4. COMMERCIAL PLANT STUDY DESIGN AND ECONOMICS (Reporting Category 3)

The engineering research and development under the Predevelopment Program culminated with the preparation of a new Exxon Catalytic Coal Gasification (CCG) Commercial Plant Study Design. This "CCG Study Design" reflects the current concept of a pioneer commercial plant producing 257 billion Btu per stream day of substitute natural gas (SNG) via catalytic gasification of Illinois coal. The objectives of the CCG Study Design effort were:

(1) To estimate the investment and product SNG cost for a pioneer commercial-scale catalytic coal gasification plant

(2) To identify process areas requiring additional data, correlation work, and/or technology development

(3) To provide a framework "base case" for evaluating new data, process improvements, and optimum process conditions.

The process bases for the Study Design were based on the results of the laboratory and engineering studies carried out during the Predevelopment Program. The key findings of the Study Design are summarized below:

(1) The estimated total investment for a pioneer commercial plant feeding Illinois No. 6 coal and producing 257 billion Btu per stream day of SNG is 1,640 million dollars (M$). This is for a January, 1978, cost level at an Eastern Illinois location. A "process development allowance" and a "project contingency" totaling 470 M$ are included in this estimate. Consistent with Exxon practices for actual projects, these contingencies have been added to predict the total investment required for a pioneer plant reflecting the current early stage of technology development and the uncertainties in project definition.

(2) The estimated cost of SNG produced from this pioneer gasification plant is 6.40 $ per million Btu, based on a nominal January, 1978, startup. This gas cost is a required initial selling price based on 100% equity financing, 15% current dollar DCF return, SNG product revenue escalation of 6% per year, operating costs and by-product revenues escalation of 5% per year, 90% capacity factor, and an Illinois No. 6 coal cost of 20 $ per ton.

(3) On an alternative financing basis of 70% debt/30% equity with 9% interest on debt, the initial gas cost is 4.80 $ per million Btu. The DCF return, escalation, and other economic bases are the same as outlined above.

(4) Several factors could reduce the SNG cost for a pioneer catalytic coal gasification plant below the Study Design range of 4.80-6.40 $/MBtu. These include larger plant capacities, use of surface-mined coals, and increased government financial incentives. Furthermore,

for plants built after the pioneer plant, gas cost savings can be expected by incorporating the learning experience gained in operating the pioneer plant and in carrying out further research and development work.

The Study Design economics are believed to be a realistic prediction of the final costs (in 1978 dollars) for a pioneer commercial plant. However, caution must be used when comparing these economics with published estimates for other coal gasification processes. Such estimates can vary widely depending on the philosophy used to set the process and offsites bases, the detail of the equipment design, and the approach to and time frame for the investment estimate. In addition, as indicated above, the method of financing, plant size, coal type, and the maturity of the technology can have substantial impacts on SNG costs. It is expected that a consistent comparison with state-of-the-art gasification technology, which is currently in progress, will show a significant incentive for further development of the Exxon Catalytic Coal Gasification Process.

4.1 STUDY DESIGN STEPS AND DEPTH OF ENGINEERING DETAIL

The following steps were carried out as part of this detailed engineering study design:

- Project basis setting
- Process basis setting
- Detailed material and energy balances for onsites (process) sections
- Equipment design and specification for onsites sections
- Overall balances for steam, cooling water, electric power, and other plant utilities
- Equipment sizing and specification for offsites sections (*i.e.*, materials handling, utilities, and general offsites)
- Investment estimate
- Economics, including economic basis setting and calculation of product SNG cost.

The CCG Study Design was a substantial effort involving over five man-years of engineering.

The facilities in the gasification plant have been grouped in eight onsites areas (Sections 100-800) and eleven offsites areas (Sections I-XI), as listed in Table 4.1-1. This table also indicates the depth of engineering detail for each section in the CCG Study Design. Complete material and energy balances were developed for most onsites sections, as well as for the steam system, to serve as the basis for equipment specification. Individual equipment items were designed and specified for nearly all onsites facilities and for specialized offsites facilities such as materials handling. The equipment specifications included type, major dimensions, design pressure and temperature, materials of construction, and special mechanical details. Consulting

TABLE 4.1-1

CATALYTIC COAL GASIFICATION
COMMERCIAL PLANT STUDY DESIGN

DEPTH OF ENGINEERING DETAIL

Plant Section Number/Description		Material & Energy Balance	Design Detail
ONSITES SECTIONS			
100	Coal Drying and Catalyst Addition	Complete	Most equipment specified
200	Reactor System	Complete	All equipment specified
300	Product Gas Cooling and Scrubbing	Complete	All equipment specified
400	Sour Water Stripping and Ammonia Recovery	Complete	All equipment specified
500	Acid Gas Removal and Sulfur Recovery	--	Duty specifications
600	Methane Recovery System	Complete	All equipment specified
700	Refrigeration	Complete	All equipment specified
800	Catalyst Recovery	Complete	All equipment sepcified
OFFSITES SECTIONS			
Materials Handling			
I	Coal Handling and Storage	--	All equipment specified
II	Coke/Char Handling	--	All equipment specified
III	Chemicals Handling and Storage	--	All equipment sepcified
IV	By-Products Storage and Shipping	--	All equipment specified
V	Waste Solids Handling and Disposal	--	All equipment specified
Utilities			
VIII	Raw Water/BFW Treating	--	Duty specifications
	Steam Generation and Distribution	Complete	Duty specifications
	Cooling Water	--	Duty specifications
	Electric Power Distribution	--	All equipment specified
	Miscellaneous Utilities	--	Duty specifications
VII	Flue Gas Desulfurization	--	Duty specification
General Offsites			
VI	Wastewater Treating	--	Duty specifications
IX	Safety and Fire Protection	--	Duty specifications
XI	Site Preparation	--	Duty specification
X	Miscellaneous Offsites	--	Duty specifications

assistance was obtained from engineering technology specialists for several key equipment items such as the coal feed system, gasifier, and preheat furnace. Conventional offsites systems such as boiler feed water treating, coal-fired boilers, and cooling towers, were specified based on the overall capacities or duties of component sub-sections. In general, considerable attention was given to the offsite facilities required in a commercial gasification plant. All offsites equipment lists were developed by engineers specializing in offsites design.

Starting with these onsites and offsites equipment specifications, investment estimates were developed using a variety of techniques. To the extent possible, the same computer programs were used as are used to prepare cost estimates for Exxon's commercial petroleum and chemicals projects. These proprietary programs estimate equipment costs and, based on historical correlations derived from actual Exxon projects, also estimate associated labor man-hours and bulk equipment quantities for structures, foundations, piping, etc. Vendor quotes were obtained for several equipment types for which Exxon's commercial experience is relatively limited (e.g., silos, conveyors, coal-fired boilers, hydroclones, etc.). Indirect costs were estimated based on recent experience with large projects. Contingencies were included in the total investment estimate, based on Exxon practices for actual projects.

The study design steps and results are described in more detail below.

4.2 PROJECT BASIS

The major project basis items for the CCG Study Design are listed below:

- Location: Minemouth, Eastern Illinois
- Coal Feed: Illinois No. 6 Bituminous
- Product: Substitute Natural Gas (SNG)
- Plant Size: 257 Billion Btu/Stream Day (HHV)
- Utilities: Steam: Generated in Coal-Fired Boilers with Regenerative Flue Gas Desulfurization
  Electric Power: Purchased
- Environmental: Based on Projected Mid-1980's Air and Water Quality Criteria

All coal mine, coal benefication ("cleaning"), and coal transportation facilities are excluded from the study design. In addition to coal, the gasification plant also receives potassium hydroxide solution and lime for use in catalyst makeup and recovery. Ammonia, sulfur, and sulfuric acid are produced as by-products. Extensive storage is provided for all feeds and by-products. Facilities are provided to supply all plant utilities except electric power. This power is purchased at high voltage (138 kV) from a local utility. Facilities are also provided for wastewater treating (through tertiary treatment with activated carbon) and for waste solids handling, storage, and trucking to a nearby disposal site. Facilities at the disposal site are excluded from the study design: a disposal charge is included in the gasification plant operating costs. A complete summary of the project basis for the CCG Study Design is provided in Table 4.2-1.

4.3 PROCESS BASIS

The second major step in developing the CCG Study Design was to set the process basis. The key process bases for the design of the eight onsites sections are summarized in Table 4.3-1. The laboratory and engineering results obtained in the Predevelopment Program played a central role in setting the process bases, especially in the critical gasification reactor system and catalyst recovery areas. The key process bases for each onsites section are described below.

4.3.1 Coal Drying and Catalyst Addition

In the Coal Drying and Catalyst Addition section, the feed coal is crushed to minus 8 mesh, and potassium catalyst is added to the coal in a water solution. Most of the water in the feed coal and catalyst solution is then removed by evaporation. A small amount of potassium hydroxide solution is purchased as makeup to supplement the potassium salts in the solution recycled from the Catalyst Recovery Section. Engineering screening studies (see Section 3.1.4) showed that KOH produced by electrolysis of KCl is likely to be the lowest cost form of makeup catalyst for CCG plants. The catalyst loading was fixed at 15 wt% K_2CO_3 equivalent on dry coal. This was the approximate catalyst level employed in the majority of the pilot Fluid Bed Gasifier (FBG) runs. Engineering studies to evaluate the impacts of gasifier operating conditions (see Section 3.2.3) indicated that reducing catalyst loading to 10 wt% provided only a marginal economic advantage.

To reduce the heat load on the gasifiers, it was judged that the catalyzed coal feed should be pre-dried to a relatively low moisture level. Coal drying studies carried out prior to the Predevelopment Program indicated that it is probably economical to design entrained coal dryers to attain exit coal moisture levels as low as 4 wt% (dry basis). The prepared coal moisture rate used in the CCG Study Design material and energy balance was slightly above this level: 4.4 wt% on dry coal. The optimum prepared coal moisture depends on cost tradeoffs between the drying and reactor system areas. Further development work will be needed to better define the performance and costs of drying systems for this service and to provide the data base for optimization studies.

4.3.2 Reactor System

In the Reactor System, the prepared coal is pressurized and fed into four gasification reactors. No pretreatment is required because the catalyst reduces agglomeration of caking coals. In the gasifiers, steam reacts with a fluidized bed of catalyzed coal char, in the presence of recycled carbon monoxide and hydrogen. Methane and carbon dioxide, as well as hydrogen sulfide and ammonia, are produced. The gas-phase shift and methanation reactions are maintained essentially at equilibrium over the catalyzed char in the gasifiers. There is no significant net production of carbon monoxide and hydrogen. The resulting overall reaction,

$$\text{Coal} + H_2O = CH_4 + CO_2 + H_2S + NH_3$$

TABLE 4.2-1

CATALYTIC COAL GASIFICATION
COMMERCIAL PLANT STUDY DESIGN

PROJECT BASIS

Plant Location

- Minemouth coal gasification plant in eastern Illinois.

Coal Feed

- Type: Illinois No. 6 bituminous coal, washed ("cleaned") in benefication plant at mine.
- Analysis, wt% (dry):

C	69.67
H	5.05
O	9.45
N	1.84
S	4.19
Cl	0.08
Ash	9.72
	100.00

- Moisture, wt% (as received): 16.5 Average
  19.0 Design Maximum
- Higher heating value, Btu/lb (dry): 12,730
- Transportation: via overland conveyor(s) from mine/benefication plant.
- Storage: 21 days "dead storage" in open pile; 35 hours "live storage" in closed silos to bridge scheduled Sunday mine shutdowns.
- Mine facilities: all mine, coal benefication, and coal transportation facilities are excluded from the study design. (Operating costs and capital charges for these facilities are assumed to be included in the coal feed cost charged to the gasification plant.)

Other Bulk Feeds

- Potassium hydroxide solution (30 wt% KOH) received via rail in tank cars (KOH solution is the makeup gasification catalyst).
- Lime (97 wt% CaO) received via rail in hopper cars (lime is used in catalyst recovery).
- Storage: 31 days for KOH solution; 20 days for lime

Plant Size/Products

- Major product: substitute natural gas (SNG), delivered to pipeline at 1000 psig.
- SNG rate: 257 billion Btu/stream day (higher heating value).
- By-products: aqueous ammonia (20 wt% NH_3); molten sulfur; sulfuric acid (98 wt%).
- By-products storage: 21 days.

Utilities

- Raw water: supplied from a river approximately 1/2 mile from plant battery limits.
- Steam: generated in coal-fired boilers with regenerative flue gas desulfurization.
- Cooling water: supplied via a recirculating cooling tower system. (Rate reduced by using air fin coolers for all major cooling services above 170°F.)
- Electric power: purchased from local utility at 138 kV.
- Plant fuel gas: supplied from methane backed out of product SNG (the primary source) and from lock hopper vent gases.

Environmental

- General basis: provide facilities to meet projected mid-1980's EPA and Illinois air and water quality criteria.
- Wastewater treating: processing includes biological oxidation and tertiary treatment with activated carbon. Facilities also provided for partial reuse of treated water and other plant effluent water.
- Waste solids handling and disposal: facilities are provided to handle and store waste solids and truck them to a nearby disposal site. Facilities at the disposal site are excluded from the study design; a disposal charge is included in the gasification plant operating costs.

TABLE 4.3-1

CATALYTIC COAL GASIFICATION
COMMERCIAL PLANT STUDY DESIGN

KEY PROCESS BASES FOR ONSITES SECTIONS

Coal Drying and Catalyst Addition	
• Number of Trains	3 normal/1 spare
• Makeup Catalyst Type	30 wt% KOH solution
• Makeup Catalyst Rate	189 ST/SD KOH (Contained)
• Prepared Coal Particle Size	Minus 8 mesh
• Prepared Coal Catalyst Loading	15 wt% K_2CO_3 equivalent on dry coal
• Prepared Coal Moisture Content	4.4 wt% on dry coal
Reactor System	
• Number of Coal Feed Trains	4 normal/1 spare
• Number of Gasifier Trains	4 normal
• Gasifier Operating Pressure	500 psia
• Gasifier Operating Temperature	1275°F
• Gasifier Steam Rate	1.585 lb steam/lb dry coal
• Gasifier Carbon Conversion	90% of feed coal carbon
• Gasifier Volume Basis	Gasifier kinetics/contacting model
• Coal Feed Injection Method	Dense phase at bottom
• Catalyst Reactions	Reflected in material and energy balance
• Preheat Furnace Outlet Temperature	1543°F normal/1575°F design(1)
Product Gas Cooling and Scrubbing	
• Number of Trains	4 normal
• High Level Heat Recovery	Via gasifier feed gas/effluent gas exchangers and waste heat boilers
• Final Solids Removal	Venturi scrubbers
• COS Handling	COS hydrolysis converters(2)
• Final NH_3 Removal	Via water scrubbers
Sour Water Stripping and Ammonia Recovery	
• Number of Trains	2 normal
Acid Gas Removal and Sulfur Recovery	
• Number of Trains	2 normal
• Acid Gas Removal	Selective (two-stage) heavy glycol solvent absorption
• Outlet H_2S Content	2 ppm
• Outlet CO_2 Content	500 ppm (0.05 mol%)
• Sulfur Recovery	Claus plants and tail-gas cleanup
Methane Recovery System	
• Number of Trains	2 normal
• System	Cryogenic distillation in a single tower
• Feed Preparation	Cyclic adsorption using molecular sieves and activated carbon
• CH_4 Content of Recycle CO/H_2	10 mol%
• CO Content of Product CH_4 (SNG)	0.1 mol%
Refrigeration	
• Number of Trains	2 normal
• System	Conventional three-level cascade
• Refrigerants	Propylene, ethylene, methane
Catalyst Recovery	
• Processing Basis	$Ca(OH)_2$ digestion plus multi-stage countercurrent water washing
• Number of Trains	2 normal with 2 digestors/train
• Digestion Conditions	300°F, Ca/K = 0.7 lb/lb, 2 hours residence time
• Resulting Soluble Potassium	90% of K in solids to digestion
• Solid/Liquid Separation Method	Hydroclones
• Overall Catalyst Recovery	87% of total loading
• Concentration of Recovered Catalyst Solution	37 wt% K_2CO_3 equivalent

Notes:

(1) Design preheat furnace outlet temperature includes additional heat input capability for gasifier temperature control.

(2) Design COS rate based on equilibrium at gasifier outlet.

is approximately thermally neutral. Only a small amount of heat input to the gasifiers is required, primarily to preheat the feed coal and provide for heat losses. This heat requirement is supplied by superheating the gasifier feed steam and recycle gas mixture in preheat furnaces.

Based on screening studies prior to the Predevelopment Program, the gasifier pressure for the CCG Study Design was set at 500 psia. Directionally, lower pressures lead to increased costs due to higher recycle gas rates, larger volumes of all system gases, and somewhat lower gasification rates. Higher pressures lead to increased costs due to thicker vessel walls and greater mechanical problems, particularly in the coal feeding system. The optimum pressure for operation of the CCG process should be studied further in the next development phase.

The engineering studies of gasifier operating conditions described in Section 3.2.3 of this report indicated that lowering gasifier temperature from 1300°F to 1200°F could reduce both investment and SNG cost by about two percent. Based on this, the gasifier temperature for the current study design was set at 1275°F. Although still lower temperatures could lead to additional savings, the temperature was not reduced all the way to 1200°F because the bulk of the current data base was obtained in FBG material balance periods carried out at about 1300°F. Future research should expand the data base in the 1200-1250°F temperature range to allow further optimization of the operating temperature.

It was shown in the engineering studies cited above that gasifier steam rates of about 1.5-1.6 lb/lb dry coal are probably close to optimum. Lower steam rates substantially increase gasifier volume and raise the preheat furnace outlet temperature above 1600°F. At temperatures above 1600°F, the technical feasibility of the furnace becomes a concern. A screening evaluation of a higher steam rate showed increases in both investment and gas cost of about two percent. The steam rate for the CCG Study Design was actually set so that the gasifier effluent gas was at equilibrium for the steam-carbon reaction over graphite. This is consistent with the approach used in the earlier engineering studies. As previously discussed, the carbon in coal-derived chars has a thermodynamic activity greater than graphite. This allows the steam-carbon gasification reaction to proceed at a significant rate even when the gases are at steam-graphite equilibrium. This basis resulted in a steam rate of 1.585 lb/lb dry coal for the CCG Study Design. Earlier studies also established 90% carbon conversion as the preferred target based on a balance between poor resource utilization at substantially lower levels and the more complex two-vessel reactor system probably required for higher levels (as discussed in Section 3.2.1).

In summary, gasifier conditions for the study design were set at 500 psia and 1275°F, 15 wt% catalyst loading, 1.585 lb steam per lb dry coal, and 90% carbon conversion. The gasifier volume required at these conditions was calculated using the gasifier model developed prior to and updated during the Predevelopment Program. As described earlier in this report, the gasifier model combines basic kinetic correlations developed from fixed bed catalytic gasification data with contacting equations which predict mass transfer effects in fluidized beds. The basic kinetics portion of this model is described in Section 2.2. The bed dimensions calculated for each of four gasifiers were 22 feet inside diameter by 97 feet in height. An additional 25 feet of straight-side was provided above the fluidized beds for solids disengagement. Further development work on the gasifier model is an important objective of the next development phase.

Data from FBG material balance periods were reviewed to develop bases for the rates and properties of gasifier fines carryover and gasifier char withdrawal. These bases were used to design solids recovery equipment for the gasifier effluent gas stream and to design the catalyst recovery system. Data on properties of the "mid-char" from the middle of the FBG fluid bed were also reviewed to help set the basis for gasifier volume calculations using the catalytic gasifier reactor model.

Methods of feeding coal to the gasifiers were studied in considerable depth. It was concluded that the coal feed injection method which appears to have the best chance of technical success is dense-phase pneumatic conveying upward through the gasifier bottom head in multiple feed lines. This approach is expected to provide very good mixing of the feed coal with the char in the gasifier fluid bed. Thus, dense bottom injection was judged less likely to lead to stagnation than the alternatives of dilute-phase conveying or side injection. In addition, dense conveying requires less injection gas than dilute conveying. The injection gas bypasses the steam/ recycle preheat furnace radiant sections. This is required to eliminate any possible coking and plugging which may occur in the coal feed lines if the hot gas (1575°F) from the preheat furnaces is used for injection. With dense phase injection, more gas passes through the furnaces' radiant sections to carry heat into the gasifiers. As a result, the preheat furnace coil outlet temperature in the dense-phase approach can be lowered by about 70°F relative to its value in the dilute phase approach. Thus, there is a significant reduction in preheat furnace investment with the dense-phase approach.

Catalyst reactions believed to occur in the gasifier were incorporated into the material and energy balance. It was assumed that all feed K_2CO_3 decomposes to release CO_2 gas and that the potassium subsequently reacts with the char and ash to produce water soluble and water insoluble forms. The estimated net heat input required for these reactions is large; about 90 MBtu/hr are added to the total gasifier heat input requirements supplied by the four preheat furnaces. Thus, the coil outlet temperature of the furnaces must be raised about 70°F to provide this added heat. If only a portion of the feed K_2CO_3 decomposes, the heat input requirements may have been substantially overestimated. Future research should seek to better quantify the significant material and energy balance effects of catalyst reactions.

Gas-phase reactions, in particular the water-gas shift reaction, may occur to some extent in the preheat furnace tubes. If the mildly exothermic shift reaction occurs in the preheat furnaces, the outlet temperature must be increased to maintain the gasifiers in heat balance. An allowance of 20°F was added to the furnace outlet temperature to reflect this possibility. However, the required outlet temperatures could increase by an additional 40°F if full shift equilibrium were obtained in the furnace tubes. Experimental work to determine whether significant reaction is likely to occur in the furnaces is planned as part of the next development phase.

The required outlet temperature for the four steam/recycle preheat furnaces was calculated to be 1543°F in normal operation. This includes the heat required for coal feed preheat and catalyst reactions, as well as gasifier system heat losses (10 MBtu/hr/train) and the 20°F temperature allowance for shift reaction in the furnace tubes. The furnaces were designed for an outlet temperature of 1575°F to provide 10 MBtu/hr/train additional heat input capability for gasifier temperature control.

4.3.3 Product Gas Cooling and Scrubbing

The next plant section, Product Gas Cooling and Scrubbing, recovers heat and entrained solids from the gasifier effluent gas. Heat is recovered at high temperature levels using both gas-gas exchangers which preheat gasifier feed gases and waste heat boilers which generate 600 psig steam. Venturi scrubbers are utilized for the removal of fine solids. Carbonyl sulfide (COS) is converted to H_2S by hydrolysis in fixed bed reactors. This step was included in this study to avoid potential operating and/or environmental problems in downstream acid gas removal. Further study may show that COS hydrolysis converters are not required. After the gasifier effluent gas passes through low level heat recovery and condensate separation drums, water scrubbing towers are provided to insure complete removal of ammonia. The need to include these scrubbers should be reviewed in future studies.

4.3.4 Sour Water Stripping and Ammonia Recovery

Facilities are provided in the Sour Water Stripping and Ammonia Recovery section to strip H_2S and NH_3 from the sour condensates and venturi scrubber fines slurry produced in Product Gas Cooling and Scrubbing. Ammonia-rich gases pass through further processing steps designed to make aqueous NH_3 for by-product sales. Sulfur-rich gases are sent to Sulfur Recovery.

4.3.5 Acid Gas Removal and Sulfur Recovery

Acid gases (H_2S and CO_2) in the solids- and NH_3-free gasifier product gas from Product Gas Cooling and Scrubbing are separated in Acid Gas Removal. A selective (two-stage) heavy glycol solvent absorption process was chosen as a result of a screening-quality evaluation of three alternative acid gas removal processes (see Section 3.3.1). The H_2S content of the gas is reduced to less than 2 ppm in the first-stage absorbers, and the CO_2 content is reduced to 500 ppm in the second-stage absorbers. The H_2S-containing stripper overhead gases from Acid Gas Removal and Sour Water Stripping are processed in Sulfur Recovery to make by-product elemental sulfur. Claus plants and tail-gas cleanup facilities are included here.

4.3.6 Methane Recovery System

The clean gas leaving Acid Gas Removal consists primarily of CH_4, CO, and H_2. In the Methane Recovery System, this stream is separated into a CO/H_2 stream which is recycled to the gasifiers and a CH_4 stream which becomes the product SNG. A small portion (5%) of the CH_4 stream is used for plant fuel and refrigerant makeup. The separation is accomplished using a simple cryogenic distillation tower supported by refrigeration facilities. To protect the cryogenic equipment from potential freezeout of small amounts of CO_2, H_2O, and light hydrocarbons, the feed gas is first passed through cyclic adsorption vessels packed with molecular sieves and activated carbon. In the distillation tower, the CO content of the product SNG is reduced to 0.1 mol% to meet the specification commonly proposed for SNG from coal. The SNG product stream is now essentially pure (99.9 mol%) methane and has a heating value of 1,010 Btu/SCF. The CH_4 content of the recycle CO/H_2 stream is set at 10 mol%. This specification can be readily obtained without nitrogen refrigeration, and the resulting methane in the recycle gas can be handled by the recycle facilities at a fairly small cost. Higher levels of methane recycle would lead to substantial increases in gasifier steam requirement and total recycle gas rate.

4.3.7 Refrigeration

The Refrigeration Section supplies the auxiliary cooling needed to carry out the cryogenic distillation in the Methane Recovery System. Refrigeration is provided via a conventional three-level cascade system with propylene, ethylene, and methane refrigerants.

4.3.8 Catalyst Recovery

The final onsites section is Catalyst Recovery. Spent solids from the gasifiers (i.e., coarse char/ash particles withdrawn from the gasifier bottom and fines recovered from the overhead gases in tertiary cyclones) are fed to this section as slurries. The spent solids are processed by aqueous $Ca(OH)_2$ digestion and subsequent multi-stage countercurrent water washing to recover most of the contained potassium salts in a concentrated water solution for recycle to Catalyst Addition. An engineering screening study (see Section 3.1.6) showed that $Ca(OH)_2$ digestion to recover water insoluble catalyst is preferred over water wash alone if KOH makeup is priced at or somewhat below the current market level.

The process basis for Catalyst Recovery (see Table 4.3-1) draws directly on laboratory results obtained in the Predevelopment Program. Pilot-scale digestion experiments indicated that at 300°F, at least 90% of the potassium in the solids fed are made water soluble by digestion with a Ca/K weight ratio of 0.7 and a residence time of 2 hours. For example, Run Number 9 in the Secondary Catalyst Recovery Unit (see Section 1.5) achieved well over 90% recovery with this temperature and residence time and a slightly higher Ca/K ratio of 0.81 mole/mole (0.83 lb Ca/lb K). There appeared to be little advantage for higher temperatures or longer residence times. While higher Ca/K ratios did enhance K recovery, the incremental costs for purchasing and processing the additional calcium (as lime) may not be justified.

The solid-liquid separations required in the multi-stage countercurrent washing sequence are accomplished using "hydroclones." (Hydroclones are small, cyclonic solid-liquid separators; they are also referred to as hydrocyclones or hydraulic cyclones. Commercial hydroclone units typically consist of multiple individual cyclones, sometimes several hundred, arranged compactly within a single outer shell.) The particle sizes of the gasifier char and fines were assumed to be unaltered in catalyst recovery. The overall recovery achieved by this system was calculated to be 87% of the total catalyst loading in the coal feed.

The wash water rate was set based on recovering catalyst in a relatively concentrated solution, i.e., 37 wt% K_2CO_3 equivalent. In order to choose the optimum recovered solution concentration, it would be necessary to quantify several effects: e.g., equilibrium and kinetic limitations in $Ca(OH)_2$ digestion, type and performance of the solid/liquid separations equipment, maximum acceptable solids concentration in slurries, and costs of evaporating incremental water. Such optimization work was not possible from the data base obtained during the Predevelopment Program. A major objective of the next development phase will be to obtain more data on the catalyst recovery process and required separations to enable selection of preferred process conditions and separations hardware. Attention will also be directed toward obtaining quantitative closure of the catalyst loop with $Ca(OH)_2$ digestion to examine the potential buildup of soluble species other than catalytically active potassium salts.

4.4 DETAILED MATERIAL AND ENERGY BALANCES

Starting with the process bases just described, detailed material and energy balances were prepared for most of the onsites sections in the CCG Study Design. The tools and techniques used were similar to those used by Exxon for material and energy balances in commercial plant design. Figure 4.4-1 summarizes the overall material balance on a schematic block flow diagram. Eleven major gas streams are tabulated. Rates are also shown for feed coal and the other feed and by-product streams entering and leaving the plant.

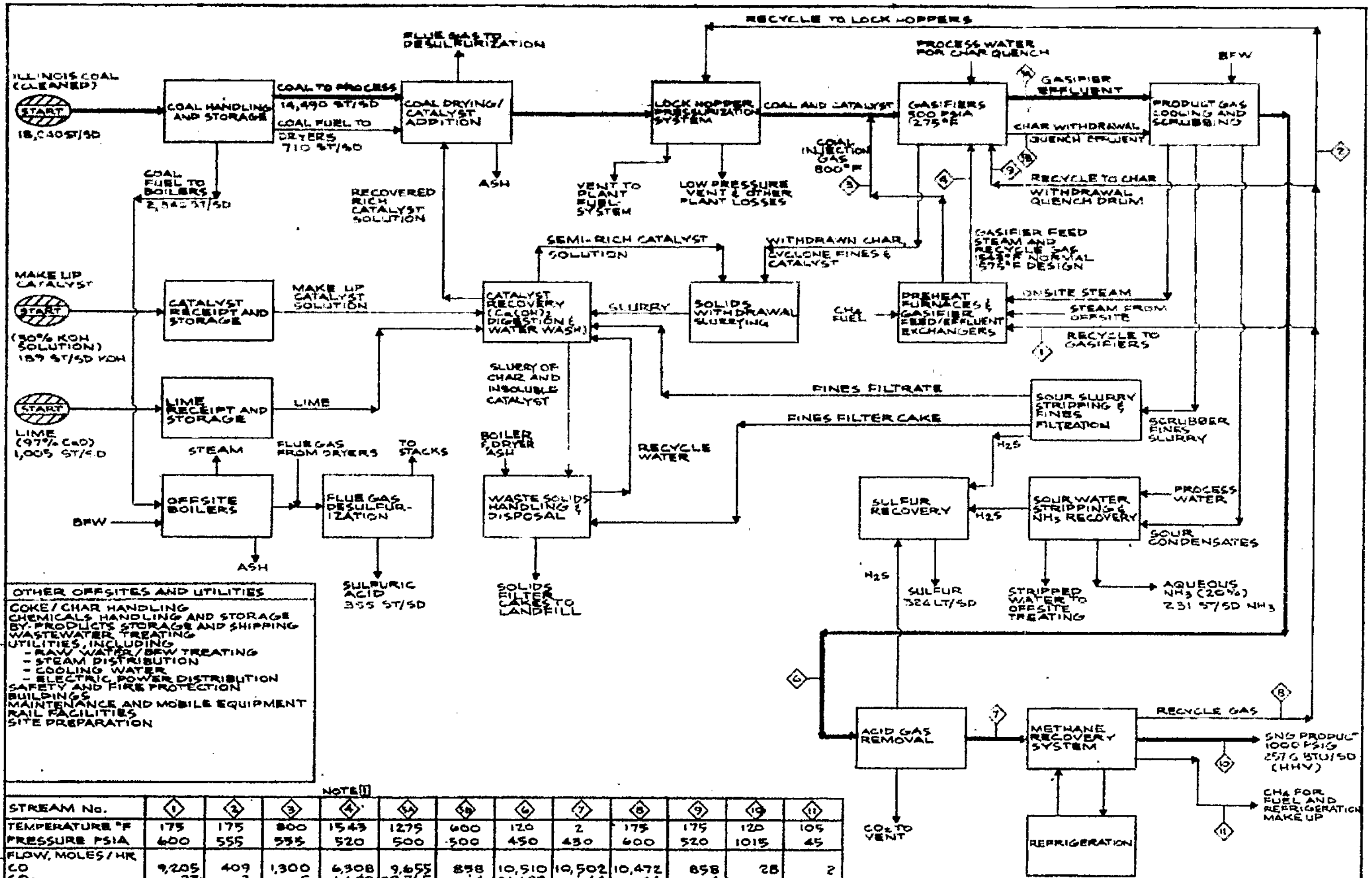
4.5 EQUIPMENT DESIGN FOR ONSITES

After the preparation of material and energy balances, the next major step in the CCG Study Design was the design and specification of equipment for the eight onsites sections. As previously discussed, individual equipment pieces were specified for most onsites sections. The specifications included equipment type, major dimensions, design pressure and temperature, materials of construction, and special mechanical details.

The major plant equipment items and most minor items are shown in the onsites coordination flowplan, Figure 4.5-1. For each section, the coordination flowplan shows facilities for a single process train. The total number of trains provided in the CCG Study Design varies from section to section and is indicated on the flowplan directly under each section title. The onsites sections are described below.

4.5.1 Section 100 -- Coal Drying and Catalyst Addition

Cleaned Illinois No. 6 coal is received from the offsite Coal Handling and Storage facilities and fed at a rate of 14,490 ST/SD to integrated coal crushing/drying systems. These systems combine gas-swept impact mills with entrained drying columns. Four trains of crusher/dryers are provided--three normally operating and one spare. The coal is crushed to minus 8 mesh size



NOTE [1]

STREAM No.	1	2	3	4	5A	5B	6	7	8	9	10	11
TEMPERATURE °F	175	175	800	1543	1275	600	120	2	175	175	120	105
PRESSURE PSIA	600	555	555	520	500	500	450	430	600	520	1015	45
FLOW, MOLES/HR												
CO	9,205	409	1,300	6,308	9,655	858	10,510	10,502	10,472	858	28	2
CO_2	27	3	5	1,629	22,765	4	21,633	44	44	4	—	—
H_2	33,094	1,489	4,674	30,017	34,597	3,082	37,671	37,665	37,665	3,082	—	—
H_2O	11	0	2	84,008	51,380	1,045	416	12	12	1	—	—
CH_4	5,050	227	713	4,337	35,196	470	35,457	35,300	5,747	470	27,942	1,611
C_2H_6	—	—	—	—	5	—	5	3	—	—	3	0
C_6H_6	—	—	—	—	—	—	—	—	—	—	—	—
H_2S	—	—	—	—	947	—	878	—	—	—	—	—
COS	—	—	—	—	14	—	2	—	—	—	—	—
NH_3	—	—	—	—	1,180	—	—	—	—	—	—	—
N_2	3,150	142	445	2705	3,292	293	3,585	3,585	3,585	293	—	—
TOTAL FLOW	50,547	2,270	7,139	131,004	159,031	5,752	110,957	87,111	57,525	4,708	27,973	1,613

NOTE:

[1] 1597 MOLES/HR HAVE REACTED BY THE WATER-GAS SHIFT REACTION IN THE PREHEAT FURNACES

FIGURE 4.4-1

EXXON RESEARCH & ENGINEERING CO.
SYNTHETIC FUELS ENGINEERING

CATALYTIC COAL GASIFICATION
COMMERCIAL PLANT STUDY DESIGN
SCHEMATIC FLOW PLAN AND
OVERALL MATERIAL BALANCE

GPS/RF

77-1779-3

133-A

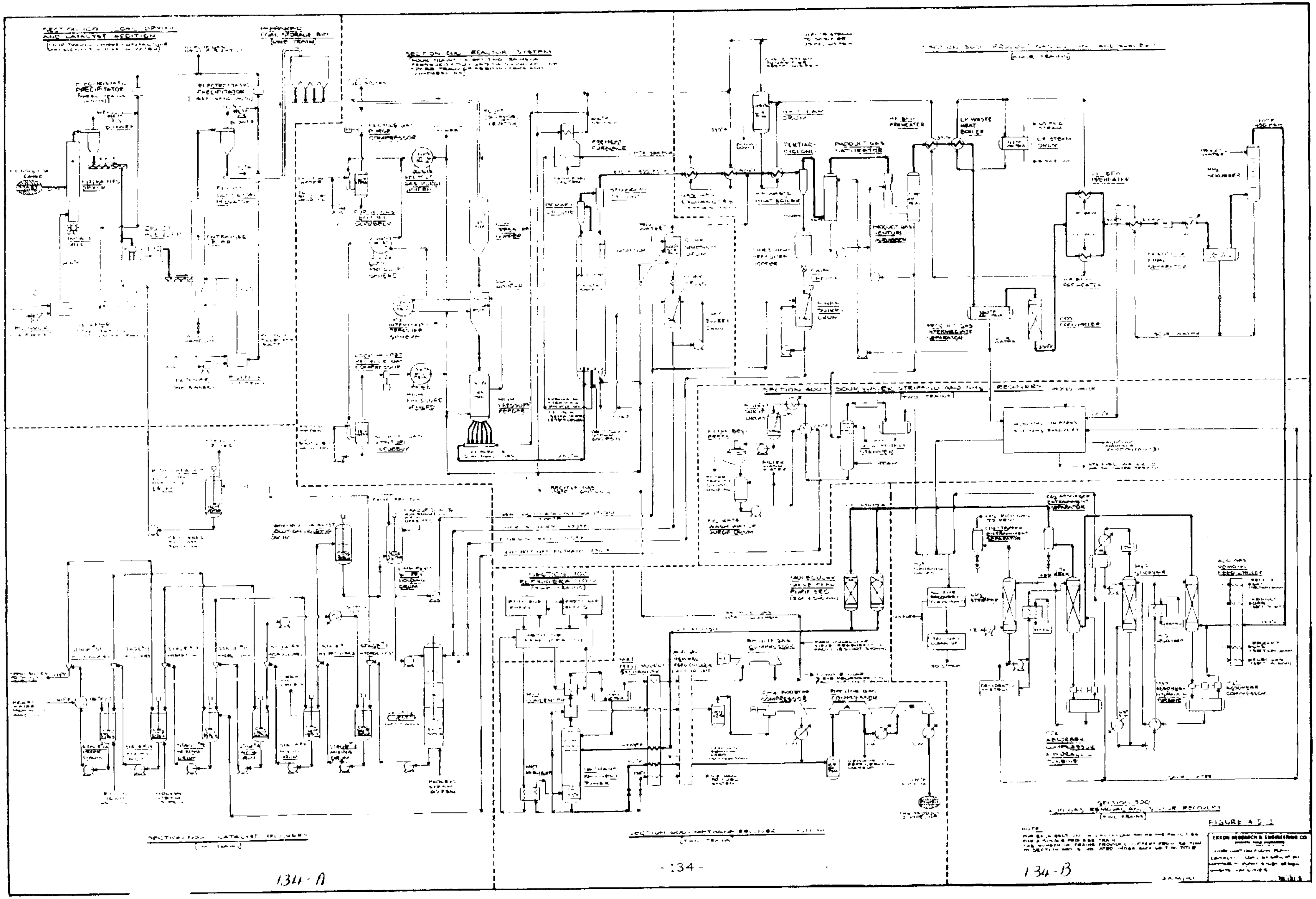


FIGURE 4.5.1
EXXON RESEARCH & ENGINEERING CO
134-A
134-B

and dried to 4 wt% moisture. Flue gas is recirculated to supply the heat and carrier gas for the drying columns. This gas is heated to 900°F in a single coal-fired burner.

Catalyst solution recycled from the Catalyst Recovery section is added to the dried coal in a gentle mixing step. The catalyst solution contains a mixture of potassium salts, with KOH being the predominant form. Lesser amounts of K_2S, $K_2S_2O_3$, and K_2CO_3 are also present. The mixing operation is carried out in zig-zag blenders. The mixture is then dried in entrained drying columns. The heat for this drying is supplied by recirculating flue gas and burning coal as in the first drying step. A total of 710 ST/SD of coal is used as fuel in the two drying steps. Flue gas from both drying steps is sent to electrostatic precipitators for fines removal and then to offsite flue gas desulfurization for SO_2 removal in admixture with flue gas from the offsite boilers. The KOH in the catalyst solution is carbonated to K_2CO_3 by the CO_2 in the flue gas. The resulting prepared coal feed contains 15 wt.% K_2CO_3 equivalent and 4.4 wt% moisture, both expressed on a dry coal basis.

4.5.2 Section 200 -- Reactor System

The catalyzed coal feed prepared in Section 100 is conveyed to the top of the reactor system structures in flight conveyor/elevators and pressurized to 545 psia in lock hopper systems. In addition to four normally-operating lock hopper trains serving the four gasifiers, a fifth spare train capable of feeding each of the gasifiers is provided. The lock hoppers are pressurized with recycle synthesis gas. To minimize gas losses and recompression costs, as well as to smooth out the flows in compressors and vent streams, two pressurization gas systems with gas storage spheres operating at four pressure levels are specified. Each of the two pressurization systems serves two lock hopper trains. The lock hoppers are pressurized in two steps and depressurized in three steps. Portions of the recycle gas used for pressurization are vented to the fuel system and to the flare system. These purges serve to limit nitrogen buildup in the recycle gas loop; otherwise, the N_2 introduced with the feed coal as well as any N_2 produced in gasification would build up indefinitely. The bottom vessel of each lock hopper system serves both as a high pressure storage hopper and as a coal feeder. Six coal feed lines run from this vessel up through the bottom head of the gasifier vessel and pass a short distance into the gasifier fluidized bed. The prepared coal particles are pneumatically conveyed through these feed lines in dense-phase flow using partially preheated (800°F) recycle gas as the conveying medium. Multiple coal injection points are used to assure good mixing and distribution of coal into the bed. One of the six lines feeding each gasifier can be shut down for maintenance without reducing coal feed rate below the design capacity.

The gasifiers operate at 500 psia and 1275°F. The catalyzed char solids in the gasifiers are fluidized with a preheated mixture of steam and recycled CO and H_2 injected through distributors. Four gasifiers are specified with a lining inside diameter of 22 ft and a tangent-to-tangent height of 122 ft. The fluidized bed height is 97 ft, and the outlet superficial gas velocity is 1.09 ft/sec. The residence time is sufficient to gasify 90% of the feed carbon. To reduce heat losses and protect the low-alloy steel shell

(C-1/2 Mo), the gasifier vessels are lined with a 9-inch thick layer of refractories. This lining consists of a 3-inch erosion resistant layer and a 6-inch insulating layer. The inside diameter of the steel shell is thus 23 ft 6 in. The gasifiers are designed to Section VIII Division 2 of the ASME Pressure Vessel Code at a design pressure of 605 psig.

The main reactions taking place in the gasifiers are the highly endothermic steam gasification reaction, the mildly exothermic water-gas shift reaction, and the highly exothermic methanation reaction. The fluidized bed consists of a continuous emulsion phase with intimate gas-solids contact and gas "bubbles" rising up through the emulsion phase. Since steam passes through the bed in bubbles, it must be transferred into the emulsion phase to react with the carbon. CO and H_2 from the recycle gas are also transferred across the bubble-emulsion interface to react via the catalytic action of the potassium-char complex to form methane. The reaction rate in the gasifiers is primarily kinetically limited, although bubble-emulsion mass transfer effects are not insignificant.

As stated previously, the overall reaction is essentially thermo-neutral. Although a detailed analysis has not yet been made, it appears that the different zones of the gasifiers will not differ greatly in temperature. The feed coal is the major external heat sink added to the gasifiers, and the preheated steam plus recycle gas is the major external heat source. These are both added in the same zone near the bottom of the gasifiers, in a fashion conducive to good mixing and heat transfer. In the bulk of the bed, the primary heat effects are the heat-balanced steam-carbon and methanation reactions. In addition, the solids mixing resulting from bed fluidization also contributes to an essentially uniform bed temperature.

The top section of each vessel contains a solids disengagement zone and two external cyclones in series to minimize fines carryover. The use of internal cyclones is an option that could be investigated. The gases leaving the gasifier are essentially in shift and methanation equilibria. With coal feed to the bottom of the bed, pyrolysis products are cracked and only traces of hydrocarbons heavier than methane leave the gasifier. At the bottom of the bed, a char solids stream is withdrawn to control bed level and ash buildup. This char stream flows into a small fluidized quench drum where it is cooled with recycle synthesis gas and process water. It is then fed into a second vessel where it is slurried with semi-rich catalyst solution for feed to catalyst recovery.

The catalytic gasifier is a single-vessel reactor with only one bed and without complicated internals. It is believed that this coal gasification system has the potential for reliable extended-term operation because (1) it is simple, (2) the catalyst prevents caking, and (3) the use of hot steam and synthesis gas for heat input prevents the slagging which can occur when oxygen is used for heat input. Thus, the CCG process has a potential advantage in capacity factor over other developing coal gasification processes which employ more complicated gasifier systems.

The four steam/recycle gas preheat furnaces are designed for radiant section outlet conditions of 1575°F and 520 psia and are analogous to commercial furnaces used for steam reforming of light hydrocarbons in hydrogen and ammonia plants. Convection section heat is used to raise the temprature of the recycle gas used for coal injection to 800°F. The preheat furnaces are fired with gaseous fuel and equipped with air preheaters to reduce fuel consumption. The design methods and approaches used are the same as those for Exxon's commercial steam reforming furnaces. Relatively low heat fluxes are maintained in order to avoid high tube metal temperatures which would require greater-than-commercial tube wall thicknesses. The fuel consumption in these furnaces is relatively small, i.e., only about 5% of the product SNG.

4.5.3 Section 300 -- Product Gas Cooling and Scrubbing

Since the gasifier exit temperature is only 1275°F and heavy hydrocarbons are present only in trace quantities, the high level sensible heat in the effluent gas can be recovered and used for steam/recycle gas preheat and for high pressure steam generation. Four stainless steel gas-gas exchangers are provided for each gasifier train. These are arranged in two parallel trains, each train consisting of two exchangers in series. Here, the steam/recycle gas mixture is heated to 1175°F before entering the preheat furnaces in Section 200. To minimize problems associated with handling residual fines, the gasifier effluent flows downward through each exchanger in a single tube pass. The tubes are supported primarily by a fixed hot end tube sheet. Special provisions are made to support the floating cold end tube sheet and allow for differential tube/shell expansion. The waste heat boiler (one per gasifier) also incorporates downward flow in a single tube pass. The 600 psig steam generated here supplies about 32% of the gasifier steam requirement.

Downstream of the high pressure waste heat boilers, tertiary cyclones recover most of the fines leaving the gasifiers. These fines are slurried with semi-rich catalyst solution and sent to catalyst recovery. Remaining fine solids are removed in a two-stage water-scrubbing system consisting of spray saturator towers and venturi scrubbers. The solids-free scrubber overhead gas is further cooled in high pressure boiler feed water preheaters and low pressure waste heat boilers generating 65 psig steam. Carbonyl sulfide (COS) in the gas stream is catalytically hydrolyzed to H_2S in fixed-bed reactors. Remaining gas cooling equipment includes additional boiler feed water heaters, air-cooled finned exchangers, and cooling water exchangers. Finally, the 120°F product gas is scrubbed with process water to remove residual ammonia.

4.5.4 Section 400 -- Sour Water Stripping and Ammonia Recovery

This two-train section handles the sour fines slurry and sour condensates produced in the previous section. The venturi scrubber fines slurry is steam-stripped to remove NH_3 and H_2S. It is then filtered to produce a solid filter cake for offsite disposal. The filtrate containing potassium salts in dilute solution is sent to catalyst recovery (Section 800). The essentially solids-free water streams from the product gas intermediate and final separators and the NH_3 scrubbers are sent to a H_2S/NH_3 stripping and NH_3 recovery system. This facility includes multiple distillation towers to strip out the contained H_2S and NH_3 in separate streams and recover the NH_3 as an aqueous 20 wt% NH_3 solution for sales. The stripped water is then sent to offsites Wastewater Treating. The H_2S-containing stream is combined with the slurry stripper overhead and sent to Section 500 for sulfur recovery.

4.5.5 Section 500 -- Acid Gas Removal and Sulfur Recovery

A heavy glycol solvent absorption process is utilized to remove the H_2S and CO_2 from the gasifier product gas. One train of these facilities serves two gasifier trains. The process used is a "selective" one, i.e., essentially all of the H_2S is removed in first-stage absorbers and recovered in steam-reboiled H_2S strippers. Most of the CO_2 is removed (down to a concentration of 500 ppm) in second-stage absorbers, stripped out with air in CO_2 strippers, and vented to the atmosphere. The H_2S-containing stream is mixed with other H_2S-containing gases from Section 400 and sent to Claus sulfur recovery plants equipped with tail-gas cleanup facilities. Over 99% of the H_2S in the combined feed gases is recovered as elemental sulfur for sales.

4.5.6 Section 600 -- Methane Recovery System

The clean gas from Acid Gas Removal consists primarily of CH_4, CO, and H_2, along with some N_2. In the Methane Recovery System, cryogenic fractionation is used to split this stream into a product CH_4 stream and a recycle stream. Small amounts of CO_2 and H_2O and traces of light hydrocarbons present in the feed gas are removed by passing the gas through feed purification vessels packed with molecular sieve and activated carbon adsorbents. This step is necessary to prevent these components from freezing in the downstream cryogenic exchangers and towers. Four parallel adsorption vessels are provided in each of the two Methane Recovery System trains. Piping, valving, and regeneration facilities including furnaces are provided to bring each vessel through a four-step cycle involving final adsorption, initial adsorption, heat-up/regeneration, and cooldown phases. A portion of the recycle gas is used for heat-up/regeneration and cooldown.

The product gas from feed purification is cooled to -205°F in plate-fin feed/product heat exchangers and fed to cryogenic fractionating towers operating at overhead condenser outlet conditions of 410 psia and -239°F. The condenser duty is removed by vaporizing methane refrigerant and methane bottoms product. Tower products are heat-exchanged to cool the feed in the feed/product plate-fins. Before entering these exchangers, the tower bottoms are expanded across valves. This lowers the temperature to provide an adequate driving force for cooling the feed.

The product methane stream from the tower bottoms contains 99.9% CH_4 and 0.1% CO. Small amounts of this stream are withdrawn for use as plant fuel and methane refrigerant makeup. The remainder, the product SNG, is compressed to 1,000 psig for delivery to a natural gas pipeline. This stream has a higher heating value of 1,010 Btu/SCF. The recycle gas stream containing CO, H_2, N_2, and 10% CH_4 is compressed for recycle to the reactor system (Section 200).

4.5.7 Section 700 -- Refrigeration

Two three-stage cascade refrigeration systems provide the supplemental heat removal needs for cryogenic methane recovery. Propylene, ethylene, and methane are used as refrigerants in the three loops. The methane recovery tower condenser heat is removed primarily by evaporating methane refrigerant in plate-fin exchangers. Part of the methane refrigerant is recondensed at intermediate pressure in the tower reboilers. The remainder is recondensed by the ethylene and propylene loops.

4.5.8 Section 800 -- Catalyst Recovery

The principal feeds to the two-train Catalyst Recovery Section are the slurry of char/ash solids withdrawn from the gasifiers in Section 200 and the slurry of relatively fine gasifier overhead solids collected by the tertiary cyclones in Section 300. These two aqueous slurries contain most of the ash, unconverted carbon, and potassium remaining from the catalyzed coal feed. The two streams are depressured into $Ca(OH)_2$ digestion vessels operating at 70 psia and 300°F. Feed lime (CaO) and makeup 30 wt% KOH solution are also fed into these digestors. The lime feed rate is 1,005 ST/SD (97% CaO). The CaO hydrolyzes in the digestors to form $Ca(OH)_2$. The ratio of calcium in the lime feed to potassium in the feed char and fines solids is 0.7 lb Ca/lb K. The char and fines slurries are soaked in the digestors for two hours. Each digestor is equipped with a three-impeller agitator to maintain the solids in suspension. Under these "digestion" conditions, 90% of the potassium in the feed solids is solubilized. The balance of the potassium leaves with the solids in water-insoluble compounds. Steam is added to the digestors to preheat the feed CaO and makeup KOH and to maintain the temperature at 300°F.

The effluent slurry from the $Ca(OH)_2$ digestors is pumped through first-stage hydroclones for solid/liquid separation. The overflow from the first-stage hydroclones, the most concentrated potassium solution in the system, contains 37 wt% K_2CO_3 equivalent. This is the recovered "rich" catalyst solution. This stream is fed to holding drums and then pumped back to Section 100 for addition to the feed coal.

The first-stage hydroclone underflow slurry is mixed with the third-stage hydroclone overflow solution in the second-stage mixing drum. The mixed slurry from this vessel is pumped through the second-stage hydroclones. The "semi-rich" overflow solution from this second stage is used to slurry the gasifier char, fines, and lime feeds to the $Ca(OH)_2$ digestors. The underflow from the second-stage hydroclones is fed into the third-stage mixing drum along with the overflow from the fourth-stage hydroclones. This countercurrent water-washing (or "leaching") sequence continues in a similar manner until the fourteenth stage, where clean makeup wash water is preheated and added to the system. The leached solids in the last stage underflow slurry are sent to offsites Waste Solids Handling facilities. Catalyst-containing filtrate from Section 400 enters the wash sequence in the ninth stage, where the concentrations are similar. The flowplan shows no "first-stage" mixing drum because the digestors serve this function.

About 98.5% of the potassium salts solubilized in $Ca(OH)_2$ digestion are recovered in the downstream water-washing stages. Overall, this section recovers 87% of the total potassium catalyst which entered the gasifiers with the feed coal. The remaining 13% is supplied by the makeup potassium hydroxide. The resulting makeup requirement is 189 ST/SD contained KOH, or 630 ST/SD of purchased 30 wt% KOH solution.

4.6 OVERALL PLANT UTILITIES BALANCES

The next major step in the CCG Study Design was to develop the overall plant utilities requirements, i.e., steam, cooling water, electric power, and others. The normal requirements for onsites equipment were calculated based on the item-by-item equipment specifications described previously. The normal requirements for offsites facilities were developed in parallel with sizing calculations. The latter requirements included utilities used by the utilities sections themselves (e.g., steam and power to drive boiler fans, pumps, and pulverizers; steam, power, and cooling water for flue gas desulfurization; steam and power to drive raw water and cooling water pumps; etc.).

In addition to the normal plant requirements, the total design capacities for utilities systems included intermittent loads. Also included in the total capacities were allowances of up to 25% to cover increases in utilities rates as facilities definition improves during project development and to provide reserve capacity in source facilities for startup and emergency needs. The utilities capacity allowances are based on Exxon's experience for a broad range of commercial process plants.

The normal and design utilities requirements for the CCG Study Design are summarized in Table 4.6-1. In every utility system, a design capacity substantially greater than the normal requirements was specified to provide for intermittent loads and capacity allowances. Offsites utilities requirements are significant, particularly in the steam and power balances. Also, almost all of the intermittent loads are for offsites users.

The plant steam balance for the CCG Study Design was done in considerable detail. A limited form of "co-generation" was incorporated. The normal net plant steam requirement of 1,740 klb/hr is generated in offsite coal-fired boilers at 1,250 psig, and expanded in non-condensing steam turbine drivers down to the 600 psig level needed for feed to the gasifiers. Other plant compressors and pumps are driven by non-condensing steam turbines with 600 psig and 150 psig inlet pressures. These steam turbines supply a combined horsepower of about 30% of the total plant horsepower. The remaining drivers are motors run by purchased electric power. Onsite waste heat boilers produce 510 klb/hr of steam at 600 psig, 90 klb/hr at 150 psig, and 680 klb/hr at 65 psig.

This steam balance was prepared based on the assumption that the offsites steam system would be fully independent of the onsites facilities. Thus, low level waste heat which is available onsite from the gasifier product gas is not used to preheat boiler feed water (BFW) makeup entering the offsite deaerators. Instead, the BFW makeup is preheated by first generating

TABLE 4.6-1

CATALYTIC COAL GASIFICATION
COMMERCIAL PLANT STUDY DESIGN

UTILITIES REQUIREMENTS

Utility	Requirements: Normal Onsites	Normal Offsites	Intermittent Loads & Capacity Allowances(1)	Total Design Capacity
Raw Water, gpm	----- 7,300 ------		3,200	10,500
Boiler Feed Water Treating, gpm(2)	----- 3,650 ------		990	4,640
Steam Generation at 1250 psig, klb/hr(3)	1,080	660	380	2,120
Cooling Water, gpm	60,000	17,000	19,000	96,000
Electric Power Distribution, MW(4)	133	14	43	190
Fuel Gas Distribution, MBtu/hr (LHV)	610	10	1,780	2,400
Compressed Air, SCFM	----- 2,400 ------		600	3,000
Inert Gas System, SCFM	------- 440 ------		2,060	2,500

Notes:

(1) This column includes:

- Capacity for intermittent requirements.
- Allowance for estimated increases in utilities loads during project development (except no allowance on gasifier steam rate).
- An additional allowance for reserve capacity in source facilities (e.g., offsite boilers, BFW treating, cooling tower, etc.).

(2) Includes treating for BFW makeup to low pressure and high pressure steam generation services. (See Table 4.7-1 for design capacity breakdown.)

(3) 1250 psig/960°F steam is generated offsite in coal-fired boilers with regenerative flue gas desulfurization.

(4) A substantial portion of plant compressors and pumps are normally driven by non-condensing steam turbines (and are thus reflected in the steam balance). The equivalent normal electric power requirement for these services would be about 47 MW for onsites and 15 MW for offsites.

substantial additional steam in the offsite boilers at 1250 psig, expanding this steam down to 15 psig in various non-condensing steam turbine drivers, and feeding the 15 psig steam into the deaerators. With this approach, the availabilities of deaerated BFW and high pressure steam are not affected by upsets which may occur in the onsite gasifier systems. Thus, the overall plant capacity factor is not reduced by interactions between the onsite and offsite facilities. On the other hand, the larger steam demand for deaeration leads to higher steam generation, steam distribution, and flue gas desulfurization costs. The boiler coal fuel requirement is increased, and the overall plant efficiency is reduced. More extensive studies may identify approaches to use the low level waste heat in the gasifier product without substantial debits in BFW and steam availability. There is potential here for significant cost savings and improved overall plant efficiency.

4.7 EQUIPMENT SIZING FOR OFFSITES

The offsites facilities in the CCG Study Design are grouped in three major areas: materials handling, utilities, and general offsites. As discussed previously, conventional offsites systems such as boilers, cooling towers, etc., were specified based on the overall capacities or duties of component sub-sections. This approach still involves carrying out detailed sizing calculations and specifying many individual components. For example, the Raw Water Section equipment list alone included separate specifications for river-side intake structure and pumps, raw water pipeline, raw water storage basin, cold lime softening and auxiliaries, gravity filters, backwash facilities, low and high pressure delivery pumps, potable water system, miscellaneous transfer pumps, and extensive intermediate and delivery piping. The equipment items were individually sized and specified in the five materials handling sections as well as in the electric power distribution section.

A summary of the offsites facilities included in the CCG Study Design is presented in Table 4.7-1. Materials handling facilities are provided to (1) receive, store, and distribute coal, makeup catalyst, and lime, (2) store and ship by-products, and (3) handle and store waste solids and truck them to a nearby disposal site. (The bases for feed coal storage facilities are discussed below.) A full range of utilities systems are provided including raw water supply, water treating, steam generation and distribution, regenerative flue gas desulfurization, cooling water, and electric power distribution. The general offsites area encompasses wastewater treating (through tertiary treatment with activated carbon), safety and fire protection systems, site preparation, buildings, and miscellaneous items such as maintenance and mobile equipment, communications, and others.

In general, high reliability was a prime design criterion for the offsite facilities in the CCG Study Design. Spares are provided in many sections to insure high system availability. For example, the Study Design includes at least one spare train of coal-fired boilers, flue gas desulfurization facilities, instrument air compressors, flares, and several key coal and lime conveyors and feeders. All normally operating pumps are spared. Several pump and fan services are driven by a combination of steam turbines and motors to maintain the availability of critical utilities such as steam, cooling water, and instrument air during a power failure. Furthermore, as discussed above, the offsites steam system is fully independent of the onsites.

TABLE 4.7-1

CATALYTIC COAL GASIFICATION
COMMERCIAL PLANT STUDY DESIGN

OFFSITE FACILITIES SUMMARY

- **Materials Handling**

  Coal Handling and Storage

  - Normal/design coal rates of 18,040/19,400 ST/TD.
  - 400,000 ST long-term (21-day) storage in open pile.
  - 32,000 ST short-term holdup in 2-14,000 ST storage silos (35 hours) and 2-2,000 ST surge silos (5 hours).
  - Conveying and distribution to users (coal dryer feed, coal dryer fuel, and offsite boiler fuel).

  Coke/Char Handling

  - Coke is stored and delivered to gasifiers for initial startup.
  - Char is removed from gasifiers during shutdown, stored, and returned to gasifiers for subsequent startups.
  - 1,800 ST closed coke storage in 2-900 ST silos blanketed by inert gas.

  Chemicals Handling and Storage

  - 630 ST/SD potassium hydroxide solution (30%); 31-day storage tank; pumps and piping to onsites.
  - 1,005 ST/SD lime (97% CaO in 1/8" granular form); 20-day closed storage silo; conveying to onsites.

  By-Products Storage and Shipping

  - 1,155 ST/SD aqueous ammonia (20%).
  - 324 LT/SD molten sulfur.
  - 355 ST/SD sulfuric acid (98%).
  - 21 days storage for each.

  Waste Solids Handling and Disposal

  - Facilities provided to collect, filter where necessary, and store waste solids from onsites (washed solids slurry from catalyst recovery, fines filter cake produced in sour water stripping section, and fly ash and bottom ash from coal drying/catalyst addition) and from offsites (fly ash and bottom ash from coal-fired boilers and cold lime softening sludge).
  - Total waste solids rate is 7,520 ST/SD (wet basis).
  - 3 days storage in multiple dump bins (8 bins total).
  - Trucks provided (in "Miscellaneous Offsites" section) for transportation to nearby disposal site.
  - Facilities at disposal site are excluded from the study design. (Cost for disposal is charged to the gasification plant.)

- **General Offsites**

  Wastewater Treating

  - Handles "dirty" rainwater and process condensates.
  - Industrial sewer system.
  - Rainwater retention pond for 28 million gallons.
  - API separator for 1,410 gpm.
  - Wastewater treating facilities for 3,700 gpm, including equalization, neutralization, biological oxidation, dual media filtration, activated carbon adsorption, and related auxiliaries.
  - Facilities for partial reuse of treated water and other plant effluent water.

  Safety and Fire Protection

  - Hydrocarbon safety valve release headers, drums, and two 250 ft high flares (including a full capacity spare) to handle releases totaling 1.8 Mlbs/hr (12 billion Btu/hr LHV).
  - Separate release system including similar facilities to handle 0.2 M lbs/hr H_2S-containing gases (1.2 billion Btu/hr LHV).
  - Firewater system for 6,000 gpm including pumps and offsites distribution piping and hydrants. (Onsites distribution piping and hydrants included in onsites "Common Facilities.")
  - Firefighting equipment.

  Site Preparation

  - Site preparation for 415 acres.
  - Roads and fencing for plant.

  Miscellaneous Offsites

  - Buildings totaling 174,000 ft^2 (administration building, laboratory, control houses, warehouse, maintenance shops, etc.).
  - Maintenance and mobile equipment.
  - 5 bulldozers for coal handling.
  - 13 75-ton dump trucks for waste solids disposal.
  - Rail facilities.
  - Communications.

- **Utilities**

  Raw Water/BFW Treating

  - 10,500 gpm raw water makeup delivered from river via pumps and 1/2-mile pipeline.
  - One-day raw water storage.
  - Cold lime softening/gravity filtration for 11,200 gpm.
  - Sodium zeolite softening for 500 gpm low pressure BFW makeup, with 8 hours storage.
  - Demineralization plant for 4,140 gpm high pressure BFW makeup, with 8 hours storage.
  - Includes pumps, piping, regeneration facilities, and other auxiliaries.

  Steam Generation and Distribution

  - 4 coal-fired boilers (1250 psig/960°F) rated at 707 klbs/hr each with all auxiliaries (3 normally operating/1 spare).
  - Steam distribution at 1250, 600, 150, 65, and 15 psig levels.

  Cooling Water

  - Recirculating system with cooling tower, pumps, and distribution piping.
  - 96,000 gpm capacity.

  Electric Power Distribution

  - Power purchased from local utility at 138 kV.
  - Two 190 MW main substations.
  - 173 MW distribution system including captive transformers for major drivers (7 total) and secondary selective local substations (21 total).

  Miscellaneous Utilities

  - Fuel gas distribution for 2,400 MBtu/hr (LHV).
  - Compressed air system for 3,000 SCFM dry instrument air, including 2-100% centrifugal compressors.
  - Inert gas system for 2,500 SCFM dry N_2/CO_2, including 2-50% inert gas generators fueled by methane.

  Flue Gas Desulfurization

  - Serves both offsite coal-fired boilers and onsite coal-fired coal drying and catalyst addition.
  - Sized to handle coal firing capacity as follows:

Flue Gas Source	Illinois Coal Firing Rate, ST/SD: Normal	Design
Offsite Boilers	2,840	3,785
Coal Drying	320	375
Catalyst Addition	390	390
Total	3,550	4,550

  - Flue gases from all three sources are combined and fed to a common FGDS system.
  - The FGDS system employs regenerative process using soda ash makeup and producing a concentrated sulfuric acid by-product.
  - Spare trains are provided for all facilities: absorption (3-50%), regeneration (3-50%), and sulfuric acid production (2-100%)

In the Coal Handling and Storage section, feed coal storage is provided both in an open pile holding a 21-day supply and in two large closed silos holding a 35-hour supply. The silos provide segregated short-term holdup to bridge scheduled Sunday mine shutdowns. The open pile is not used during these weekly shutdowns. It is believed that coal exposed to the weather for a few weeks will degrade and that this will lead to reduced plant SNG output and possibly other operating problems when such "weathered" coal is fed to the gasifiers. Therefore, plant feed will be brought in from the exposed coal pile only during extended mine shutdowns. Further studies of the properties and reactivity of "weathered" coal are desirable to determine whether segregated weekend storage is justified.

4.8 PLANT INVESTMENT

4.8.1 Investment Cost Estimating Approach

An investment cost estimate was prepared for the CCG Commercial Plant Study Design covering the onsite and offsite facilities described above. The first step was to prepare section-by-section estimates of direct material, labor, and subcontract costs on a January, 1978 Eastern Illinois basis. In general, the direct costs were developed using the same estimating tools and techniques used by Exxon for commercial projects at the equivalent level of engineering detail. Proprietary computer programs were used to estimate equipment costs and also to estimate associated labor man-hours and bulk equipment quantities for structures, foundations, piping, etc., based on historical correlations derived from actual Exxon projects. Vendor quotes were obtained for several equipment types for which Exxon's commercial experience is relatively limited (e.g., silos, conveyors, coal-fired boilers, and hydroclones). Vendor quotes were also obtained for the acid gas removal process. Many of the conventional offsites systems were estimated by using proprietary Exxon investment curves. These curves have been developed based on cost estimates for and field experience from commercial projects. Although the basic format of the investment curves is cost versus capacity, the user is required to specify other important design variables in addition to capacity so that an accurate cost estimate is obtained for the particular application.

The next step in estimating plant investment was to add the "indirect" costs associated with project execution. These indirect costs include payroll burdens (payroll taxes and benefits), field labor overheads (costs of field labor supervision, construction equipment, temporary construction, and consumables), contractor engineering, and engineering and construction fees. The indirect costs were based on recent Exxon experience with very large projects. These costs amounted to approximately 50% of the total direct costs.

Finally, a "process development allowance" and a "project contingency" were added to the investment in order to predict the total investment required for a pioneer plant. The process development allowance (25% of the onsites direct and indirect costs) was added to reflect the early stage of technology development. The size of this allowance is a function of the stage of development, and is based on historical data for other Exxon process developments. The project contingency (25% of the total plant direct and indirect costs) was added to reflect uncertainties in project definition consistent with Exxon practices for commercial projects. The indirect costs, process development allowance, and project contingency are discussed further below.

4.8.2 Investment for a Pioneer Commercial Plant

The investment for the CCG Study Design is presented in Table 4.8-1. The total plant investment is 1,640 M$ for the pioneer commercial plant feeding Illinois No. 6 coal and producing 257 billion Btu per stream day of SNG with a higher heating value of 1,010 Btu/SCF. This is for a January, 1978, cost level at an Eastern Illinois location. Caution must be used when comparing this investment with published projections of plant investment for other developing coal gasification processes and for existing technology. Most of the investments reported to date in the literature have been significantly lower.

Analysis of many published estimates indicates that the differences are caused by three major factors. First, the CCG Study Design basis-setting and equipment specification approaches were aimed at providing the most likely final cost for a pioneer commercial plant. Thus, the facilities were designed on a process basis supported by the current data base; potential future process improvements were not considered. The equipment specifications were developed in detail to avoid omissions which might result from overly simplified approaches. As discussed above, the utilities systems included capacity allowances which historical experience has shown to be necessary. Also, the design philosophy incorporated features to achieve a high plant capacity factor. Some or all of these elements are not included in many published estimates of coal gasification plant investments.

The second major difference between the CCG estimate and many published estimates is the inclusion of added indirect costs in the CCG estimate to reflect the effect of "diseconomics of scale" on field labor overheads for very large projects. This previously unanticipated inefficiency has been estimated based on feedback from recent large commercial projects. "Diseconomics of scale" are typically omitted when using estimating techniques which have been developed primarily for conventional-sized projects.

The third major factor contributing to a relatively higher investment for the CCG Study Design is the inclusion of investment contingencies in the detailed CCG cost estimate. Following normal Exxon practices, these contingencies are included at this early stage of process and project development to allow prediction of the most likely pioneer plant investment. As shown in Table 4.8-1, the investment estimate includes a process development allowance of 25% applied to the direct and indirect costs for the onsite facilities. This allowance is applied to estimates for new technology to cover historical increases in investment as process developments proceed from initial research to the pioneer commercial plant. In addition, the investment includes a 25% project contingency on the total plant direct and indirect costs. The term project contingency, as used by Exxon, refers to a statistical factor applied to all estimates at each stage of project development to cover historical increases in cost resulting from more detailed design definition, firming of the project execution plan, site factors, and

TABLE 4.8-1

CATALYTIC COAL GASIFICATION
COMMERCIAL PLANT STUDY DESIGN

INVESTMENT FOR PIONEER PLANT

Basis:
- January, 1978 Instant Plant
- Eastern Illinois Location
- 257 Billion Btu/Stream Day SNG (HHV Basis)

Plant Section	Investment Breakdown Million $	%(1)
ONSITES		
Coal Drying	38	3
Catalyst Addition	24	2
Reactor System	200	17
Product Gas Cooling and Scrubbing	85	7
Sour Water Stripping and Ammonia Recovery	20	2
Acid Gas Removal and Sulfur Recovery	157	14
Methane Recovery System	44	4
Refrigeration	31	3
Catalyst Recovery	39	3
Common Onsite Facilities	63	5
ONSITES SUBTOTAL	701	60
MATERIALS HANDLING		
Coal Handling and Storage	52	
Coke/Char Handling	5	
Chemicals Handling and Storage	27	
By-Products Storage and Shipping	5	
Waste Solids Handling and Disposal	28	
MATERIALS HANDLING SUBTOTAL	117	10
UTILITIES		
Raw Water/BFW Treating	32	
Steam Generation and Distribution	117	
Cooling Water	10	
Electric Power Distribution	29	
Miscellaneous Utilities	5	
Flue Gas Desulfurization(2)	67	
UTILITIES SUBTOTAL	260	22
GENERAL OFFSITES		
Wastewater Treating	41	
Safety and Fire Protection	12	
Site Preparation	8	
Miscellaneous Offsites	33	
GENERAL OFFSITES SUBTOTAL	94	8
TOTAL DIRECT AND INDIRECT COSTS	1,172	100
PROCESS DEVELOPMENT ALLOWANCE (25% of Onsites Direct & Indirect Costs)	175	
PROJECT CONTINGENCY (25% of Total Direct & Indirect Costs)	293	
TOTAL ERECTED COST	1,640	

Notes:

(1) Percentage breakdown of investment is based on total direct and indirect costs excluding process development allowance and project contingency.

(2) Includes desulfurization for flue gases from steam generation (coal-fired boilers) and from coal drying and catalyst addition.

estimate corrections. In summary, the Study Design investment is believed to be a realistic prediction of the final cost in 1978 dollars for a pioneer CCG plant.

With regard to the breakdown of investment cost by plant area, it is worth noting that onsites facilities for catalyst addition and recovery and methane recovery amount to only about 12% of the total investment. This includes 2% for Catalyst Addition, 3% for Catalyst Recovery, and 7% for Methane Recovery and Refrigeration. For a thermal gasification process, the costs of shift conversion, methanation, and heat input via oxygen plants or another system are likely to be substantially higher. In addition, offsite steam requirements are reduced relative to thermal processes as a result of the high level heat recovery from the gasifier effluent gas and the inherent high efficiency of combining all reactions in one vessel. Also, the absence of heavy hydrocarbons in the gasifier effluent minimizes wastewater treating requirements and eliminates the need to incinerate the vent gas from acid gas removal to meet hydrocarbon emissions standards. Despite these cost-saving factors, investments in the three offsites areas still add up to 40% of the total plant direct and indirect costs. This illustrates the importance of studying coal gasification plant offsites requirements in detail at an early stage. Offsites facilities requirements can be an important factor in choosing between gasification technologies. The data needs and cost trade-offs in these offsite areas should be reflected in the overall process development and optimization effort.

4.9 ECONOMICS

4.9.1 Basis for Calculation of SNG Cost

The CCG Study Design economics have been developed for a pioneer gasification plant with a nominal January, 1978 startup. The economic basis for SNG cost calculations is summarized in Table 4.9-1.

The capacity factor for the plant has been taken as 90%. Capacity factor is defined as the actual annual plant SNG production divided by the theoretical production at full design capacity, i.e., 257 billion Btu per stream day for 365 days. A capacity factor of 90% is judged to be reasonable for the CCG process. As discussed earlier, the catalytic gasification system is believed to have the potential for high capacity factor operation because: (1) the gasifiers have only a single fluidized bed, (2) the catalyst reduces feed coal caking, and (3) the use of preheated gases for heat input prevents slagging which can occur when oxygen is used. Spare equipment trains are provided to insure high availability in sections requiring relatively frequent maintenance. Spare facilities have also been provided in coal drying/catalyst addition, coal pressurization, offsite boilers, flue gas desulfurization, coal and lime handling, and other areas. Individual catalyst recovery mixing vessels and their associated pumps and hydroclones can be by-passed temporarily for maintenance with only a slight loss in overall potassium recovery.

The required initial selling price (RISP) for the SNG product has been calculated using the discounted cash flow (DCF) method. The basis used

TABLE 4.9-1

CATALYTIC COAL GASIFICATION
COMMERCIAL PLANT STUDY DESIGN

ECONOMIC BASIS FOR SNG COST

General Basis

• Time	January, 1978 "Instant" Plant
• Location	Minemouth, Eastern Illinois
• Capacity Factor	90%

Basis for Discounted Cash Flow (DCF) Return Calculation

• Project Life	25 Years
• Debt/Equity Ratio	0% Debt/100% Equity Financing(1)
• DCF Return	15% Return on Equity (Current Dollar Basis)
• Income Tax Rate	50% (48% Federal and 2% Local)
• Depreciation Method	13 Years, Sum-of-the-Years'-Digits
• Investment Tax Credit	7% in Year of Expenditure
• Working Capital (Incl. Land)	6.1% of Total Erected Cost
• Startup Expense	6.0% of Total Erected Cost
• Investment Expenditures Profile	15% 3rd Year Before Startup 45% 2nd Year Before Startup 30% 1st Year Before Startup 10% 1st Year After Startup
• Production Schedule(2)	50% 1st Year After Startup 87.5% 2nd Year After Startup 100% 3rd Year After Startup and Thereafter

Raw Materials and Operating Costs and By-Product Revenues

• Unit Costs (1978 Basis)	Listed in Table 4.9-2

Escalation After 1978

• Raw Materials and Operating Costs	5%/Year
• By-Product Revenues	5%/Year
• SNG Product Revenues	6%/Year

Notes:

(1) Alternative financing case also developed with 70% debt/30% equity, 9% interest on debt, 25-year straight-line pay-back of debt principal, and 11-3/4% interest during construction period on debt portion of investment expended.

(2) Production percentages shown are relative to full output after applying the 90% capacity factor. All operating costs and revenues are reduced accordingly in the first two years after startup.

for DCF calculations is summarized in Table 4.9-1. This economic basis has been described in a report covering a study design for the Exxon Donor Solvent Coal Liquefaction Process ("Exxon Donor Solvent Coal Liquefaction Commercial Plant Study Design," Exxon Research and Engineering Company, Interim Report FE-2353-13, Prepared for DOE under Contract No. E(49-18)-2353). Key basis items are highlighted in the following discussion of the SNG cost.

4.9.2 SNG Cost for a Pioneer Commercial Plant

Consistent with the investment of 1,640 M$, the estimated cost of SNG produced from Illinois coal is about 6.40 $/MBtu on a 1978 basis, as shown in Table 4.9-2. This gas cost is a required initial selling price based on 100% equity financing with a 15% current dollar DCF return. It was assumed that SNG product revenues will escalate at 6% per year and that operating costs and by-product revenues will escalate at 5% per year. On a financing basis of 70% debt/30% equity with 9% interest on the debt, the initial gas cost is about 4.80 $/MBtu as shown in Table 4.9-3. This cost is also based on the same DCF return and escalation assumptions.

In calculating these SNG costs, the "cleaned" Illinois No. 6 coal feed was charged at 20 $/ton. The KOH solution used for catalyst makeup was charged at 300 $/ton (contained KOH). This is approximately 15% less than the January, 1978 market price of 355 $/ton. Engineering studies reported in Section 3.1.4 indicated that discounts ranging from 25-45% of current market prices may be possible if a dedicated KCl electrolysis plant is used to produce KOH solution in the relatively large quantities and low purities (98-99%) required for commercial CCG plants. No shipping cost was included, since it was assumed that an electrolysis plant would be located in or adjacent to a commercial CCG plant. (If shipping from current KOH manufacturing facilities in tank cars were required, it would add about 50 $/ton contained KOH.) Lime used in catalyst recovery was charged at 39 $/ton. This cost was based on January, 1978 price quotes averaging 33 $/ton (obtained from lime vendors in the Illinois area), plus 6 $/ton for shipment to the gasification plant via unit train. The bases used for other operating costs and by-product credits are listed in Tables 4.9-2 and 4.9-3.

Out of the total SNG cost of 6.40 $/MBtu in the 100% equity case, slightly over 50% is attributable to capital charges. Coal for gasifier feed and for coal drying and boiler fuel accounts for about 22%. The potassium hydroxide and lime used for catalyst makeup and recovery contribute only 6% to the gas cost.

It is important to recognize that several factors could reduce the SNG cost for a pioneer CCG plant below the Study Design range of 4.80-6.40 $/MBtu. For example, the construction of plants with capacities larger than 257 billion Btu/SD could reduce the gas cost between 0.25 and 0.50 $/MBtu, depending on the actual plant size constructed. The use of surface-mined coal instead of deep-mined coal could reduce SNG cost 0.50 to 0.75 $/MBtu, depending on coal heating value and mining costs. The combined effect of these items could result in a total reduction in gas cost from the pioneer plant of 0.75 to 1.25 $/MBtu. In addition, tax credits, loan guarantees, or other government incentives could further reduce the SNG cost from the pioneer plant.

TABLE 4.9-2

CATALYTIC COAL GASIFICATION
COMMERCIAL PLANT STUDY DESIGN

COST OF SNG FROM PIONEER PLANT WITH 100% EQUITY FINANCING

Basis:
- January, 1978 Instant Plant, Eastern Illinois Location
- 257 Billion Btu/Stream Day SNG (HHV Basis)
- 90% Capacity Factor
- 100% Equity Financing
- 15% Current Dollar DCF Return
- Escalation Rates:
  - Operating Costs and By-Product Revenues at 5%/Year
  - SNG Revenues at 6%/Year
- Total Erected Cost of 1,640 M$ (From Table 4.8-1)

SNG Cost Components	Requirements (At Full Capacity)	Unit Costs (1978)	SNG Cost Breakdown $/Million Btu (1978)
• Illinois No. 6 Coal (Cleaned)			
- To Gasifiers	14,490 ST/SD(2)	20 $/ST	1.128
- To Coal Dryer Fuel	710 ST/SD	20 $/ST	0.055
- To Offsite Boiler Fuel	2,840 ST/SD	20 $/ST	0.221
Subtotal	18,040 ST/SD		1.404
• Major Chemicals			
- KOH Solution (30 wt%)	189 ST/SD (Contained)	300 $/ST	0.221
- Lime (97% CaO)	1,005 ST/SD	39 $/ST	0.152
Subtotal			0.373
• Other Operating Costs			
- Purchased Electric Power	147 MW	2.5 ¢/kWh	0.343
- Raw Water	7,300 gpm	15 ¢/k gal	0.006
- Other Catalysts and Chemicals	Many Items	5.9 M$/yr	0.070
- Wages and Benefits	1,025 Men	21 k$/man/yr	0.255
- Salaries and Benefits	275 Men	25 k$/man/yr	0.081
- Labor Overheads and Supplies	20% of Wages, Salaries and Benefits		0.067
- Materials and Overheads	3.3% of Total Erected Cost/Year		0.641
- Ash Disposal	7,520 ST/SD (Wet)	1 $/ST	0.029
Subtotal			1.492
• By-Product Revenues			
- Ammonia (20 wt%)	231 ST/SD (Contained)	160 $/ST	(0.144)
- Sulfur	324 LT/SD(2)	25 $/LT	(0.031)
- Sulfuric Acid (98 wt%)	355 ST/SD	10 $/ST	(0.014)
Subtotal			(0.189)
• Capital Charges	Per Above Basis (See Table 4.9-1 for Full Basis)		3.343
TOTAL SUBSTITUTE NATURAL GAS COST (RISP)(3)			6.423

Notes:

(1) $k = 10^3$, $M = 10^6$, $G = 10^9$.

(2) ST/SD = short tons/stream day (i.e., one day's operation at full plant capacity). LT = long tons.

(3) Required initial selling price in first year of plant operation (1978).

TABLE 4.9-3

CATALYTIC COAL GASIFICATION
COMMERCIAL PLANT STUDY DESIGN

COST OF SNG FROM PIONEER PLANT WITH 70% DEBT/30% EQUITY FINANCING

Basis:
- January, 1978 Instant Plant, Eastern Illinois Location
- 257 Billion Btu/Stream Day SNG (HHV Basis)
- 90% Capacity Factor
- 70% Debt/30% Equity Financing
- 9% Interest on Debt
- 15% Current Dollar DCF Return
- Escalation Rates:
  - Operating Costs and By-Product Revenues at 5%/Year
  - SNG Revenues at 6%/Year
- Total Erected Cost of 1,640M$ (From Table 4.8-1)

SNG Cost Components	Requirements (At Full Capacity)	Unit Costs (1978)	SNG Cost Breakdown $/Million Btu (1978)
• Illinois No. 6 Coal (Cleaned)			
- To Gasifiers	14,490 ST/SD(2)	20 $/ST	1.128
- To Coal Dryer Fuel	710 ST/SD	20 $/ST	0.055
- To Offsite Boiler Fuel	2,840 ST/SD	20 $/ST	0.221
Subtotal	18,040 ST/SD		1.404
• Major Chemicals			
- KOH Solution (30 wt%)	189 ST/SD (Contained)	300 $/ST	0.221
- Lime (97% CaO)	1,005 ST/SD	39 $/ST	0.152
Subtotal			0.373
• Other Operating Costs			
- Purchased Electric Power	147 MW	2.5 ¢/kWh	0.343
- Raw Water	7,300 gpm	15 ¢/k gal	0.006
- Other Catalysts and Chemicals	Many Items	5.9 M$/yr	0.070
- Wages and Benefits	1,025 Men	21 k$/man/yr	0.255
- Salaries and Benefits	275 Men	25 k$/man/yr	0.081
- Labor Overheads and Supplies	20% of Wages, Salaries, and Benefits		0.067
- Materials and Overheads	3.3% of Total Erected Cost/Year		0.641
- Ash Disposal	7,520 ST/SD (Wet)		0.029
Subtotal			1.492
• By-Product Revenues			
- Ammonia (20 wt%)	231 ST/SD (Contained)	160 $/ST	(0.144)
- Sulfur	324 LT/SD(2)	25 $/LT	(0.031)
- Sulfuric Acid (98 wt%)	355 ST/SD	10 $/ST	(0.014)
Subtotal			(0.189)
• Capital Charges	Per Above Basis (See Table 4.9-1 for Full Basis)		1.709
TOTAL SUBSTITUTE NATURAL GAS COST (RISP)(3)			4.789

Notes:

(1) k = 10^3, M = 10^6, G = 10^9.

(2) ST/SD = short tons/stream day (i.e., one day's operation at full plant capacity). LT = long tons.

(3) Required initial selling price in first year of plant operation (1978).

For plants built after the pioneer plant, gas cost savings can be expected by incorporating the learning experience gained in operating the pioneer plant and in carrying out further research and development work. Historical data from other Exxon process developments suggest that, on a 1978 cost basis, the gas cost from mature technology plants could be 0.75 to 1.00 $/MBtu less than that for the pioneer plant.

As previously discussed, estimates of coal gasification costs can vary widely depending on the philosophy used to set the process and offsites bases, the detail of the equipment design, and the approach to the investment estimate. In addition, as just indicated, the method of financing, plant size, coal type, and the maturity of the technology can have significant impacts on SNG costs. The time frame for which costs are presented is also an important factor. Thus, caution must be used when comparing these economics with published estimates for other coal gasification processes.

From the SNG cost percentages discussed above, it is obvious that coal usage and investment are the crucial factors in coal gasification costs. CCG is believed to have substantial advantages over existing technology in both of these areas. Thus, it is expected that a consistent comparison with state-of-the-art gasification technology, which is currently in progress, will show a significant incentive for further development of the Exxon Catalytic Coal Gasification Process.