plant are grouped in Area 100. They are:

Plant Number	Plant Description
101	Coal Receiving and Storage
102	Coal Drying and Grinding
103	Shell Coal Gasification
104	COS/HCN Hydrolysis
105	Sour Water Stripping
106	Acid Gas Removal
107	Sulfur Plant (Claus/TGT)
108	Sulfur Polishing
109	Syngas Wet Scrubbing
110	Air Separation

The block flow diagram of these process plants is shown in Figure 2-1. Process designs for Plants 101, 102, 103, 104, 106, 108, and 109 were reported in the Third Quarterly Report. Details for the remaining plants, Plants 105, 107, and 110, have now been completed and are included in this section.

The design is based on feeding Illinois #6 coal to the gasifiers. The typical coal analysis is given in Table 2-1.

2.1 SOUR WATER STRIPPER (PLANT 105)

The water produced in Plant 109 is stripped in Plant 105. The waste water is sent to the waste water treatment area and the stripped gas is fed to the sulfur plant.

2.1.1 Design Basis and Considerations

This plant employs one operating train and no spare train is provided. The design capacity of this plant is 140 gpm. A sour water feed surge tank is provided which is designed for 5 days (24,000 barrels) storage capacity to allow continued operation of the rest of the facility in case of an unscheduled shutdown of the stripper.

2.1.2 Process Description and Process Flow Diagram (105-B-01)

Process flow diagram 105-B-01 shows the sour water treating plant. The combined flow of scrubber bottoms from the eight Syngas Wet Scrubbing Plants (Plant 109) enters the feed surge tank and is flashed to 2 psig. The sour water is first preheated in the feed/effluent exchanger, 105E-1, then fed to the top of the stripper, 105C-1, which is reboiled using 150 psig steam. The overhead vapor from the sour water stripper, containing principally H_2S, CO_2, and H_2O with small amount of NH_3 and HCN, is cooled in the overhead condenser, 105E-4, where most of the water is condensed. The condensate from this cooling step is separated in the overhead accumulator, 105C-2, and refluxed to the stripper. The net overhead gas from the sour water stripper is sent to the Claus unit in the sulfur recovery plant, Plant 107.

The stripped water from the bottom of the sour water stripper is first cooled in the feed/effluent exchanger, 105E-1, then cooled to 100°F against the cooling water in the stripper bottoms cooler, 105E-2. The stripped and cooled water is pumped to the plant waste water treatment facility for a final treatment before discharge or reuse.

2.1.3 Material Balance and Utilities

The material balance around Plant 105, Sour Water Treating, is shown in Table 2-2. The plant will consume 9,333 lbs/hr of 150 psig saturated steam, which is the primary utility requirement.

2.2 SULFUR RECOVERY AND TAILGAS TREATING (PLANT 107)

This plant is designed to receive the sour gas streams generated in the Coal Gasification/Syngas Cleanup section (Process Area 100) and convert the H_2S to elemental sulfur and the NH_3 to nitrogen. The treated offgas is vented to atmosphere via a stack after incineration. The product sulfur is stored and shipped out as prills.

2.2.1 Design Basis and Considerations

This plant is designed with spare capacity so that continuous operation of the facility is allowed without violating the sulfur emission requirements due to an unscheduled shutdown. Total design capacity is 150% of normal operating capacity in order to achieve high on-stream reliability with minimum capital investment cost. Thus the plant is configured with three (3) parallel trains each having 50% of the normal total plant capacity. Two trains will be in operating mode and the third on stand-by during normal operations.

Each train consists of three sections: a three-stage Claus unit, a SCOT tail gas treating unit, and a catalytic incinerator with a stack. The Claus/SCOT combination was selected because it is a proven combination and is becoming an industry standard. This plant is designed to meet the environmental compliance strategy as outlined previously in the in Second Quarterly Report - June 1992, Section 3.2.3.

The three stage Claus unit is designed to recover a minimum of 95% of the incoming sulfur on a combined feed basis. This reduces the recycle from the tail gas treating unit to an economic level. Unconverted sulfur in the gas from the Claus unit is sent to the SCOT tail gas treating unit where the sulfur is converted to H_2S and recycled back to the Claus unit. The overall sulfur recovery for the combined Claus/SCOT system is 99.8%. Sulfur production is 560 tpd corresponding to 68.2 pounds of sulfur per ton of coal gasified (MAF), or 94.4% of the sulfur in the coal. The stack gas has an SO_2 concentration of 0.025 vol% on a dry basis to meet the emission requirement at the Illinois site.

2.2.2 Process Description and Process Block Flow Diagram

A process block flow diagram of the Claus/SCOT units is shown in Diagram 107-B-01, Sheets 1 and 2. The major stream flows are shown in Table 2-3.

Acid gas from the AGR stripper, sour gas from the sour water stripper, and NH_3 and H_2S from the ammonia stripper are sent to the sulfur recovery plant (Sheet 1), where the gases are partially combusted to produce a stoichiometric quantity of SO_2, and then reacted to produce liquid elemental sulfur. About one third of the H_2S is initially combusted with air, which has been preheated to 450 °F, in order to achieve a temperature of 2200-2400 °F. Sufficient air is added to produce sufficient SO_2 via thermal reaction to satisfy the stoichiometric ratio of 2:1 for H_2S/SO_2 required to form sulfur and water.

$$2H_2S + SO_2 = 3S + 2H_2O$$

Over half of the conversion to sulfur occurs in the thermal reactor at about 2200-2400 °F, the remainder in three catalytic converters at about 450 °F. Any ammonia present reacts to form molecular nitrogen and water. NO_X formation is minimized by the use of low NO_X burners and stage-wise addition of air. The reaction products from the thermal reactor are cooled in a boiler generating 600 psig steam and then cooled further in the primary sulfur condenser, a three section boiler producing low pressure (50 psig) steam, where molten sulfur is condensed out between stages. Part of the 600 psig steam is used to reheat the uncondensed vapors back to 450 °F before going to the next catalytic reactor stage. Sulfur in the outlet vapor from the first two catalytic stages is condensed in the remaining two sections of the primary condenser. Sulfur from the last stage catalytic converter is condensed against cooling water in order to achieve maximum sulfur recovery.

Molten sulfur flows to an underground storage pit where it is kept heated with low pressure steam prior to prilling. Prilled sulfur is stored for shipment

Tail gas from the Claus unit is sent to a SCOT unit (Sheet 2), where unreacted S, SO_2 and COS are converted/reduced to H_2S in a combination reactor consisting of a fired combustion chamber and a catalytic reaction vessel. The tail gas is preheated using 600 psig steam minimizing the need for fuel gas addition to the combustion chamber. A slip stream of syngas provides the fuel gas and the excess fuel gas provides the necessary reducing gas. The reactor effluent is first used to generate 50 psig steam then further cooled in a quench column by direct contact with circulating water. The bottoms from this tower is sent to wastewater treating. The quench column offgas enters an amine absorber, where essentially all of the H_2S from the quenched gas stream is removed. Rich amine solvent leaving the absorber is sent to a stripper. The stripper offgas is recycled to the Claus unit for recovery of sulfur.

2.2.3 Technology/Vendor Selection

The Claus sulfur process is established commercially and, consequently, the equipment requirements are well known. The Shell Claus Offgas Treating (SCOT) process is selected because of its simplicity, cost advantage, and proven track record.

2.2.4 Material Balance and Utilities

The overall material balance is shown in Table 2-3.

The sulfur recovery plant will have 91,000 lbs./hr of excess 600 psig steam for export. It will also have a net consumption of 169,000 lbs/hr of 50 psig steam. These are the major utility requirements. Other utilities will be developed as the study progresses.

2.3 AIR SEPARATION (PLANT 110)

This plant primarily provides the required oxygen for the gasification of coal. In addition, a small quantity of oxygen also is provided for the autothermal reforming operation in the F-T synthesis loop.

2.3.1 Design Basis and Considerations

A total of eight operating trains, each at 2,100 STPD, will be used in the design. No spare train will be provided. The design incorporates a backup system including a liquid oxygen storage capacity of 16,800 tons, which is equivalent to eight days of single train production, and a gaseous oxygen storage of 58 tons per train, which is equivalent to 40-minutes of production from one train. This will protect the facility from an unscheduled shutdown of one or more oxygen trains.

The purity of oxygen will directly impact the size of the F-T recycle loop. The use of high purity oxygen will minimize the N_2 content in the syngas and, consequently, the N_2 built-up in the recycle loop. This will, in turn, minimize the flow of recycle gas for a given purge rate. A trade-off study was performed to examine the capital and operating costs of the process units in the F-T recycle loop as well as those for the air separation plant at two oxygen purity levels, 95 versus 99.5 mole percent. The results indicate that the savings on capital and operating costs for the affected process plants are far greater than the increase in oxygen plant costs as purity level increases. An oxygen plant producing 99.5% O_2 purity is used for the baseline design.

2.3.2 Process Description

A simplified flow diagram of one air separation train is shown in Diagram 110-B-01.

Ambient air is filtered and compressed in a two-stage axial centrifugal compressor with interstage cooling The air from the final stage of compression enters a direct contact aftercooler where it contacts cooling water and chilled cooling water in two separate packed sections.

The cooled air from the top of the aftercooler has lost the majority of its ambient water vapor. Removal of the residual water vapor, carbon dioxide and other atmospheric contaminants occurs in the molecular sieve adsorbers. Two vessels containing the adsorbent are used in a cyclic process. While one vessel is on line purifying the incoming air, the other vessel first is heated with dry waste nitrogen gas to remove the adsorbed contaminants and then cooled to the operating temperature before being placed back in service. The regeneration heater uses high pressure steam to raise the temperature of the regeneration gas to the proper level.

The dry air enters the "cold box" where it is cooled to cryogenic temperature in the main heat exchangers and is separated into oxygen and nitrogen by cryogenic distillation. Final cooling is by expansion. The oxygen stream is further purified in a argon column to 99.5 mole %. The main heat exchangers are brazed aluminum, multipass, plate-fin units in which air is cooled against the cold oxygen and nitrogen streams leaving the distillation columns.

The oxygen product stream leaving the cryogenic separation section is warmed in the main heat exchangers and compressed to final delivery pressure in a centrifugal compressor. In order to insure a continuous supply of oxygen, backup storage systems are included in the design. A portion of the oxygen produced in the distillation is withdrawn as liquid and stored at low pressure in an insulated tank. In addition to liquid oxygen storage, a high pressure gaseous oxygen storage tank also is provided. This tank is periodically filled by pumping liquid oxygen to over 2000 psia, vaporizing the liquid and storing it at pressure at ambient temperature. Oxygen from the gaseous storage tank is always available to meet a shortfall in supply pressure regardless of the cause. Sufficient gaseous storage capacity is provided to allow time to cool the main liquid oxygen supply pumps to operating temperature. Oxygen supply is then provided by pumping liquid to supply pressure and vaporizing the liquid with steam.

2.3.3 Technology/Vendor Selection

There are at least four major vendors who can supply turnkey air separation plants of this size: Liquid Air Engineering, Airco, Air Products, and Union Carbide. Each design has a slightly different power requirement and capital cost. The Air Products design is selected for the baseline study.

2.3.4 Material Balance

The detailed material balance for eight 2100 tpd oxygen plants, supplying eight Shell gasifiers plus the estimated autothermal reformer requirement, is shown in Table 2-4. The autothermal reforming requirement is not yet fully defined and may change the total oxygen demand slightly. Eight standard oxygen plants should still suffice.

Table 2-1
Gasifier Feed Coal Analysis
(Illinois No. 6 Coal)

Item	Gasifier Feed Coal	Dry
Higher Heating Value, Btu/lb (measured)	11,193	12,246
Proximate Analysis, wt %		
Moisture	2.00	–
Ash	11.26	11.49
Volatile Matter	41.39	42.23
Fixed Carbon	45.35	46.28
Ultimate Analysis, wt %		
Moisture	2.00	–
Ash	11.26	11.49
Carbon	69.59	71.01
Hydrogen	4.70	4.80
Nitrogen	1.37	1.40
Sulfur	3.13	3.19
Chlorine	0.10	0.10
Oxygen (by difference)	7.85	8.01

Table 2-2
Total Plant Material Balance
Plant 105- Sour Water Stripping
(Illinois No. 6 Coal)

Stream No.	105.1 Sour Water from Wet Scrubbing	105.2 Scur Gas to Sulfur Recovery	105.3 Waste Water to Treating	105.4 Acid Makeup
Phase	Liquid	Vapor	Liquid	Liquid
Flow Rate, lbmol/hr				
H2	0.36	0.36		
N2				
O2				
H2S	18.14	18.11	0.03	
CO	0.01	0.01		
CO2	13.12	13.12		
H2O (or Moisture)	3704.88	28.24	3927.73	251.09
COS				
NH3	0.76	0.56	0.20	
HCl				
HCN	1.06	0.47	0.58	
CH4				
Coal MAF				
Ash	(73 lb/hr)		(73 lb/hr)	
C	(18 lb/hr)		(18 lb/hr)	
Cl				
Sulfur	(1 lb/hr)		(1 lb/hr)	
NaOH	30.76		30.76	
NaCl	34.39		34.39	
H2SO4			15.38	15.38
Total Lbmol/hr	3803.48	60.86	4009.08	266.47
Total Lb/hr	71,296	1,726	75,601	6032

Table 2-3
Total Plant Material Balance
Plant 107- Sulfur Plant
(Illinois No. 6 Coal)

Stream No.	107.1 NH3 Gas	107.2 Sour Gas from SWS	107.3 Acid Gas from AGR	Combined Feed Gas	107.4 Treated Tail Gas	107.5 Sulfur Product	107.6 Combust -ion Air	107.7 Waste Water	107.8 Reduc- ing Gas
Phase	Vap	Vap	Vap	Vap	Vap	Liq	Vap	Liq	Vap
Flow Rate, lbmol/hr									
H2	1.31	0.36	34.01	35.68	35.68				70.00
N2			0.66	0.66	3184.87		3157.14		
O2							836.51		
H2S	13.88	18.11	1425.76	1457.75	1.46				
CO	2.79	0.01	61.91	64.71	0				
CO2	48.77	13.12	1750.15	1812.04	1877.54		100.24		
H2O (or Moisture)	87.77	28.24	146.58	262.59				1969.55	
COS									
NH3	52.53	0.56	0.25	53.34					
HCl									
HCN	0.33	0.47		0.80					
CH4									
Coal MAF									
Ash									
C									
Cl									
Sulfur						1456.30			
NaOH									
NaCl									
H2SO4									
Total Lbmol/hr	207.38	60.87	3419.32	3687.57	5099.55	1456.30	4093.89	1969.55	70.00
Total Lb/hr	5,185	1,726	130,074	136,985	171,971	46,689	117,124	35,482	141

Table 2-4
Total Plant Material Balance
Plant 110 - Air Separation Unit
(Illinois No. 6 Coal)

Stream No.	110.1 Air Dry Basis	110.2 Oxygen Product	110.3 Waste Nitrogen	Loss
Phase	Vap	Vap	Vap	Vap
Flow Rate, lbmol/hr				
N2	166762.17		166604.92	157.25
O2	44739.07	43749.89	946.85	42.34
Ar	1990.46	219.74	1768.70	2.02
H2O (or Moisture)				
Total, Lbmol/hr	218956.40	43969.63	174785.17	201.61
Total, Lb/hr	6,182,687	1,408,722	4,768,124	5,841

NOTE: BASIS EIGHT 2100 SHORT TPD OXYGEN TRAINS

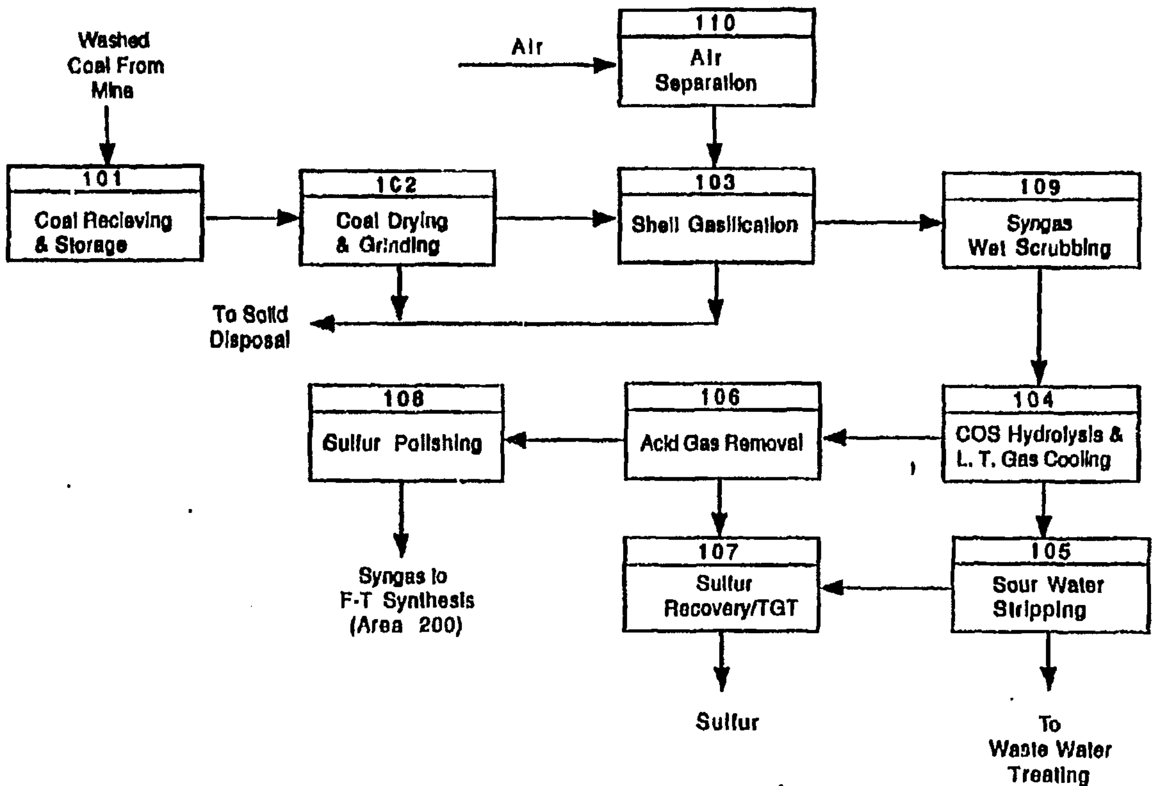


Figure 2-1 Block Flow Diagram for Area 100 - Syngas Production

105D-1 SOUR WATER SURGE TANK

105G-2A/B TREATED WATER PUMP

105E-2 STRIPPER BOTTOM COOLER

105C-1 SOUR WATER STRIPPER

105E-3 STRIPPER REBOILER

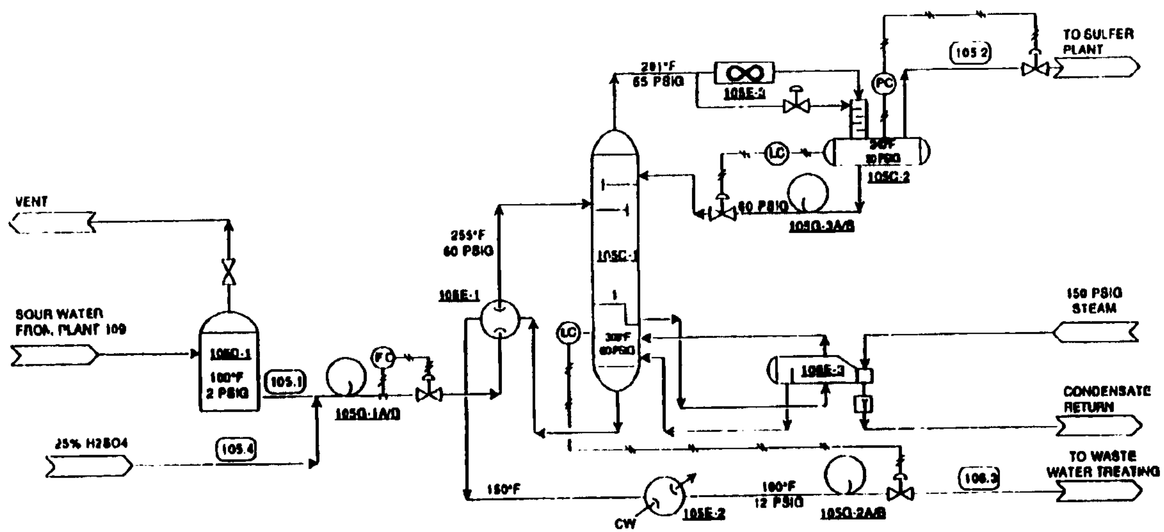
105E-4 STRIPPER OVHD COND.

105C-2 STRIPPER OVHD ACCUMULATOR

105G-3A/B REFLUX PUMPS

105G-1A/B SOUR WATER PUMP

105E-1 FEED/EFFLUENT EXCHANGER



NOTES

1. One train for the whole plant.
2. MM = 1,000,000 BTU/HR
3. GPM shown are at flow conditions.
4. Air Cooling Design temperature is 96°F in.
5. Cooling water design temperatures 87°F in; 115°F out
6. CW = cooling water

THE PROCESS CONDITIONS OF FLOW QUANTITIES, TEMPERATURES, PRESSURES, COMPOSITIONS AND INSPECTIONS AND EQUIPMENT DUTIES SHOWN ON THIS FLOW DIAGRAM ARE FOR DESIGN PURPOSES ONLY, AND WHILE USEFUL AS GUIDES IN OPERATION DO NOT REPRESENT EXACT OR GUARANTEED OPERATING CONDITIONS

GUARANTEES ARE SPELLED OUT ONLY IN THE CONTRACT

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Process Flow Diagram
Plant 105 - Sour Water Stripping
Illinois No. 6 Coal

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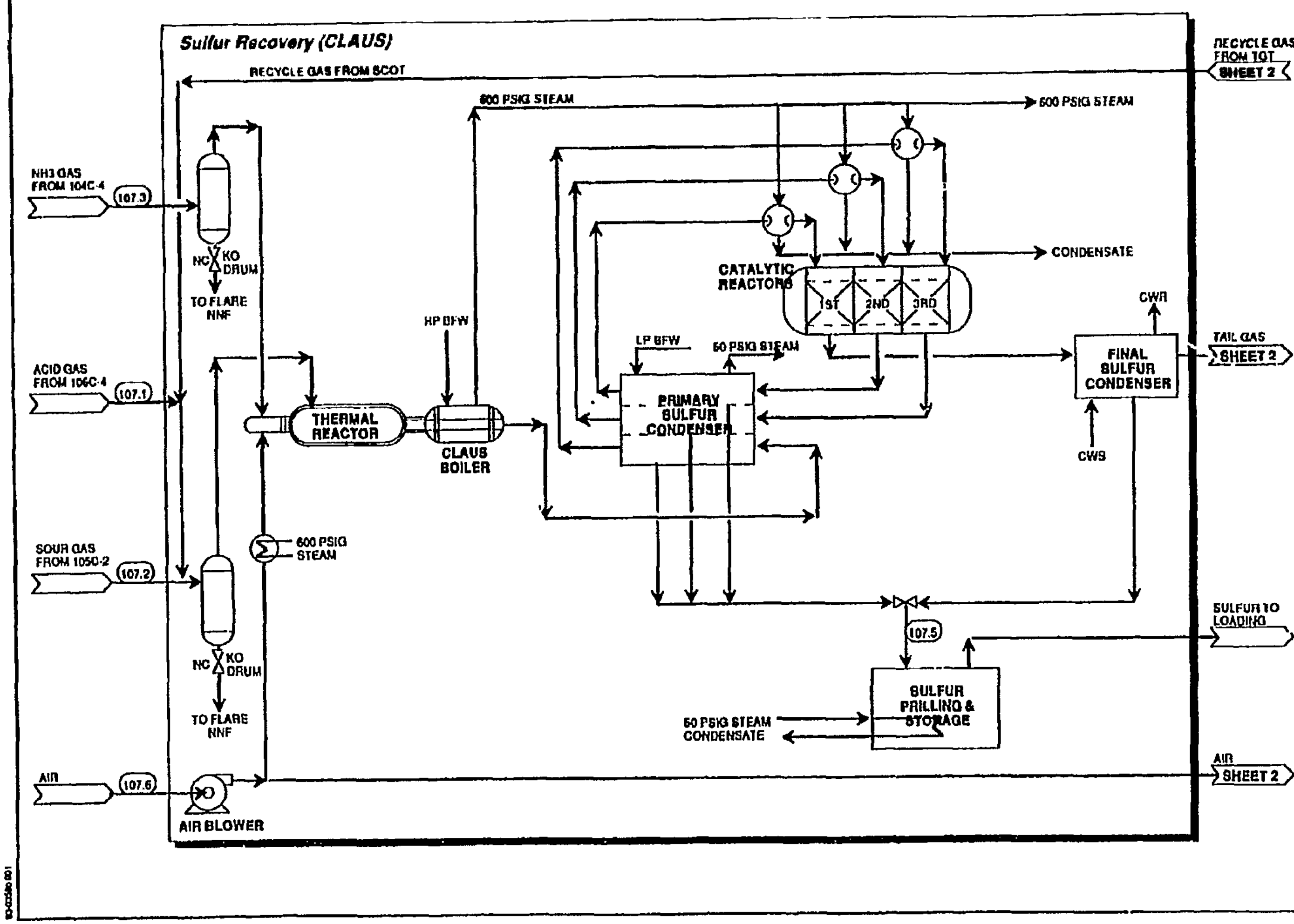

Notes

1. Process schematic is for one train only.
2. Total of three (3) trains are required for the whole plant.

Train	Design Capacity Long Ton/Day	Status
1	250	operating
2	250	operating
3	250	spare

3. CWS = Cooling Water Supply
   CWR = Cooling Water Return
4. NC = Normally Closed
   NNF = Normally No Flow
6. Cooling water design tempoartures
   87°F In
   115°F Out

REV	DATE	ISSUE DATE	BY	PROJ ENG	ENG SUPV	PROC	PROJ	MGR
		Review						
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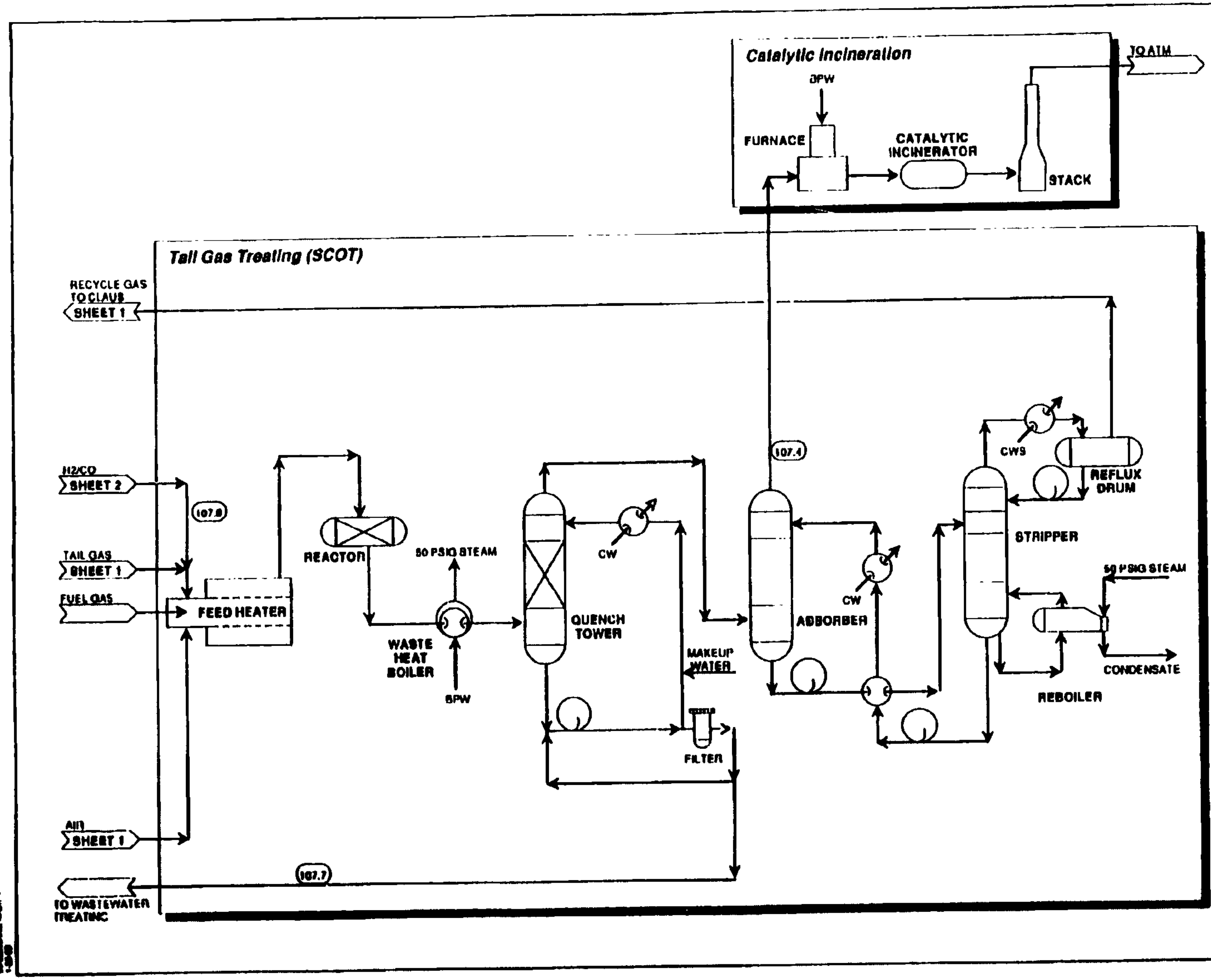
Simplified Process Flow Diagram
Plant 107-Sulfur Recovery and Tailgas Treating
Illinois No. 6 Coal

JOB No.	DRAWING No.	REV
21655	107-B-01 SHEET 1 OF 2	A



Notes

1. Process schematic is for one train only.
2. Total of three (3) trains are required for the whole plant.

Train	Design Capacity Long Ton/Day	Status
1	250	operating
2	250	operating
3	250	spare

3. CWS = Cooling Water Supply
   CWR = Cooling Water Return
4. Cooling water design temperatures:
   87°F In
   115°F Out

THE PROCESS CONDITIONS OF FLOW QUANTITIES, TEMPERATURES, PRESSURES, COMPOSITIONS AND INSPECTIONS, AND EQUIPMENT DUTIES SHOWN ON THIS FLOW DIAGRAM ARE FOR DESIGN PURPOSES ONLY, AND WHILE USEFUL AS GUIDES IN OPERATION, DO NOT REPRESENT EXACT OR GUARANTEED OPERATING CONDITIONS.

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REV	DATE	ISSUE DATE
	1/12/93	4th Qtrly
	10/12/92	Review

SCALE None | DRAWN MSA

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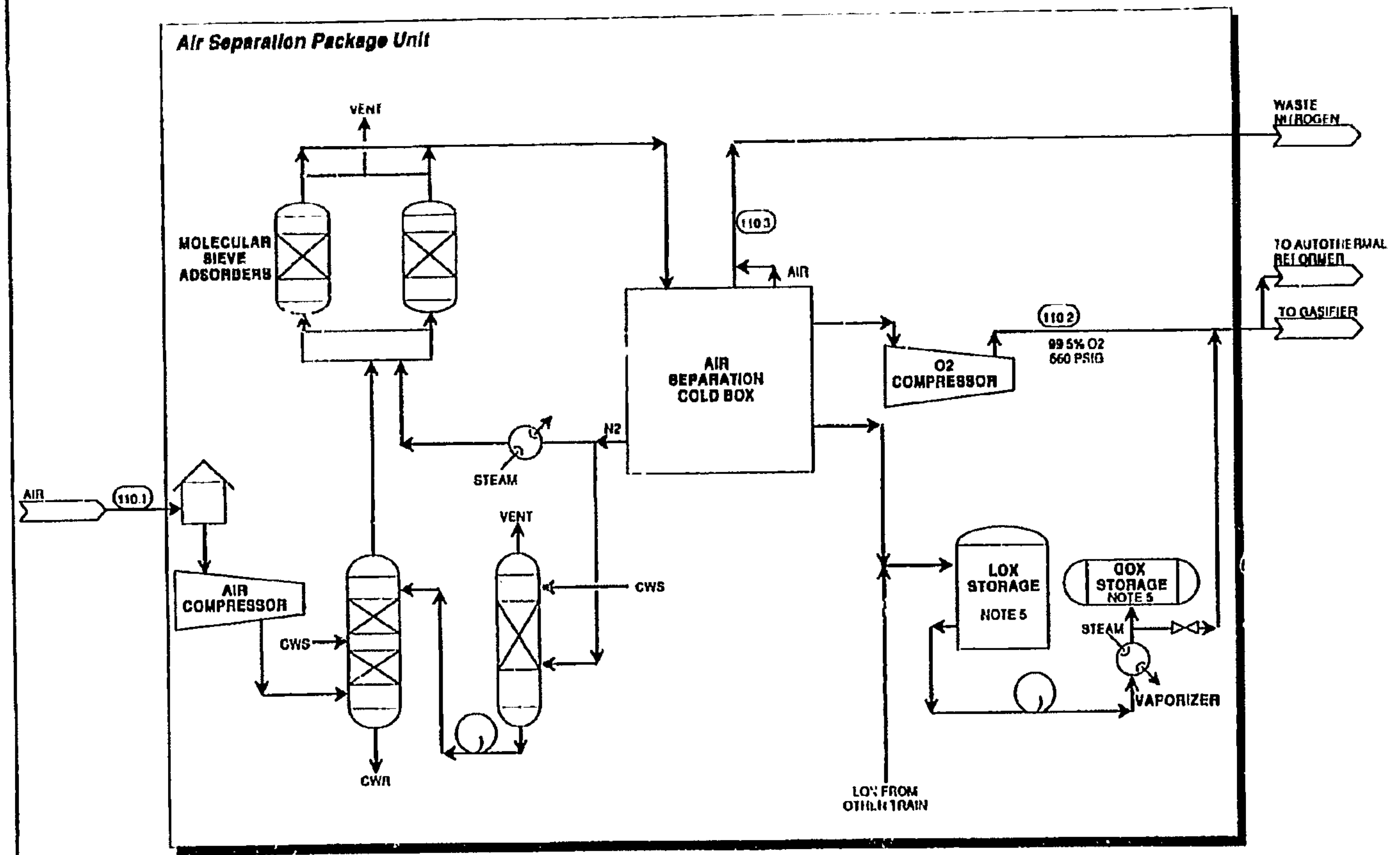
Simplified Process Flow Diagram
Plant 107-Sulfur Recovery and Tailgas Treating
Illinois No. 6 Coal

JOB No.	DRAWING No.	REV
21666	107-B-01 SHEET 2 OF 2	A



Notes

1. Block flow diagram is for one train only.
2. Total of eight (8) trains are required for the whole plant.
3. CWS = Cooling Water Supply
   CWR = Cooling Water Return
4. LOX = Liquid Oxygen
   GOX = Gaseous Oxygen
5. LOX and GOX storage facility is shared by 2 trains.
6. Cooling water design temperatures:
   87°F in
   115°F out

THE PROCESS CONDITIONS OF FLOW QUANTITIES, TEMPERATURES, PRESSURES, COMPOSITIONS AND INSPECTIONS, AND EQUIPMENT DUTIES SHOWN ON THIS FLOW DIAGRAM ARE FOR DESIGN PURPOSES ONLY, AND WHILE USEFUL AS GUIDES IN OPERATION, DO NOT REPRESENT EXACT OR GUARANTEED OPERATING CONDITIONS.

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REV	DATE	ISSUE DATE	BY	PROJ ENG	ENG SUPV	PROJ	PROJ	UNIT
	1/13/93	Quarterly						
	10/13/92	Review						

SCALE None | DRAWN MSA

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Simplified Process Flow Diagram
Plant 110-Air Separation Unit
Illinois No. 6 Coal

JOB No	DRAWING No	REV
21655	110 B-01	A

Preliminary F-T Reaction Loop Design

Section 3

Preliminary F-T Reaction Loop Design

The process units within the F-T reaction loop are designated to be in Area 200. They are:

Plant Number	Plant Description
201	F-T Synthesis
202	Carbon Dioxide Removal
203	Compression & Dehydration
204	Hydrocarbons Recovery
205	Hydrogen Recovery
206	Autothermal Reforming

The block flow diagram of these process units is shown in Figure 3-1. Process designs for Plants 202 and 206 have been completed, although the material balances and utility requirements still are preliminary. Detailed descriptions of these plants are included in this section. Material balances and utilities will be given in the next quarterly report.

3.1 CARBON DIOXIDE REMOVAL (PLANT 202)

3.1.1 Design Basis and Considerations

CO_2 is the primary byproduct in the F-T reactor when operated at low H_2/CO ratio with a catalyst active for the water gas shift reaction. It must be removed from the F-T recycle loop recycle gas before the gas is returned to the reactor and, the sooner this is done, the better since this will reduce the size of the downstream equipment. The general process requirements and preferences for the CO_2 removal unit are as follows:

(1) The CO_2 removal system should preferably operate at loop pressure before the gas is recompressed in the recycle compressor (approximately 260 psig);

(2) The CO_2 stream should preferably be produced at elevated pressure and has to be free of hydrocarbons and inert gases such as nitrogen so it can be used in the gasifier coal feed system (Plant 102) without further compression or purification;

(3) If vented to the atmosphere without additional treatment, the CO_2 stream must be low in contaminants;

(4) The recycle gas after CO_2 removal must be sufficiently low in CO_2 (e.g. less than 400 ppm) so that it will not freeze out in the cryogenic section of the hydrocarbon recovery unit in the gas plant (Plant 204) to plug up the heat exchangers and other equipment;

(5) Low cost and low energy requirement are preferred.

Chemical solvents involving amines are favored for this service because of lower capital investment cost and reasonable energy requirements compared to other technologies. Furthermore, inhibited MDEA (a tertiary amine) or glycol amines are preferred over high concentration MEA (a primary amine) with corrosion inhibitors because there is less of a problem with disposal of the spent solutions. Any of these amines will produce a recycle gas with a CO_2 concentration well below the 400 ppm specification. More detailed analysis of alternative CO_2 removal technologies was performed in the Tradeoff Studies and the results documented in the Quarterly Report for April-June 1992 for this project.

3.1.2 Technology/Licensor Selection

There are several proprietary amine solvents such as Gas/Spec (Dow Chemical), UCARSOL (Union Carbide), Amine Guard (UOP), Flexsorb (Exxon), etc. which contain high concentrations of either MDEA or glycol amines plus proprietary additives. There is not much difference between these processes from a heat requirement or cost standpoint. The major difference is in the types of the corrosion protection schemes.

Dow's Gas/Spec process, which uses a 50 wt% MDEA solution plus proprietary additives, is selected as a representative process for the process design.

3.1.3 Process Description (PFD 202-B-01)

The process flow diagram for the CO_2 removal unit (Plant 202) is shown in Drawing PFD 202-B-01. This plant is configured similarly to the Acid Gas Removal unit (Plant 106) which is also an amine type of unit. Eight (8) parallel Plant 202 trains are required.

The vapors from the F-T high pressure separator (Plant 201 F-T Synthesis) and the off-gas from the deethanizer overhead (Plant 204 F-T Gas Plant) are combined in feed gas KO drum 202D-1 and sent to the amine adsorber (202C-1) for CO_2 removal. To ensure an amine-free vapor product, the absorber overhead vapor is water-washed in scrubber 202C-3, and the liquid is returned to the rich-amine knock-out drum, 202-D2. The rich-amine solution from the bottom of adsorber is flashed in 202-D2, and heated by exchange with the lean-amine solution before it is sent to one of the two Amine Regenerators, 202C-2A or B.

Because of the high reboiler duty caused by the high CO_2 removal rate, two (2) amine regenerators are required for each amine absorber. Each regenerator is in turn serviced by four (4) reboilers. The regenerated lean-amine solution from these two regenerators is combined and pumped to a common amine storage tank 202D-3. The lean-amine is then pumped from this tank, cooled and sent back to the absorber. A portion, approximately 10%, of the cooled lean-amine solution is filtered and returned to the storage tank.

A portion of the CO_2 gas from the regenerator overhead separators (202C4A and B) is sent to Plant 102 for use as coal carrier gas in the gasifier lock hoppers. Excess CO_2 is vented to atmosphere.

3.2 AUTOTHERMAL REFORMER (PLANT 206)

3.2.1 Design Basis and Considerations

The autothermal reformer is included in the recycle loop to:

(1) Minimize the built-up of light ends by converting them to syngas. This will produce more F-T liquids and improve the overall economics.

(2) Help modify the H_2/CO ratio in the F-T feed stream to satisfy the target conversions of H_2 and CO in the F-T reactors.

(3) Provide operating flexibility such that an unexpected failure of the Gas Plant (Plant 204) does not cause the shutdown of the entire recycle loop. In this case the gas stream, containing unrecovered C_3/C_4 hydrocarbons, can be bypassed to Plant 206 for disposal. For this matter, excess plant fuel gas can also be sent to this plant to be converted to valuable F-T feedstock.

A reactor feed/effluent heat exchange system is incorporated in the design for reasons of efficiency. A waste heat boiler and feed charge heater system would cost about the same, but would be less energy efficient.

3.2.2 Technology/Vendor Selections

Autothermal Reaction System

Both Lurgi Corporation of Germany and Haldor Topsoe, Inc. of Houston, TX are licensors of autothermal reforming technology. In the baseline design, Lurgi's technology is used because of their commercial experience at Sasol.

Autothermal Reforming Catalysts

Lurgi would use catalyst supplied by vendors such as United Catalysts, Inc. Haldor Topsoe will provide their own catalyst.

Reactor Feed/Effluent Heat Exchangers

Due to the high temperature differentials (reactor effluent cooling from 1800°F to 500°F and reactor feed heating from 72°F to 1400°F), the heat exchanger metallurgy and internal mechanical design require special attention. A TEMA AFT shell-and-tube heat exchanger with four (4) shells in series and countercurrent flow would meet the thermal design criteria. However, the design of internals, such as the stress considerations for the tubesheet and shell, need first-of-its-kind calculations. While this high-temperature/high-pressure design principle has been demonstrated in ICI's Gas Heated Reformer or GHR used in hydrogen and ammonia plant designs, heat exchanger vendors do not yet fabricate such equipment for standard heat exchange services.

In order to eliminate the tubesheet and the exposure of the shell inside wall to high temperatures, Struthers Wells Corporation of Warren, PA has proposed a feed superheater to replace the first two shell-and-tube heat exchangers which see the highest temperatures. This feed superheater would be internally lined with the refractory and would be similar to a superheater box using hot furnace flue gas to superheat other process gases or steam except that this feed superheater box would be under pressure.

3.2.3 Process Description (PFD 206-B-01)

The process flow diagram of the autothermal reforming unit (Plant 206) is shown in PFD 206-B-01. The oxygen feed from the air separation unit (Plant 110) is heated to 500°F by high pressure steam in heat exchanger 206E-2. The warm oxygen is subsequently combined with the process steam before being sent to the mixing chamber located at the top of the autothermal reforming reactor, 206C-1. The preheated feed gas also is introduced to this mixer via separate nozzles. The mixing chamber and the reactor are proprietary designs of the licensors.

The normal autothermal reformer reactor outlet temperature is 1,800°F. The combined recycle gas from the hydrogen recovery unit (Plant 205) at a temperature of approximately 72°F is heat exchanged with the autothermal reformer effluent to reach a feed preheat temperature of 1706°F. The heat exchange takes place in feed superheater 206E-1 and heat exchangers 206E-3A & B. As a result, the reactor effluent is cooled to 500°F before it is sent to the F-T reactors in Plant 201. Preheat levels are maximized in order to minimize oxygen consumption. No steam is produced using this heat exchange arrangement.

Four (4) parallel reactor heat exchanger trains are provided for this service. This four train configuration is dictated by a combined consideration of turndown capability of the autothermal reformers and pressure drop limitations of the superheater and heat exchangers.

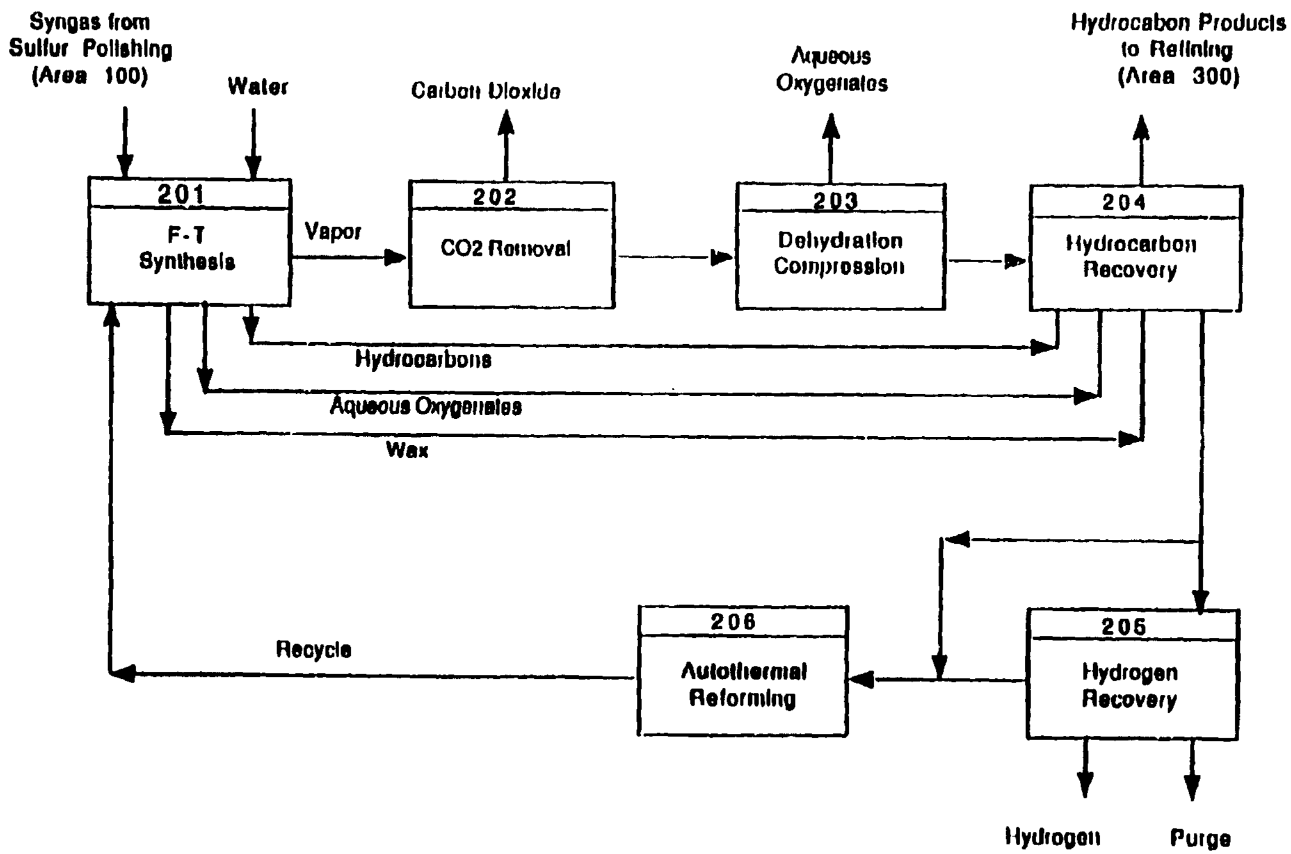
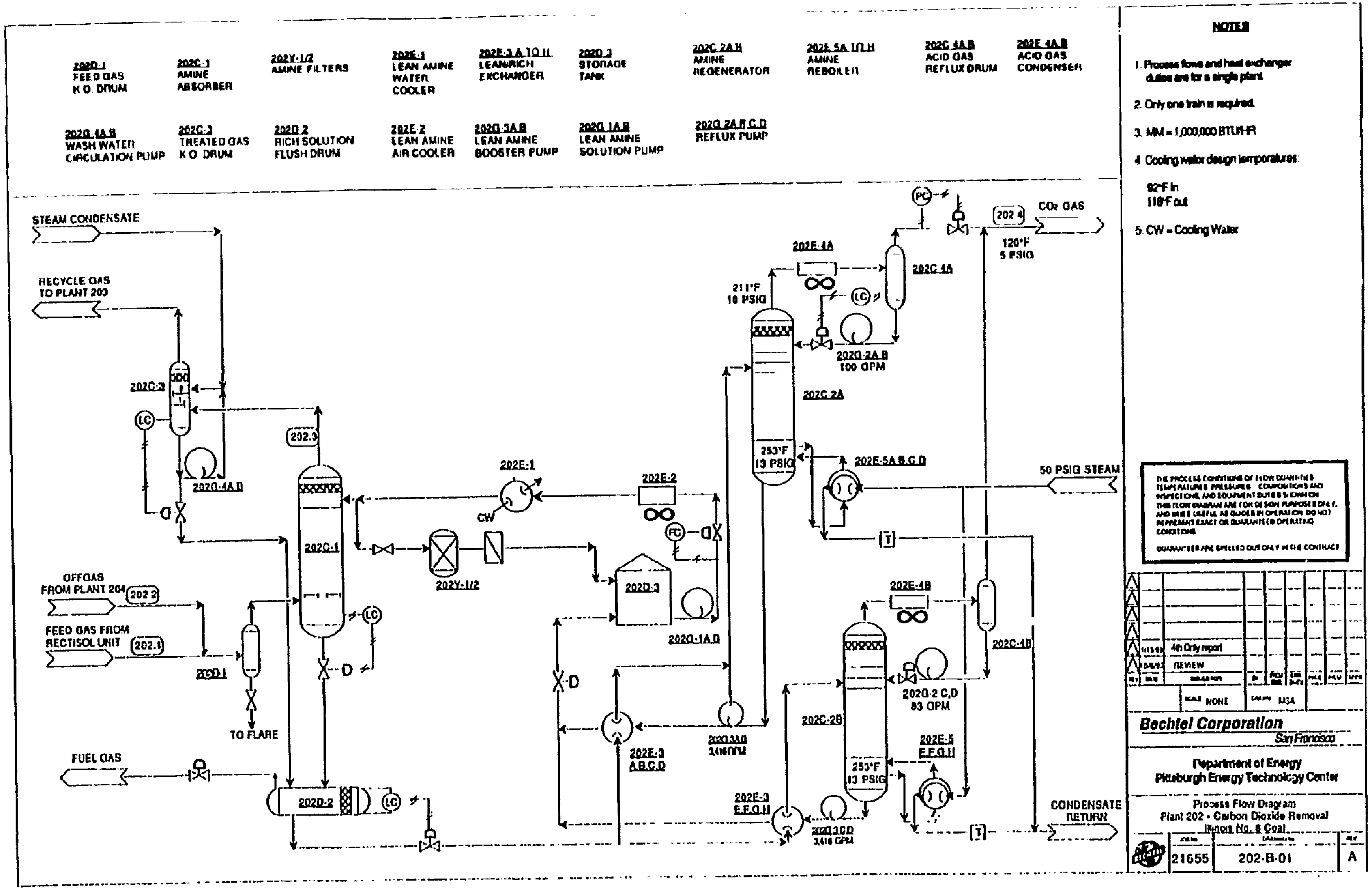


Figure 3-1
Block Flow Diagram for Area 200- F-T Reaction Loop



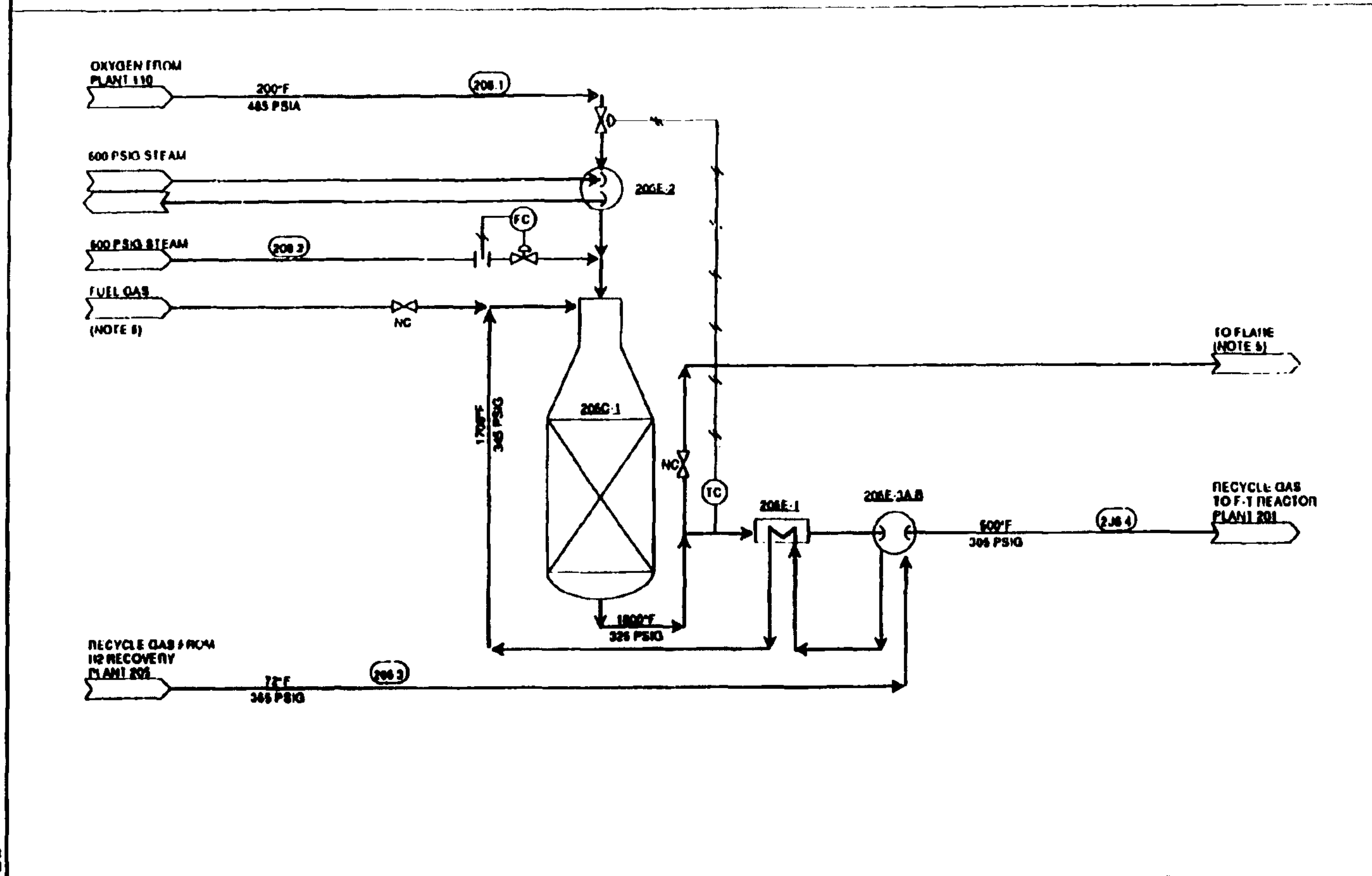
202D-1 FEED GAS K.O. DRUM
202C-1 AMINE ABSORBER
202Y-1/2 AMINE FILTERS
202E-1 LEAN AMINE WATER COOLER
202E-3 A TO H LEAN/RICH EXCHANGER
202D-3 STORAGE TANK
202C-2A,B AMINE REGENERATOR
202E-5A TO H AMINE REBOILER
202C-4A,B ACID GAS REFLUX DRUM
202E-4A,B ACID GAS CONDENSER
202G-4A,B WASH WATER CIRCULATION PUMP
202C-3 TREATED GAS K.O. DRUM
202D-2 RICH SOLUTION FLUSH DRUM
202E-2 LEAN AMINE AIR COOLER
202G-3A,B LEAN AMINE BOOSTER PUMP
202G-1A,B LEAN AMINE SOLUTION PUMP
202G-2A,B,C,D REFLUX PUMP
STEAM CONDENSATE
RECYCLE GAS TO PLANT 203
202C-3
202G-4A,B
202.3
OFFGAS FROM PLANT 204
202.2
FEED GAS FROM RECTISOL UNIT
202.1
202D-1
TO FLARE
FUEL GAS
202D-2
202C-1
202E-1
CW
202Y-1/2
202E-2
202D-3
202G-1A,B
202E-3 A,B,C,D
202G-3A,B 3,416 GPM
202E-3 E,F,G,H
202G-1C,D 3,416 GPM
202E-4A
202C-4A
211°F 10 PSIG
202G-2A,B 100 GPM
202C-2A
253°F 13 PSIG
202E-5A,B,C,D
50 PSIG STEAM
202 4
CO2 GAS
120°F 5 PSIG
202E-4B
202C-4B
202G-2 C,D 83 GPM
202C-2B
253°F 13 PSIG
202E-5 E,F,G,H
CONDENSATE RETURN
NOTES
1. Process flows and heat exchanger duties are for a single plant.
2. Only one train is required.
3. MM = 1,000,000 BTU/HR
4. Cooling water design temperatures:
92°F in
118°F out
5. CW = Cooling Water
4th Qtrly report
REVIEW
SCALE NONE
MSA
Bechtel Corporation
San Francisco
Department of Energy
Pittsburgh Energy Technology Center
Process Flow Diagram
Plant 202 - Carbon Dioxide Removal
Illinois No. 6 Coal
21655
202-B-01
A

206E-2
OXYGEN PREHEATER
0.55 MM

206C-1
AUTOTHERMAL REFORMER
7'-1" ID X 18'-0" T-T

206E-1
AUTOTHERMAL REFORMER
FEED SUPERHEATER
32.00 MM

206E-3A,B
AUTOTHERMAL REFORMER
FEED-EFFLUENT EXCHANGER
65.65 MM



Notes

1. Process flows and heat exchanger duties are for one train only.
2. Total of four (4) trains are required for the whole plant.
3. MM = 1,000,000 BTU/HR.
4. NC = Normally closed
5. Fuel gas is used for start-up only.

THE PROCESS CONDITIONS OF FLOW QUANTITIES, TEMPERATURES, PRESSURES, COMPOSITIONS AND INSPECTIONS AND EQUIPMENT DUTIES SHOWN ON THIS FLOW DIAGRAM ARE FOR DESIGN PURPOSES ONLY, AND WHILE USEFUL AS GUIDES IN OPERATION, DO NOT REPRESENT EXACT OR GUARANTEED OPERATING CONDITIONS.

GUARANTEES ARE SPELLED OUT ONLY IN THE CONTRACT.

REV	DATE	ISSUE DATA	BY					
	11/13/93	4th Qtrly						
	10/4/92	Review						

SCALE: None
DRAWN: MSA

Bechtel Corporation
San Francisco

U.S. Department of Energy
Pittsburgh Energy Technology Center

Process Flow Diagram
Plant 206-Autothermal Reformer
Illinois No. 6 Coal

JOB No.	DRAWING No.	REV
21946	206-B-01	A

Development of a Process Simulation Model

Section 4

Development of a Process Simulation Model

Amoco continued developing block models for the Area 100 syngas production section of the facility. Preliminary user models have been developed for all Area 100 section plants except for Plant 107, the sulfur recovery plant. All these plant models have been tied together in a single ASPEN/SP simulation of the syngas production section.

4.1 Coal Receiving and Storage (Plant 101)

The preliminary Plant 101 (coal receiving and storage) Fortran user block model has a single input and a single output stream. It was developed only to predict the utilities consumptions, operating labor and capital cost for the coal receiving and storage facilities.

4.2 Coal Drying and Grinding (Plant 102)

The preliminary Plant 102 (coal drying and grinding plant) Fortran user block model has a single input stream and four output streams. This model was adapted from the coal cleaning model that was developed for the direct coal liquefaction baseline design study. Since this block model included coal cleaning and drying, it can also be used for the indirect coal liquefaction model provided appropriate parameters are supplied. Thus, the four output streams are the dried (clean) coal, middling coal, refuse and waste water streams. Since the indirect liquefaction baseline design only calls for a coal drying and grinding plant, the middling and refuse streams are null streams.

4.3 Shell Gasifier (Plant 103)

The preliminary Plant 103 (Shell gasifier plant) model for Illinois No. 6 coal is a simple Fortran user block model that predicts the utilities consumptions, operating labor and capital cost that works in combination with an in-line Fortran block which sets the inlet carbon dioxide and product syngas and slag production rates as a function of the coal feed rate. The H1 and H2 heater blocks are present only to force ASPEN/SP to recognize the CO2 and SLAGOUT streams. This preliminary Plant 103 model will be replaced with a more detailed equilibrium-based one that will be developed later. This simple model was developed at this time to allow parallel development of the individual plant models and the combined flowsheet model.

4.4 Syngas Treating and Cooling (Plant 104)

The preliminary Plant 104 (syngas treating and cooling plant) model is a Fortran user block model that simulates both the COS and HCN hydrolysis reaction and product separation sections of the plant. For simplicity, this model ignores any sodium hydroxide addition to the plant since all the sodium eventually leaves in the waste water (as sodium chloride).

4.5 Sour Water Stripping (Plant 105)

The preliminary Plant 105 (sour water stripping plant) model consists of an ASPEN/SP separator block model followed by a Fortran user block model. The Fortran user block model predicts the utilities consumptions, operation labor and capital cost for the sour water stripping plant, and the separator block model does the mass balance calculations. Again, for simplicity, this model was developed to ignore any sodium actually present in the entering sour water.

4.6 Acid Gas Removal (Plant 106)

The preliminary Plant 106 (acid gas removal plant) model also consists of an ASPEN/SP separator block model followed by a Fortran user block model. The Fortran user block model predicts the utilities consumptions, operating labor and capital cost for the acid gas removal plant, and the separator block model does the mass balance calculations.

4.7 Sulfur Polishing (Plant 108)

The preliminary Plant 108 (sulfur polishing plant) model consists of Fortran user block model followed by an ASPEN/SP separator block model. The Fortran user block model predicts the utilities consumptions, operating labor and capital cost for the acid gas removal plant, and the separator block model simulates the sulfur removal as a separate product stream. In reality, the sulfur is removed by adsorption on a ZnO bed. Once the sulfur production is known, the ZnO consumption can be calculated.

4.8 Syngas Wet Scrubbing (Plant 109)

The preliminary Plant 109 (syngas wet scrubbing plant) model is a Fortran user block model which has two input streams and two output streams. This model was developed using fresh water for scrubbing the syngas rather than using water from Plant 104. This simplifying assumption allows the removal of the water recycle loop between Plants 104 and 109 from the overall plant simulation model.

4.9 Air Separation (Plant 110)

The preliminary Plant 110 (air separation plant) model is a Fortran user block model which has a single input stream and two output streams. This model was adapted from the air separation plant model that was developed for the direct coal liquefaction baseline design study, and ignores any other components in the air other than oxygen and nitrogen. If appropriate input parameters are supplied, this model also can be used to study the effect of different product oxygen concentrations.

4.10 Overall Plant Simulation - Area 100

Figure 4-1 is an ASPEN/SP block flow diagram that shows how these plant models are connected to model the Area 100 syngas production section of the complex. The block shown representing the Plant 107 sulfur plant is preliminary and has been included only to provide an overview of the entire Area 100 syngas production section.

In Figure 4-1, the streams are labeled with the plant number that produces them followed by an S and a sequence number; i.e., 102S1, 102S2, and 102S3. Any stream with a C suffix is identical to that stream having the same stream name without the C suffix, but is of the ASPEN/SP CONVEN (only containing conventional components) stream class rather the the MIXNC (containing conventional and non-conventional solid components) stream class. Major input and output streams have names that resemble English words; i.e., MINECOAL and SULFUR. Some other input streams also have meaningful names; such as the make-up water stream going to Plant 104 which is called H2OTO104.

The ASPEN/SP blocks are labeled with a P followed by the plant number, and possibly, a letter suffix. ASPEN/SP blocks without a letter suffix model the entire plant. Suffix C represents an ASPEN/SP stream class changer, suffix F represents a Fortran user block model; suffix M represents a stream mixer; and suffix S represents a component separation block or a flow splitter. A sequence number may be added following the letter suffix, if needed, to provide unique block names; i.e., the P104C1 and P104C2 class changer blocks.

All streams occurring in the simulation before the P104C1, P104C2 and P105C ASPEN/SP class changer blocks are of stream class MIXNC because most of them any contain solids (non-conventional components); whereas, all subsequent streams are of the ASPEN/SP CONVEN stream class. The switching from the MIXNC to CONVEN stream class allows the resulting process simulation model to run faster and use less computer memory and disk storage.

Not shown in Figure 4-1 are several in-line Fortran blocks and two convergence blocks, One in-line Fortran block sets both the Plant 103 inlet carbon dioxide stream and product syngas and SLAGOUT stream flow rates. Three additional in-line Fortran blocks are used to set the make-up water stream flow rates to Plants 104, 106 and 109. Also not shown are two convergence blocks; one of which is used to set the ROM coal feed rate to the complex (stream MINECOAL) to obtain the desired dried coal feed rate to the coal gasification plant, and the other which sets the correct inlet air stream flow rate to the air separation plant (stream AIRTO110) as a function of the coal feed rate to the coal gasification plant.

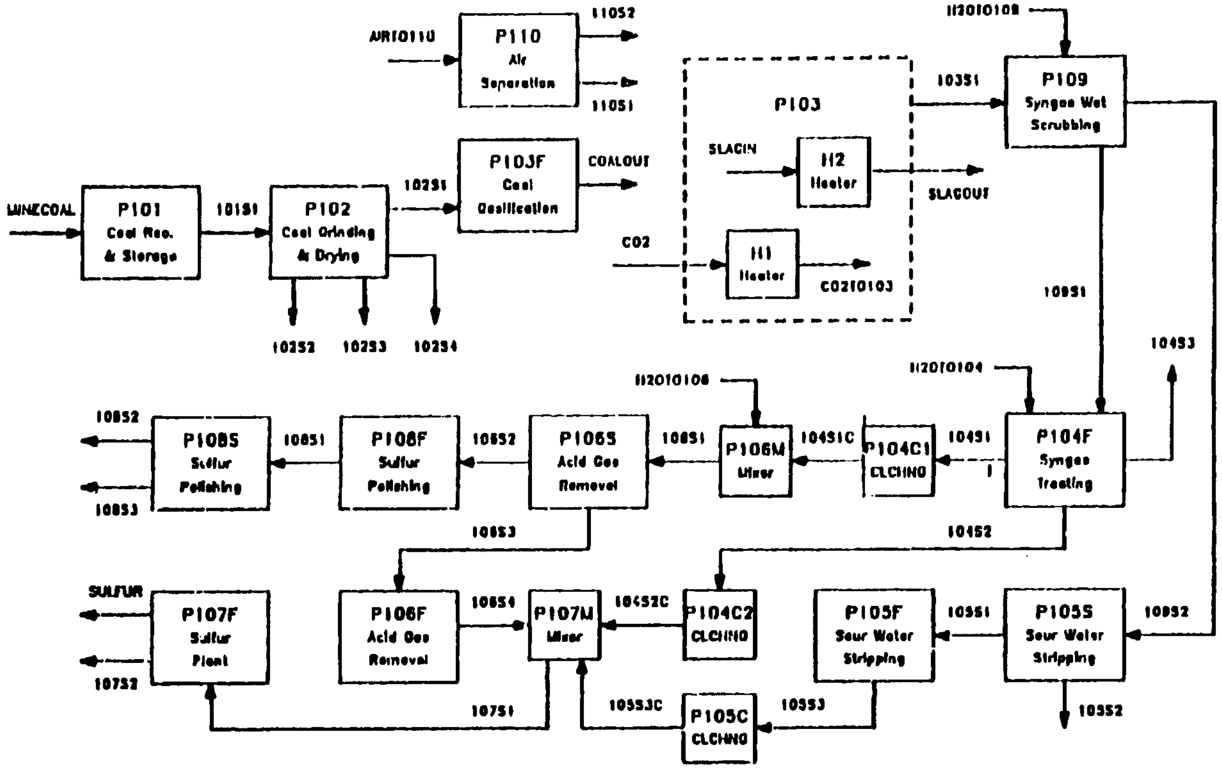


Figure 4–1
ASPEN/SP BLOCK FLOW DIAGRAM FOR AREA 100 – SYNGAS PRODUCTION

1/6/93

Project Management & Staffing Report

Section 5

Project Management & Staffing Report

5.1 TASK 7 - PROJECT MANAGEMENT

The start of Task 4 (Process Flowsheet Simulation Model) is pushed back to the fourth quarter of 1992 so that most of the process design work for the baseline case can be completed. This is to avoid any rework on the PFS model due to the last minute changes on the plant design. This delay is not expected to affect the original study schedule and budget. The overall project schedule status at the end of the reporting period is shown in Figure 5-1.

During this reporting period, a project progress meeting was held at Bechtel San Francisco office. In conjuction with the DOE Contractors' Review Conference, another project progress meeting was held at PETC Pittsburgh office so that the Bechtel and Amoco project teams may have direct interaction with PETC in-house research staff. Bechtel also participated at the Slurry Reactor Hydrodynamics Workshop sponsored by PETC at Pittsburgh. Joe Fox and Sam Tam made a short slide presentation. Sam Tam also presented a paper summarizing progress on the baseline study at the PETC Contractors' Review Conference.

5.2 KEY PERSONNEL STAFFING REPORT

The key personnel staffing report for this reporting period (June 22, 1992 through September 13, 1992) as required by DOE/PETC is shown below:

Name	Function	% Time Spent[a]
Bechtel		
Bruce D. Degen	Process Manager	35
Charles R. Brown	Offsite Facilities	0[b]
Gary Lucido	Cost Estimating	0[c]
Samuel S. Tam	Project Manager	68
Yang L. Cheng	Process Supervisor	80
Amoco		
A. Schachtschneider	Subcontract Manager	6
S. S. Kramer	Process Model/Simulation	7

(a) Number of hours spent divided by the total available working hours in the period and expressed as a percentage.

(b) C. Brown of Bechtel did not spend any time in this reporting quarter because no offsite facilities work was required.

(c) G. Lucido of Bechtel did not spend any time in this reporting quarter because no cost estimating work was required.

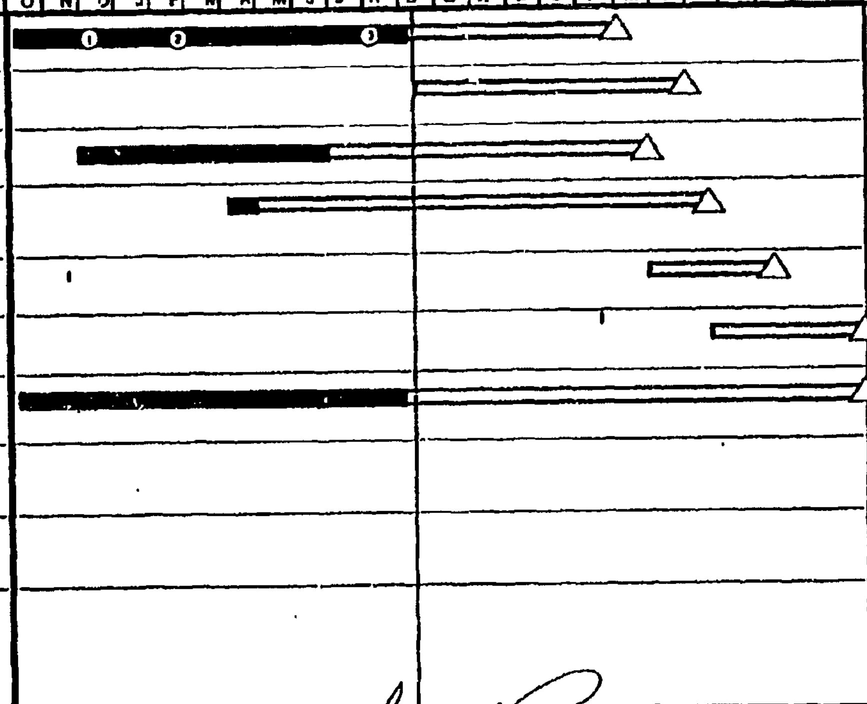
DOE F1332.3
(11-84)

PLAN ■ STATUS REPORT

FORM APPROVED
OMB NO 1901-1400

1. TITLE Baseline Design/Economics for Advanced Fischer-Tropsch Technology	2. REPORTING PERIOD 8/17/92 to 9/13/92	3. IDENTIFICATION NUMBER DE-AC22-91PC90027
4. PARTICIPANT NAME AND ADDRESS Bechtel Corporation, 50 Beale Street, San Francisco, CA 94105		5. START DATE 9/26/91
		6. COMPLETION DATE 9/25/93

9. DURATION: FY 92 (O N D J F M A M J J A S), FY 93 (O N D J F M A M J J A S)



7. ELEMENT CODE	8. REPORTING ELEMENT	10. PERCENT COMPLETE a. Plan	b. Actual
Task 1	Baseline Design	67	67
Task 2	Economic Evaluation		
Task 3	Engineering Design Criteria	74	40
Task 4	Process Flowsheet Simulation Model	31	4
Task 5	Sensitivity Studies		
Task 6	Documentation and Training		
Task 7	Project Management & Administration	49	46

△ Completion
① ASPEN/SP software delivered
② First progress meeting
③ Second progress meeting

11. SIGNATURE OF PARTICIPANT'S PROJECT MANAGER AND DATE Samuel S. Tam Samuel S. Tam 10/13/92

Figure 5-1 Overall Milestone Schedule (as of September 13, 1992)

U.S. DEPARTMENT OF COMMERCE
Technology Administration
National Technical Information Service
Springfield, VA 22161 (703) 605-6000