SECTION 16

PROJECTED PERFORMANCE

The conceptual commercial Fischer-Tropsch process plant described herein has been designed for a high thermal efficiency, to process the design feed at the design rate, and to make products of stated quantity and quality. The design considered prior industrial experiences, including published information from SASOL, who operate an industrial Fischer-Tropsch plant. Also considered were the German World War II plants, American post-World War II attempts at commercialization, and the U.S. Government-sponsored work at Louisiana, Mo. A contributing data source is the ERDA (Bureau of Mines) experimental work spanning the past 15 years.

This conceptual design obtains a high thermal efficiency with high recovery of reaction heat by using techniques investigated in the ERDA experimental program with flame sprayed catalyst on heat transfer surfaces.

The design is considered to be workable. Commercialization would require favorable results from additional development work, which will be discussed below.

The selection of materials of construction consists of generally high alloy resistant materials specified to provide confidence that they will perform.

16.1 GASIFIER

The entrainment gasifier performance is expected to conform to flowsheet conditions subject to successful operation of in-progress pilot plant units, including the ERDA-sponsored Bi-Gas unit. Results of studies and successful PDU work performed using this type of gasifier during the past 10 years are indicative of its practicality. This experimental work has resulted in the construction of a two-stage entrainment slagging-type pilot plant scale gasifier at Homer City, Pa. The results of the imminent startup and operation of this pilot plant are expected to corroborate the gasifier design used while possibly suggesting some modifications.

16.1.1 COAL FEED SIZE

A nominal coal feed size of minus 20 mesh x 0, rather than the conventional 70% minus 200 mesh size used in conventional pulverized coal fed boilers, was used for this design. The coarser feed is expected to have better characteristics in the compression screw feeders used to feed the gasifiers. Char recovery efficiencies are expected to be higher due to the coarser average particle size as compared with the 70% minus 200 mesh alternative. The medium-to-fine coal feed size is expected to handle satisfactorily

in the entrainment gasifier since it more closely resembles a cyclone burner type of boiler than the pulverized coal type. Velocities specified within the gasifier are such that the coarser particles will be satisfactorily entrained.

The BCR pilot plant is equipped to produce a similar coal feed, and also a finely-ground feed. The operating results of the two-stage entrainment gasifier on both medium-to-fine and pulverized coal feed can be used to confirm or modify the design presented here.

16.1.2 COAL FEEDING

The conceptual plant gasifier is fed with dry coal by compression screw feeders. Screw feeders that will transfer ground coal from atmospheric conditions into reactors operating at moderate pressures, such as the Fischer-Tropsch gasifier, and at high pressures are under development; ERDA-FE is sponsoring three such development activities. The power requirement used in this design was estimated and extrapolated from similar operations at lower pressure.

Gasification of the extruded feed coal is dependent on breaking of the feed coal by steam jets. Additional fracturing is also expected from the sudden vaporization of volatiles, including inherent moisture in the feed coal.

The compression screw feeder was selected on the presumption that present developmental work will be successful and timely. Lower pressure compression screw feeding has been reported by General Electric.[24] It is felt that ultimately this method will prove to be the simplest and most economical.

The compression screw feeder used in this design consists of a heavy-duty screw having a decreasing pitch such that the conveyed material would be compressed at the point of discharge to a pressure slightly above the 468-psig gasifier pressure. Mixing within the unit would be minimized, and a high pressure die plate is not considered necessary. Four such coal feed injectors with water-jacketed barrels are provided for feeding each gasifier. Four screw injectors are also included to transfer char from the recovery system at approximately 440 psig into the gasifier's bottom stage. These will not require the degree of compression required of the coal injectors because of the low pressure differential.

16.1.3 SLAG DISPOSAL

The slag disposal system provided in the gasifier design is of the type used successfully with slagging coal-fired boilers. These systems are also available for pressurized boiler operation.

16.2 SYNGAS HEAT RECOVERY

Syngas flows from the top of the gasifier through four parallel, abrasion-resistant, ceramic-lined ducts into five vertical straight-tube heat exchangers, 13-1601, -1602, and -1604 (3 shells). The gas contains approximately 30 wt% entrained char or 0.125 pound solids/acfm entering the heat exchangers at 1,700°F, and 0.2 pound solids/acfm leaving the heat exchanger at 950°F. This is considered a light solids loading by pneumatic conveying standards. Velocities downward through these heat exchanger tubes are approximately 75 fps, which is within the design range for pneumatic conveying systems. Accordingly, abrasion of the heat exchange tubes is expected to be within commercial experience limits.

A suitable expansion joint will be provided for each heat exchanger commensurate with each unit's service. This will consist of a modified floating-head design to allow for the differential thermal expansion between the shells and the tubes.

In order to minimize tube abrasion, the tube inlets are designed with fitted abrasion-resistant ceramic ferrules. Thus, the turbulent entry area of the tubes will be protected. Stainless steel No. 321 with 1/16 in. corrosion allowance, equivalent to Schedule 80 pipe, was selected as the tube material because of its toughness. This material is deemed satisfactory for service up to 1,500°F. Maximum tube temperature is expected to be 1,300°F. The heat exchanger inlets and outlets are lined with ceramic for shell abrasion resistance in these areas.

The syngas coolers are patterned after known 450-psig units used in petroleum fluid cracker regenerator gas outlet coolers and fluid catalyst line exchangers. These exchangers are also similar to the 750 psig tubular boilers cooling syngas and ash under similar conditions in a coal-based ammonia plant[25].

The four parallel streams of 950°F gas, with entrained char, leave the heat exchangers at the bottom and enter individual sets of two-stage cyclones (four sets in all). Connecting ducts are refractory lined, short as possible, with a minimum of bends. All bends are long sweep and designed with solids retention areas where impingement occurs so that direct impingement on duct walls is minimized.

Dependable performance is expected from these synthesis gas cooling units.

16.3 GAS CLEANING

The production of a synthesis gas feed free of solids contaminants is essential to proper catalysis in the Fischer-Tropsch reactors. The series of solids removal equipment consisting of two-stage, high efficiency cyclones, a hot dry electrostatic precipitator, a Venturi wet scrubber, and a final wet electrostatic precipitator is expected to ensure a suitably clean synthesis gas.

Perhaps more than sufficient solids removal capacity was included in the design; however, the incentive to assure supply of a cleaned synthesis gas is considered adequate to warrant the equipment selection.

16.4 SHIFT CONVERSION

The sour gas shift selected in preference to a sweet shift conversion is workable. Both sour and sweet shift conversion processes are currently in commercial use.

16.5 EXTENDED SURFACE (ISOTHERMAL PLATE TYPE) CATALYTIC REACTORS

Extensive review of research work on catalyst-coated plate-type reactors by ERDA laboratories led to selection of this type of reactor for the three catalytic reactions which are shift conversion, Fischer-Tropsch synthesis, and methanation.

It is considered that the success achieved by the ERDA (PERC) laboratory at Bruceton, Pa. with experimental catalyst-coated plate reactors in Fischer-Tropsch syntheses and methanation presented a practical catalytic process. The principle of conducting reaction heat directly from a catalyst coating through the supporting metal into a coolant in contact with that metal is considered basically sound. Advantages were such that this type of catalytic reactor was also adapted to the shift conversion reaction for this Fischer-Tropsch design. Detailed discussion of the basis for this type of catalytic reactor selection was given in Section 15. Further comments are included in Section 18 under paragraph 18.5.

In order to successfully design the FSC reactors, it was necessary to independently develop design procedures. These procedures define the effects of simultaneous (1) liberation of exothermic reaction heat on the external surface of an extended surface heat exchanger and, (2) heat transfer from the reaction mass to boiling water inside the tubes. A summary of this design procedure development is included at the end of this report as APPENDIX C.

Discussions with flame-spray equipment manufacturers further strengthened the opinion that the developmental work underway on this type of reactor will culminate in commercially attractive processes.

The foregoing covers the primary conceptual items of the Fischer-Tropsch plant design. Regardless of the confidence that there may be in the design presented, a final design should be based on successful pilot plant data.

A discussion covering details and the various alternates considered during the development of the design is contained in Section 18.

SECTION 17

POTENTIAL IMPROVEMENTS

The conceptual Fischer-Tropsch commercial plant design includes developments and improvements to give a high-thermal efficiency. Some concepts that are considered candidates for further improvement will be discussed in this section.

17.1 SLAG QUENCH HEAT RECOVERY

The molten slag draining to the bottom of the gasifier is quenched to a low temperature so it may be handled as a coarse sand slurry. There is no heat recovery. A slag quench by boiling water would recover most of the heat in the slag as steam. This contaminated steam could go to the sour shift reactors. Plant efficiency would increase about 0.6%.

Slag recovery, handling and disposal procedures would require extensive analysis and innovative design.

17.2 SYNTHESIS AT 1000 PSI

The majority of the ERDA experimental data on flame-sprayed catalyst is at 400 psig for Fischer-Tropsch synthesis. This pressure has proven satisfactory in commercial operation to avoid carbon formation. Some tests have been run by ERDA at 1000 psig, and more are planned. Although more oxygenates were formed, no radical yield change is anticipated. The economy of operating at 1000 psig appears favorable, with reduced equipment sizes in the gas-treating sections, and reduced pipeline SNG compression. However, there appears to be no efficiency gain; added costs of oxygen compression tends to be offset by savings in SNG compression. Reduced gas-treating cost could increase the overall thermal efficiency 1 to 2%.

17.3 GAS CLEANING SIMPLIFICATION

Information has been received regarding the commercial development of an "ionizer" purported to agglomerate fine particles at high velocities. Effective performance at velocities as high as thirty feet per second is claimed as contrasted to electrostatic precipitators' operation at velocities of approximately five feet per second.

Particle collection at a capacity increase of 90% plus greater recovery efficiencies at a cost equal to about 60% of that for an equivalent wet

or dry electrostatic precipitation is claimed to be possible. For dry solids separation, the ionizer would be installed in series with a small precipitator which would serve primarily as a finishing unit.

Assuming successful development, the ionizer could substantially simplify and reduce the capital cost of gas cleaning.

17.4 ASSESSMENT AND DELETION OF MARGINAL HEAT RECOVERY CAPACITY

The complex as designed produces approximately 140 MW of electric power in excess of that required to operate the facility. The sales value of this export power determines the justification for use of heat-recovery equipment. A low sales value may not be sufficient to justify the combined larger electrical generators and the extra heat-recovery capital costs. The breakeven point for the power generation as presented appears to be approximately 1.1¢ per KWH. Deletion of the equipment to produce this 140 MW would result in a decrease of approximately 1.5 percent thermal efficiency and would require extensive redesign.

17.5 EXTENDED SURFACE COATED CATALYST REACTOR DEVELOPMENT

The type of reactor selected for the Fischer-Tropsch synthesis, shift conversion, and methanation steps is based on research done by the ERDA-PERC laboratories. Discussions with flame spray equipment manufacturers strengthened the opinion that this type of equipment could be commercially feasible.

Extensive development using large scale equipment and automated methods of applying catalyst coatings is required for commercialization of this procedure. This development work is currently under way. Acceleration of this effort could result in realization of timely commercial success.

17.6 SINGLE-TRAIN PLANT

Two trains were provided in the plant design to reduce vulnerability to complete shut down in case of operational failure. Capital costs and production costs savings over the present design could be effected in many areas with a single-train plant design.

A single-train plant gasifier handling the full 30,000 tons of coal per day would have an internal diameter of 23'8" compared to 16'9" for each of gasifiers specified in this design. This would reduce the cost of the gasifier unit by approximately 15%. Consolidations throughout the plant could be effected for a single-train design, with a resultant lower capital cost.

A single-train plant design may be justified at some future date following successful demonstration plant design and operation.

SECTION 18

PROCESS DESIGN CONSIDERATIONS

The development of a conceptual design for a multi-step process plant involves a number of choices in most of the major process steps. In each of the process steps an attempt was made to analyze and select the best method. This section presents a discussion of alternates considered during the course of development of the design.

18.1 GASIFIER SELECTION

The various known types were considered and comparisons made.

18.1.1 FIXED BED TYPE

The continuous, fixed-bed gas producer at atmospheric pressure is used in industry and adapted to oxygen-steam gasification in the Lurgi pressurized design. Fixed-bed designs are adaptable to smaller units with mechanical bed-stirring devices, use large amounts of steam, and require selectively sized non-caking coal. Oxygen usage is low. Syngas is normally high in methane and hydrogen content, and production of tars and phenols are high. Capital cost and maintenance are high.

18.1.2 FLUIDIZED BED GASIFIER

The characteristics are similar to the fixed-bed type except the fluid gasifier can be built in large sizes. Char loss with the ash can be a problem. Caking coals must be pretreated. Steam use is high. Syngas is high in methane and hydrogen.

18.1.3 ENTRAINMENT GASIFIER

The entrainment gasifier was selected because any feed coal, caking or non-caking, can be charged, and char feed can be used. This design adapts to a high throughput in a single unit. The syngas has a lower hydrogen content adaptable to adjustments required for the Fischer-Tropsch synthesis. Organic sulfur, phenol, and tar production is at a minimum. Oxygen consumption is high, resulting in excess heat which is recoverable for utilities generation.

18.1.4 TWO-STAGE ENTRAINMENT GASIFIER

The conceptual design uses a developmental two-stage gasifier with the coal feed converted to syngas and char; the char is recovered and

reacted in the lower stage (stage I). Conversion of the coal to methane in the gasifier is lower than comparable fluid or fixed-bed gasifiers. The two-stage entrainment gasifier is expected to operate at higher efficiency and with reduced oxygen use, relative to a single-stage unit. Recovery and feeding of the hot recovered char is required.

Table 18-1 presents the results of a screening-type economic comparison of the projected economics based on analysis of available information on the Lurgi-Westfield fixed bed, fluid-bed synthane type, and the two-stage entrainment gasifiers. The lowest production cost is indicated for the two-stage entrainment gasifier.

18.1.5 SINGLE-STAGE ENTRAINMENT GASIFIER

The present commercial entrainment slagging gasifier, Koppers-Totzek, is single stage, and operates at atmospheric pressure. This design has proven itself reliable. This reliability factor must be weighed against a 25% larger oxygen plant requirement and a 14% greater coal consumption for a given rate of syngas production. A greater quantity of heat would be recoverable relative to use of a two-stage entrainment gasifier, which requires additional excess electricity generation. A minimum net thermal efficiency drop of 3%, for a total net efficiency of 66.7%, would occur.

The capital cost increase of the added oxygen plant capacity must be weighed against partial elimination of the hot-char recovery system and a less costly gasifier. Production costs using the single-stage gasifier would be higher than those from a two-stage entrainment gasifier plant.

Table 18-2 presents the results of a screening-type economic comparison of the single- and two-stage entrainment gasifiers. The two-stage system shows an annual savings, expressed as equivalent uniform annual cost, over the single-stage of approximately $19 million or 12¢ per million Btu based on a 12% DCF and a 20-year project life. Figures 18-1 and 18-2 contain comparative flow diagrams for the one- and two-stage systems. Figures 18-3 and 18-4 present sketches of the two-stage and single-stage gasifiers, respectively.

18.2 COAL FEEDING

The compression screw feeder was selected on the basis that present developmental work will be successful. Lower pressure compression screw feeding has been accomplished. It is felt that ultimately this method will be assured as the simplest and most economical.

Lock hopper systems were considered but rejected because of their complexity, high capital, and operating costs. Lock hoppers also work best on coarse coal without fines as is the case with the Lurgi gasifier systems.

Water slurry feed was discarded because of the substantial thermal efficiency loss resultant from vaporization of the slurry water. This loss would be an approximate 10% drop in thermal efficiency. A 60% overall thermal efficiency process would be considered undesirable.

18.3 WATER GAS SHIFT UNIT

18.3.1 GENERAL

The conceptual plant design is based upon the use of a sour-shift conversion unit, as opposed to a sweet-shift unit in which only trace quantities of sulfides are present. Isothermal FSC type catalyst reactors are used.

The reactor design is the same basic design as used for the Fischer-Tropsch and methanation reactors, although there is as yet no experimental data to indicate the performance of this type of design for shift conversion.

18.3.2 EVALUATION OF SOUR AND SWEET SHIFT CONVERSION

The shift reaction could take place either before or after acid-gas removal. The conventional sweet-shift operation takes place after acid-gas removal, in the presence of minute quantities of sulfur, and small quantities of carbon dioxide.

The sweet-shift unit handles a smaller volume of gas than the sour-shift unit since the acid gas constitutes 9.5% of the gas volume from the gasifier. However, a second acid gas treating unit is required to reduce the carbon dioxide content of the hydrogen-enriched syngas to the Fischer-Tropsch unit.

Sour-shift conversion results in an overall shift steam and utility saving equivalent to an improvement in overall plant thermal efficiency of 1.3%. The saving in reduced acid gas treating utilities account for approximately 1% of this overall efficiency improvement. The FSC type reactor design is particularly suited to the sour-shift operation, as compared to a fixed-bed granular reactor, since it can more easily process the dusty syngas with low-pressure drop, as dust or carbon are expected to sweep through without buildup. The surfaces can be readily cleaned of solids by flushing. The saving in compression horsepower resulting from the use of a FSC reactor rather than a granular-bed reactor is about 10,500 bhp. This is equivalent to an increase in overall plant efficiency of approximately 0.6%.

18.3.3 SHIFT CONVERSION STEAM REQUIREMENT

The shift conversion and gas cleaning units operate at minimum operating costs when the shift reactor uses the lowest steam-to-gas ratio required for satisfactory performance of the catalyst.

The steam ratio used for design is the lowest possible for a 50°F approach to equilibrium. In operation, this steam ratio can be increased and accommodated by the design, as can the flow through the shift unit be raised to full flow of syngas. The temperature of operation of the reactor can vary from 600°F to 750°F within limits of the Dowtherm coolant. These factors allow considerable latitude to secure optimum catalyst performance.

18.3.4 CONVERSION OF TARS, ORGANIC SULFIDES, AND CARBONYLS

Tars, if present, are expected to be hydrocracked on the FSC shift reactor catalyst. Granular catalysts tars may coke, resulting in some pore blockage with subsequent loss in activity. Some conversion of phenols to hydrocarbons is expected. Organic sulfides will be hydrogenated to hydrogen sulfide, and carbonyl sulfide content is reduced.

The inherent low-pressure drop of the FSC reactor would permit all the syngas to be passed through the shift converter, thereby reducing the quantity of organic sulfides to be removed by the downstream units, without a substantial penalty in compression horsepower. However, the minimum shift charge steam to dry gas ratio must be maintained. An additional 664,700 lb/hr of shift steam would otherwise be required, which is equivalent to a reduction in the overall plant thermal efficiency of 2.1%.

18.4 GAS TREATING AND CLEANING

18.4.1 ACID GAS REMOVAL

The acid gases are formed in the gasification, shift reaction, and Fischer-Tropsch synthesis units. Both hydrogen sulfide and carbon dioxide are present in the raw syngas from the gasifier. More carbon dioxide is produced by the water-gas shift reaction, and hydrogen sulfide is produced from organic sulfur decomposition in the shift reactor. These acid gases are removed in Acid Gas Removal Unit 14. Removing the carbon dioxide from the shift reaction in the same step as the hydrogen sulfide results in an increase of plant efficiency of about 1% when compared with the alternate of removal of hydrogen sulfide first, through a sweet-shift reactor, followed by subsequent carbon dioxide removal.

Carbon dioxide product of the Fischer-Tropsch synthesis is removed from the recycle gas and net gas stream. A concentration less than 7% at the reactor outlet is maintained. Alternatively, the net Fischer-Tropsch vent carbon dioxide could be removed from the final SNG. Removal of the carbon dioxide by hydrogenation to methane was deemed preferable to adding another scrubbing step.

For acid gas removal from the gasifier-shift syngas units, the physical solvent process was deemed to consume less utilities. Selexol was used for this design as being representative of proprietary physical solvent processes. The syngas has little methane and hydrocarbons. The reduction of vent gas loss was a design consideration. The utilities specified by the licensor were used for the efficiency calculation, capital cost and economics.

A low vent loss, resulting in less than 0.1% reduction in overall efficiency, is expected; a carbon monoxide level of less than 200 PPMV is predicted in the vent gas.

The hot potassium carbonate system is used in the conceptual design for removal of carbon dioxide from the Fischer-Tropsch recycle and net tail gas. Alcohols are absorbed in the process and are removed and recovered from the carbon dioxide vent stream.

The hot potassium carbonate process must be controlled to prevent the formation of formic acid salts.

18.4.2 CHAR REMOVAL

The quantity of char entrained in the syngas from the gasifier is large. The design char removal is set for a quantity approximately equal to the weight of the coal feed. This must be removed because ash can poison the catalyst in the Fischer-Tropsch synthesis. Steps are as listed below for 15,000 TPD char and dust removal:

| | Removed | Outlet Loading | |
	%	Char Lbs/Hr	Grains/100 SCF
Cyclones	97.7	29,250	
Dry electrostatic precipitator	99.5	146	
Venturi water scrubber	97.9	3	
Wet electrostatic precipitator	97.5	0.075	0.001

17.4.3 METAL CARBONYL

Iron-carbonyl, as well as nickel-carbonyl, may be formed from the reaction of carbon monoxide. Formation is favored by low temperatures and elevated pressure. While iron carbonyl has been detected in gas from low-pressure distribution systems, the time of contact in the Fischer-Tropsch plant is too short to form any appreciable amount. Iron-carbonyl does not appear to be a problem in past Fischer-Tropsch plants.

Semi-commercial experience for a fixed-bed gasifier has detected nickel-carbonyl in low-Btu fuel gas. Concentration of 0.04 PPMV have been reported in the raw syngas train. This carries through a bed shift reactor and Benfield treating unit with a gas containing about 25% carbon monoxide. Reducing the carbon monoxide to about 7.5% reduced the nickel-carbonyl to 0.002 PPMV.

The Fischer-Tropsch complex produces SNG with 0.1% carbon monoxide. The equilibrium and actual concentration of metal-carbonyls in the SNG should be below detection.

Nickel-carbonyl and iron-carbonyl can be formed by contact of cooling syngas and entrained ash as these are cooled by exchangers in the

gasifier outlet. These should be removed to a concentration manyfold below the sulfur level of 0.1 PPMV from the acid gas removal unit. The carbonyls may be considered a small part of the usual contaminants fed to the Fischer-Tropsch synthesis. The carbonyls formed in the Fischer-Tropsch synthesis circuit are not expected to cause a serious problem in the methanator. The carbonyls from the methanator are removed by equilibrium as the carbon monoxide is reduced to extremely low levels.

18.4.4 HYDROGEN CYANIDE AND AMMONIA

These impurities, tar and organic sulfur appear in very low levels in the syngas from an entrainment gasifier because of exposure to higher reaction temperatures than experienced in fixed or fluidized bed gasifiers. Hydrogen cyanide and ammonia are removed by water scrubbing of the syngas and reduced to undetectable levels in the acid gas scrubbing.

18.5 FISCHER-TROPSCH SYNTHESIS

18.5.1 DESIGN OF REACTORS

The conceptual reactors for the Fischer-Tropsch synthesis, shift conversion, and methanation are based on ERDA-developed flame-sprayed catalysts. Flame sprayed catalyst is also used for the water gas shift reaction.[8] This concept was compared in several mechanical models, and a design with coolant in the tubes and catalyst plates outside the tubes was selected. This gave the lowest pressure drop and lowest expected cost. Details of the background experimental work and run data used as the basis for this design is included as Section 15 in this report.

ERDA has directed their experimental work to improve the commercialization of the Fischer-Tropsch process by reducing the quantity of catalyst by using high space velocities while improving per pass yields of products. Concepts of hot gas recycle cooling and oil slurry reactors were not promising because of high energy use and poor heat recovery.

The advantages of catalyst plates outside the tubes are:

(1) High-pressure steam generation in the tubes is safer and uses less metal.

(2) More volume is available outside the tubes with lower gas pressure drop while heat transfer rates can be varied to provide optimum rates.

(3) Long tubes can be used with small diameter and thin tube wall. For a large reactor, catalysts can be more easily changed outside the tubes.

Therefore, the design of the conceptual plant features catalyst flame sprayed on finned tubes. Sprayed catalyst surface is approximately 14 times the bare surface tube area. The cooled tube is close to any part of

the catalyst fin to transfer heat both through the fin to the tube or by mixing and heat transfer in the gas phase. By using coolant tubes in close contact with the reaction gas, a minimum of gas recycle for cooling is required. Figure 18-5 shows the internal configuration of the reactor unit.

Another design considered was the sprayed-plate reactor with all heat removed by hot gas recirculation. The added power to circulate the recycle gas at up to 27:1 ratio consumed about six times the amount of power as required by direct cooling by fin tubes.

Placing the plates in the tubes reduces the catalyst-surface-to-bare-tube ratio to a typical 1.27:1 for some ERDA pilot plant designs. The plates in the tube passages reduce the flow area of the gas considerably and the resultant higher pressure drop increases power requirements. Much of the ERDA pilot plant runs are with synthesis gas flow through the tubes, which function well at low gas-flow rates. The conceptual design should have better heat transfer rates.

In addition to catalyst-sprayed-on-plates, steel lathe turning catalyst was investigated. Pilot plant run data was not available for high space velocity, but this catalyst may approximate the performance of sprayed-on-plate catalyst, while requiring more surface area and higher pressure drop. Lathe turning catalyst is considered an alternate means of providing an open catalyst system.

Catalyst in pellet form gave high gas-pressure drop with lower space velocity and higher hydrogen-to-carbon monoxide ratio required. These requirements made pellet catalyst less attractive than sprayed catalyst or lathe turnings.

18.5.2 PACKED TUBE AND ENTRAINED CATALYST REACTOR CONSIDERATIONS

SASOL used fixed and entrained catalyst (Synthol) reactors for the Fischer-Tropsch process. This entrained catalyst reactor produces a higher yield of oxygenates and light unsaturates than the fixed-bed reactor. In the entrained reactor, the reaction is accomplished in the transfer line to the separator. Coolers are in the line, while a large vessel provides separation of the catalyst from the gas stream. Catalyst separates and settles in the vessel with entrained fine catalyst further separated by cyclones. Settled and collected catalyst is recirculated.

Carthage Hydrocol at Brownsville, Texas used a true field reactor with cooling coils embedded in the fluid catalyst bed, and syngas flowing through the bed. The heating surface required appears to be great.

The packed tube reactor used at SASOL relies on heat transfer from the center of a catalyst containing 2-inch tube. This is satisfactory at low space velocity where large catalyst volume is provided. Heat is conducted from the catalysts by a flow of recycle gas, in turn cooled by the tube walls

with steam generation cooling. It was estimated that there is a temperature difference of 20°F between the center of the catalyst and the tube wall. This type reactor requires a higher recycle ratio, 2.5 to 1, in order to achieve the similar degree of reaction as the coated-fin tube system operating at a 1.5 to 1 ratio. Screening-type economic comparisons were made between the SASOL-type packed tube system, Synthol entrainment system, and selected coated fin-tube system using the limited published information available. An equipment and capital cost comparison is shown in Table 18-3. The economic comparison for these three systems is shown in Table 18-4.

The entrained catalyst Synthol reactor uses a large circulation of catalyst with coolers in the circulated gas and catalyst stream. Flow rates as high as 14 feet-per-second are used for the sprayed-plate reactor. This rate would wear the cooling tubes, but a rate of 2 to 4 feet-per-second may be feasible and a 50 to 100 Btu/SF/°F/Hr heat rate possible. About 16% of the reaction heat is recovered in the coolers which circulates heating oil. About 84% of the heat is removed by charging cold recycle gas with little use of exchange heat recovery. If none of the reaction heat is recovered as high pressure steam a drop of 24% in efficiency would result, and this conceptual plant would have less than a 50% thermal efficiency. This is based on no heat recovery for all process reactors.

The particles of catalyst in the entrained or fluid reactor must be cooled by conduction to the surrounding gas. If particle sizes are large, the heat released per unit surface is high. Small particles are difficult to separate and handle. So, the hydrogen content of the gas is increased to moderate the reaction. The entrained reactor sees a combined fresh and recycle feed of 6.77:1 hydrogen to carbon monoxide ratio, while a packed tube reactor uses 1.72. The selected sprayed plate reactor hydrogen to carbon monoxide ratio is 2.17 (fresh feed ratio of 1.45 plus recycle).

Each entrained catalyst particle burns brightly fueled by the hydrogen-carbon monoxide reaction heat and cooled only by excess hydrogen. If the temperature increase is too high, carbon is formed and the catalyst deactivates. Also, the carbon can split the catalyst particle, as happened at Carthage Hydrocol and in the pilot plant at Louisiana, Missouri where a slurry system was used.

The sprayed catalyst reactor can produce more oxygenates if operated at 1000 psig. But the sprayed catalyst reactor is used because the thermal efficiency is higher by a wide margin over the entrained catalyst reactor.

The packed tube reactor could be used as an alternative to the sprayed plate reactor with lathe turning catalyst, higher recycle ratio and larger catalyst volume. Or, pellet catalyst can be used with possibly 15% efficiency reduction because of lower steam generation pressure and increased recycle compression loss.

If methanation is not used after the sprayed catalyst Fischer-Tropsch reactor, an entrained catalyst reactor could be used to continue the Fischer-Tropsch reaction to high conversion, approximately 97%. Cryogenic separation of tail gas would be a problem that the conceptual plant has avoided by producing usable SNG without cryogenic separation.

The ERDA pilot plant research fluid methanation is based on micron-size catalyst requiring very large industrial reactors. Less active, non-nickel catalyst is used at high temperatures. Conversion is low whereas, SNG requires near 100% conversion of carbon monoxide to 1/10-% level in final product gas. If catalyst size of 1/32 to 1/16-inch were used, duplicating petroleum refining fluid techniques, highly active catalyst such as commercial nickel methanation catalyst on molecular sieve carrier, then an entrained catalyst system could be used as SASOL. This would only be for final methanation with the bulk methanation done in sprayed catalyst reactors with higher heat recovery efficiency. Then the entrained catalyst techniques might be acceptable, as well as a fluid reactor.

Petroleum fluid reactors are used to greatest advantage in the catalyst regeneration step in non-hydrogenation processes. Fluid hydrogenation has been tried but with poor results and no known plants are operating. Hydrogenation and hydrocracking processes trend to fixed bed and very active catalyst.

The use of less active catalyst than nickel for higher methanation temperature is not necessary for generation of high-pressure steam if a sprayed catalyst reactor with direct removal of heat in catalyst-coated fin tubes is used. If the reaction heat is removed with recycle gas, or entrained catalyst circulation, the steam generation pressure is degraded or the heat exchange surface vastly increased.

The conceptual plant features a sprayed catalyst reactor where the plates are intimately cooled by heat exchange tubes generating high-pressures steam or Dowtherm vapor. The plates are similar to plate fins in commercial heat exchange and are designed for periodical change of catalyst by similar technique to tube cleaning now practiced, or by removal of the plates for re-spraying.

18.5.3   FISCHER-TROPSCH SYNTHESIS CATALYST

The catalyst proposed for the Fischer-Tropsch synthesis is the flame-sprayed fused iron catalysts, developed by the Bureau of Mines at Bruceton, Pennsylvania, now the Energy Research and Development Administration's Energy Research Center. The catalyst is prepared from relatively inexpensive iron ores by flame-spraying the powdered ore onto carbon steel support plates to a thickness of about 20 mils.

A variety of iron ores may be used to prepare the active catalyst provided that they do not contain elements such as sulfur, which is a catalyst poison. The ERDA Fischer-Tropsch pilot plant has tested sprayed catalysts prepared from synthetic magnetite (Fe_3O_4), and Alan Wood magnetite. Tests are planned for plate catalysts prepared from Raney Iron.

The alkali metal content of the catalyst is important in determining both the catalyst activity, and the catalyst selectivity towards the synthesis of liquid hydrocarbons boiling higher than propylene.

The sprayed catalyst is impregnated with a potassium salt, potassium hydroxide or carbonate, before being charged to the reactor. The optimum quantity of potassium to be added is in the order of 0.4 parts of K_2O per 100 parts of active iron.

The desirable form of the active iron for the Fischer-Tropsch synthesis is considered to be Hagg carbide ($Fe_{20}C_9$). Before the catalyst is placed on stream it is reduced in-situ with pure hydrogen and the reduced iron then carbided by the syngas.

18.5.4 CATALYST POISONS

Sulfur, compounds, cyanides, coal ash, and halogens are known catalyst poisons for the Fischer-Tropsch synthesis, and, in minute quantities, rapidly deactivate the catalyst. The sulfur content of the syngas must be maintained below 0.1 ppm by volume. In the conceptual plant, the Fischer-Tropsch reactors are protected by a zinc oxide guard reactor which reduces the sulfur content to less than 0.05 ppm. The success of the conceptual design depends on removal of sulfur compounds and contaminants to low design level. One year catalyst life is expected. The presence of nickel on the catalyst results in low liquid yields since it promotes methanation reactions. Nickel carbonyl, the source of nickel, is removed in the acid gas treating step.

Water and carbon dioxide reduce catalyst activity. These reaction products are continuously removed to levels of 7%. Deactivation is slight and temporary at these low levels.

18.6 LIQUID PRODUCT RECOVERY

18.6.1 LIGHT ENDS RECOVERY

A combination of refrigeration and oil absorption has been used in the conceptual plant to recover some 60% of the propane and propylene from the net Fischer-Tropsch gas stream. These components are also recovered from the net Fischer-Tropsch liquid. Total recovery of these components from the Fischer-Tropsch net product is 63.7%, and was determined by the requirement for a light hydrocarbon stock for SNG blending, and the necessity to limit the amount of hydrogen consumed by hydrocracking in the methanation unit. The selected scheme was chosen after a preliminary review of several recovery schemes. These are:

Refrigeration System
Refrigerated Gas Fractionator
Refrigerated Lean Oil Absorption
Cryogenic Turbo-expander
Absorption.

18.6.2 PRODUCT FRACTIONATION

The recovered hydrocarbons and Fischer-Tropsch liquids are fractionated to give a mixed C_4 LPG, light and heavy naptha cuts, a wide-range diesel oil corresponding to ASTM No. 1 diesel, and a heavy fuel oil.

Propane and lighter hydrocarbons are freed of carbon monoxide and used to increase the calorific value of the SNG product.

Process stream temperatures have been limited to a maximum of 450°F to permit the use of process steam reboilers on fractionators.

18.7 METHANATION

18.7.1 DESIGN CONCEPTS

The gaseous fuel produced by the conceptual plant is designed to have combustion characteristics similar to those of natural gas. Natural gas is distributed over a wide area for both domestic and industrial use. SNG produced should deviate as little as possible from the combustion characteristics of natural gas.

To reduce both the hydrogen and carbon monoxide content, the stripped Fischer-Tropsch tail gas is methanated to 0.1 vol % CO. The ethane and heavier content of the Fischer-Tropsch gas is hydrocracked to methane and olefins saturated in passing through the methanator. The hydrocarbon dew point of the gas is -100°F at 800 psia. Gas leaving the final methanator has an HHV of 907 Btu/scf.

The calorific value of the methanated gas is further increased by the addition of a light hydrocarbon stream which contains olefins. Olefins are saturated and hydrogen reduced by passage through a bed of hydrotreating catalyst. The resulting SNG has an HHV of 1035 Btu/scf.

The gas hydrocarbon dew point is +32°F at 800 psia. The gas is dried by cooling and glycol contacting to prevent hydrate formation in the distribution lines.

SNG having a lower hydrocarbon dew point can be produced by recovering ethane from the Fischer-Tropsch gas and blending with the methanated gas. It is estimated that the additional utility demand for ethane recovery will decrease the percentage plant efficiency by about 0.2.

The CO_2 content of the feed Fischer-Tropsch gas is reduced by hot-carbonate scrubbing in the Fischer-Tropsch synthesis unit. Conditions in the methanation plant are designed to reduce the final SNG CO_2 content below 1.5 vol%. Recovering CO_2 before the liquids recovery and methanation units, rather than after methanation, reduces the size of these units by about 7%.

The nitrogen content of the SNG product is 6.8 vol %. Forty-eight percent of this nitrogen is derived from nitrogen in the coal and 52% from nitrogen in the oxygen feed to the gasifier if 98% purity is used.

18.7.2 METHANATOR REACTOR

The conceptual reactor for the methanation step is based upon the flame sprayed Raney nickel plate catalysts developed by PERC.

The Raney nickel catalyst is deposited on stainless steel finned tubes, which are first sandblasted on both sides with an iron-free grit and then flame-sprayed with a light coat of bonding material. The Raney nickel alloy powder is flame-sprayed onto the prepared steel surface to the desired thickness of about 20 mils. Direct removal of the heat of reaction through the catalyst-coated tubes provides for efficient recovery of almost all the reaction heat for subsequent use in generating 1200 psig steam. This reactor design is similar to that used for the Fischer-Tropsch reactors.

The catalyst is activated by leaching approximately 70% of the aluminum in the Raney alloy with a dilute caustic solution. An alternative form of catalyst construction is one in which the catalyst support plates are made of Raney nickel alloy, and the catalyst formed by activation of the plate surface only.

The conceptual plant methanator employs a recycle of 1.7 volumes of cooled reaction product for each volume of $(CO+H_2)$ in the feed gas. This recycle is used to moderate the rate of reaction by diluting the fresh feed CO concentration, and to control the concentration of water in the reactor outlet. The concentration of water in the reaction product is controlled to limit the amount of CO_2 formation in the reactor.

The fluid bed methanator, which applies fluid-bed cooling technology to methanation, was also considered. The low conversion level of 51 to 71 typically achieved by this system, and the high-operating temperature of 455 to 510°C, did not favor this reactor design over the plate reactor proposed.

18.7.3 FINAL METHANATOR

The final methanator is designed to reduce the methanator reactor effluent CO content from a maximum of 1 mol % to 0.1 mol % on a dry basis.

The reactor uses a fixed bed of granular nickel based catalyst, and is of a radial flow design for low-pressure drop. Commercial nickel-methanation catalyst is active from about 300 to 750°F, and does not generally lose activity upon exposure to temperatures up to 1500°F. The catalysts are also very active for methanating CO_2 to form methane and water. The CO_2 content of the reactor effluent is reduced to 1.29 mol % (dry basis), when the reactor is operated at the design inlet temperature of 520°F, and at a space velocity of 3370 SCFH/CF.

Carbon forming reactions are not catalyzed by commercial granular methanation catalysts under normal operating conditions.

18.7.4 HYDROTREATER REACTOR

In order to increase the calorific value to the SNG to 1035 Btu/scf, a mixed light hydrocarbon stream is blended into the gas. The mixed stream passes through the hydrotreater reactor where hydrogen in the gas stream is used to saturate olefinic compounds present. Saturation of the olefins, and the corresponding reduction in hydrogen concentration, serves to improve the combustion characteristics of the gas.

The reactor uses a fixed bed of granular cobalt-moly catalyst and is of radial-flow design for low pressure drop. Commercial cobalt-moly catalysts are active for olefin hydrogenation from 450°F to 800°F. The conceptual plant reactor is designed to operate at an inlet temperature of 500°F and a space velocity of 2000 SCFM/CF.

18.7.5 GAS DRYING

The SNG is dryed to a water dew point of 34°F at 1000 psig by a triethylene glycol (TEG) dehydration system. This is equal to a dew point depression of 88°F. TEG regeneration units using vacuum regeneration and/or gas stripping are able to achieve dew point depression in excess of 100°F.

Table 18-1 – Comparison of Entrainment Gasifier with Lurgi and Fluidized Bed Types
(One Train)

UNITS			LURGI-WESTFIELD			FLUID BED – SYNTHANE			ENTRAINMENT		
Differing Operating Costs	Cost/Unit	Units	Units/SD	EUAC** $MM/yr	$/MMBtu	Units/SD	EUAC $MM/yr	EUAC $/MMBtu	Units/SD	EUAC $MM/yr	EUAC $/MMBtu
Raw Materials											
Coal	$15/ton	TPSD	15,824	78.3	0.95	15,741	77.9	0.94	15,000	74.2	0.90
Oxygen	$10/ton	TPSD	7,628	25.2	0.31	4,199	13.9	0.17	9,437	31.1	0.38
Steam	$ 2/MLBS	MLB/D	43,873	28.9	0.35	29,280	19.3	0.23	(24,000)	(15.8)	(0.19)
Credit for Tar and Oil	$12/BBL	BBL/D	5,684	(23.1)	(0.34)	2,907	(14.4)	(0.17)	0	None	None
Maintenance	5% of FCI										
Materials				6.3	0.08		3.7	0.04		3.5	0.04
Labor				3.1	0.04		1.8	0.02		1.8	0.02
Payroll Burden	35%			1.1	0.01		0.6	0.01		0.6	0.01
Plant Overhead	60%			2.5	0.03		1.5	0.02		1.4	0.02
Operating Labor*				--	--		--	--		--	--
Property Tax and Ins.	2.75% of FCI			5.8	0.07		3.4	0.04		3.3	0.04
G&A Overhead				2.3	0.03		1.9	0.02		1.5	0.02
Total Differing Operating Costs				125.4	1.53		10.6	1.32		101.6	1.24
Fixed Capital Investment			$MM 210	28.1	0.34	$MM 122.8	16.4	0.20	$MM 118.3	15.9	0.19
Working Capital				3.9	0.05		3.2	0.04		2.3	0.03
Taxes											
Income Tax	52%			20.5	0.25		13.1	0.16		12.2	0.15
Investment Tax Credit	10%			(2.5)	(0.03)		(1.5)	(0.02)		(1.4)	(0.02)
Subtotal				175.4	2.13		140.8	1.71		131.0	1.59
Less Entrainment				131.0	1.59		131.0	1.59			
Potential Savings with Entrainment				$44.4	0.54		$9.8	0.12			

*Operating labor assumed to be nearly similar in all cases.
**EUAC = Equivalent Uniform Annual Cost.

18-14

Table 18-2 - Economic Comparison
Fischer-Tropsch Entrainment Gasifiers
2-Stage vs. 1-Stage
Two Trains, Capacity = 500 Billion Btu/Da

Differing Operating Costs	Cost/Unit	QUANTITY 2-stage	1-stage	Difference 2-stage vs. 1-stage	SAVINGS EUAC $MM/YR	$/MM Btu
Raw Materials						
Coal TPSD	$15 per ton	29,034	33,306	-4,272	21.146	0.131
Oxygen TPSD	$10 per ton	18,874	25,426	-6,552	21.622	0.128
Steam MLBS/SD	$ 2 per MLBS	33,252	16,188	17,064	-11.262	-0.068
Compressor Turbine Steam MLB/SD	$ 2 per MLBS		1,361	1,361	0.898	0.005
Maintenance	5% of FCI					
Material					-0.320	
Labor					-0.158	
Payroll Burden					-0.054	
Plant Overhead					-0.128	
					-0.660	-0.004
*Operating Labor					---	---
Overhead					-0.323	-0.002
Property Taxes and Ins.	2.75% of FCI				-0.294	-0.002
Fixed Capital Investment						
Basic Gasifier $MM		12.300	2.600			
Char Recycle Auxil-		4.000				
iaries			3.000		-1.432	-0.008
Recycle Compressor		$16.300	$5.600	$10.700		
Working Capital					-0.420	-0.003
Income Tax					-1.744	
Investment Tax Credit					+0.128	
Total Tax						-0.010
Credit for Steam Produced MLBS/SD	$ 2 per MLBS		-15,712	15,712	-10.370	-0.063
Savings of 2-stage over 1-stage gasifier					$17.289	$0.104

*Operating labor assumed to be nearly similar for both cases.

Table 18-3 – Fischer-Tropsch Catalytic Reactor Types Comparison
500 MMM BTU/Day Plant 2 Trains
Capital Costs

Major Equipment	Packed Tube System		Synthol Entrainment System		Coated Extended Surface System	
	Physical Data	Costs, $M	Physical Data	Costs, $M	Physical Data	Costs, $M
F-T Reactor Recycle Ratio	2.5:1		2.0:1		1.5:1	
Gas Throughput Rate, ACFM	313,700		269,160		224,300	
SCFM	9,766,700		8,366,700		6,966,700	
Space Velocity	500				2,000	
Catalytic Reactors						
31-Catalyst Tube Units		158,159				
16-Entrainment Loop Units				103,094		
18-Extended Surface Units						69,794
Catalyst		9,500		5,000		12,600
Feed Exchangers						
Surface, Sq. Ft.	721,500	53,942	620,000	46,354	515,400	38,533
Compressors		2,595		2,114		1,612
Operating HP	24,300		11,000		8,250	
Gas Throughput, ACFM	244,000		179,500		134,600	
Columns		1,536		1,408		1,280
Vessels		9,835		9,015		8,195
Pumps and Drivers		216		216		216
Storage Tanks, Agitators		88		88		88
CO_2 Removal Equipment		99,740		95,206		90,673
TOTAL CAPITAL COST		335,611		262,495		222,991

Table 18-4 - Fischer-Tropsch Catalytic Reactor Types Comparison
Operating Costs (Equivalent Uniform Annual Costs)

*Annual Major Operating Costs Items	Packed Tube System		Synthol Entrainment System		Coated Extended Surface System	
	Physical Data	Costs, $M	Physical Data	Costs, $M	Physical Data	Costs, $M
Compressor Turbine Steam, M#/HR	243		110		83	
Steam Costs $2/MLB		3,850		1,742		1,320
Maintenance @4% of Capital Costs (40% Labor)		13,424		10,500		8,920
Taxes and Insurance @2-3/4% of Cap. Costs		9,229		7,219		6,132
Catalyst Replacement		9,500		5,000		6,300
Catalyst Life, Yrs	1		1		1	
Plant Overhead @60% of Maint. Labor		3,222		2,520		2,141
G&A Overhead 1-1/2% of Total Operating Costs		597		411		378
TOTAL OPERATING COSTS		39,822		27,392		25,191
Income Tax		31,976		24,931		21,218
Fixed Capital Investment		44,931		35,143		29,854
Working Capital		1,355		987		882
TOTAL COSTS		118,084		88,453		77,145
Credit: 1250 PSI Steam Produced, M#/HR	5,519		6,209		5,956	
Value @ $2/MLBS		-87,421		-98,351		-94,343
TOTAL NET COSTS		30,663		-9,898		-17,198

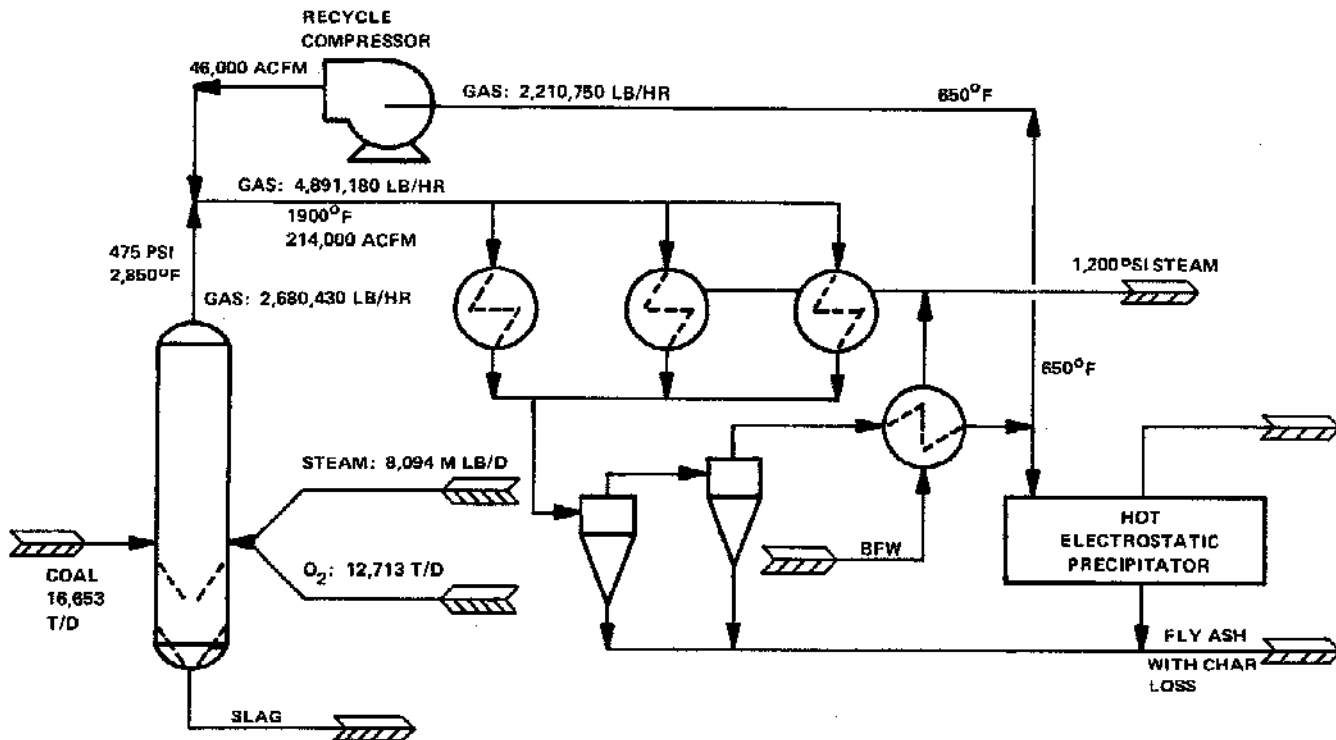*Operating labor assumed equivalent for all cases.

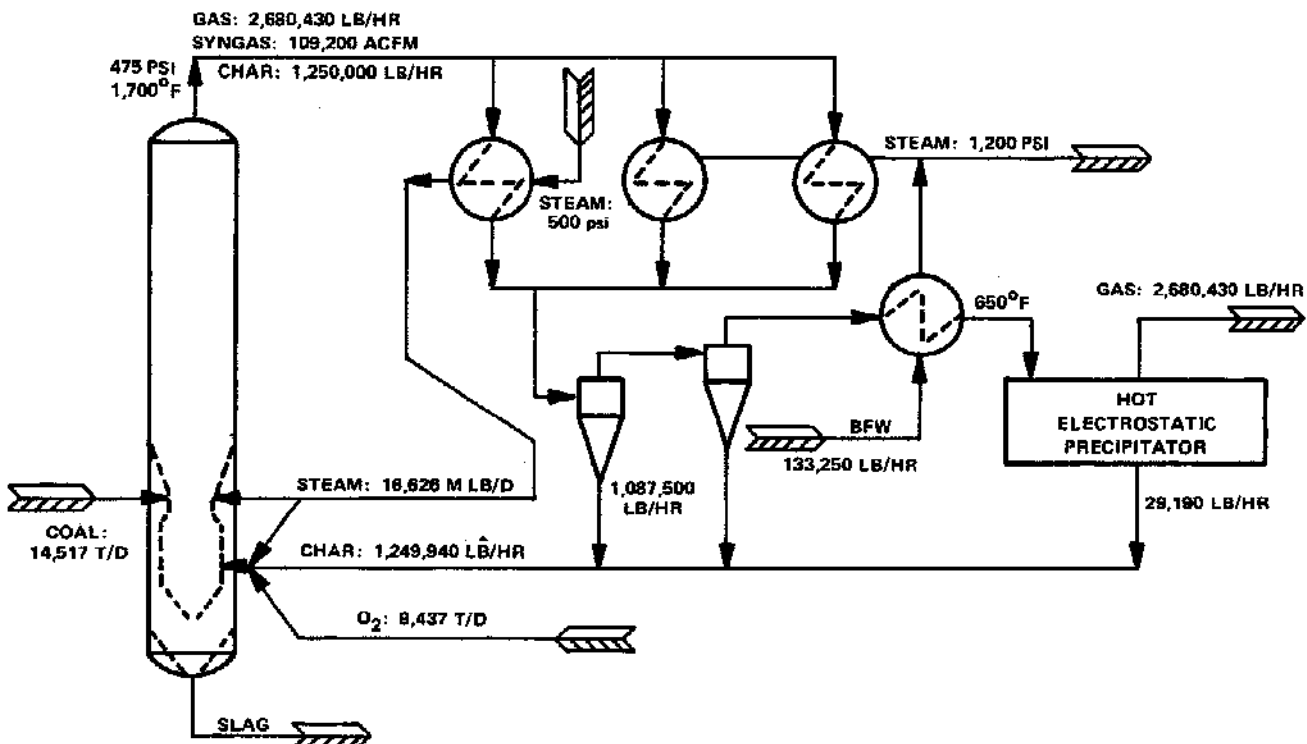Figure 18-1 - Single-Stage Entrainment Gasifier Flow Diagram (One of Two Trains)

Figure 18-2 - Two-Stage Entrainment Gasifier Flow Diagram (One of Two Trains)
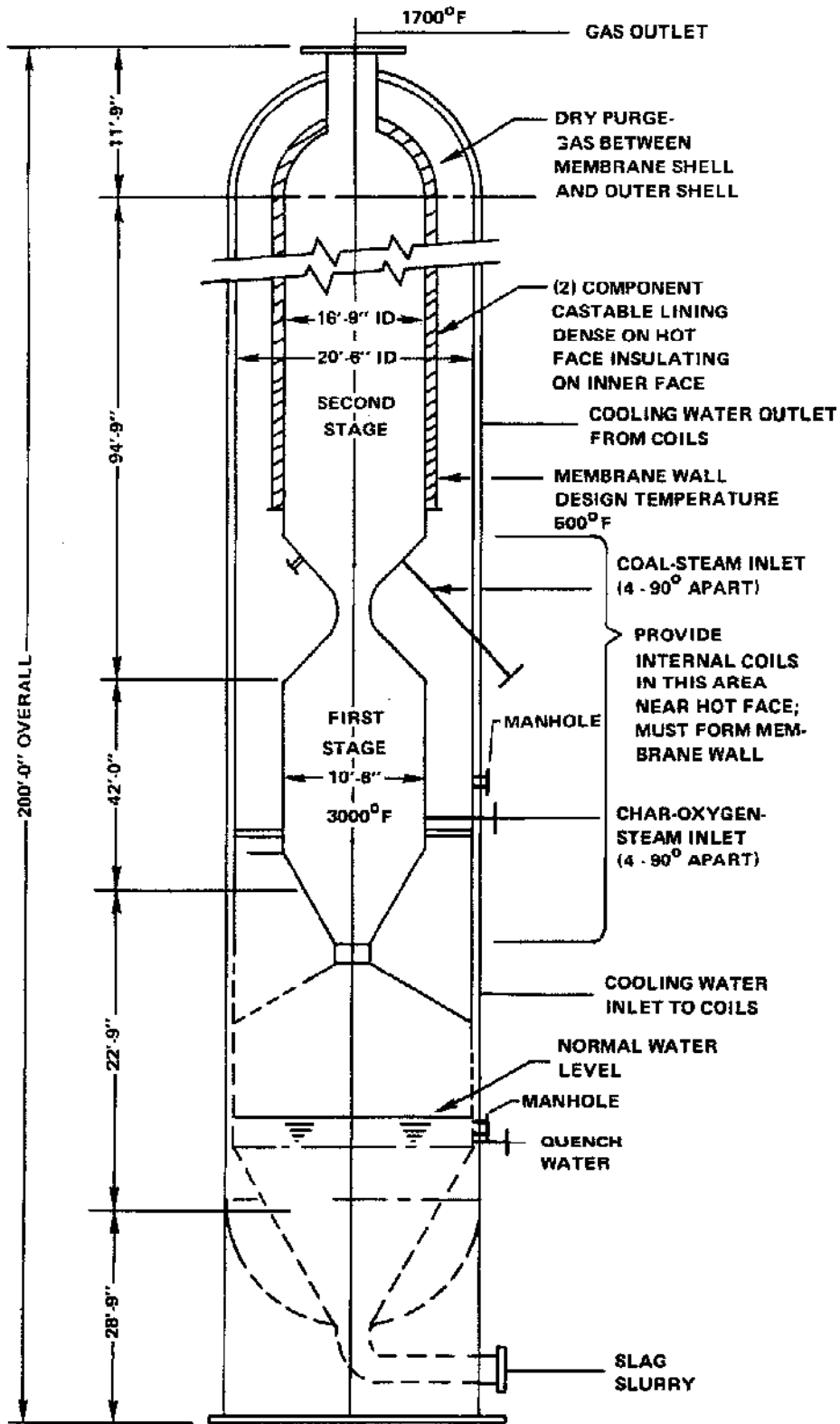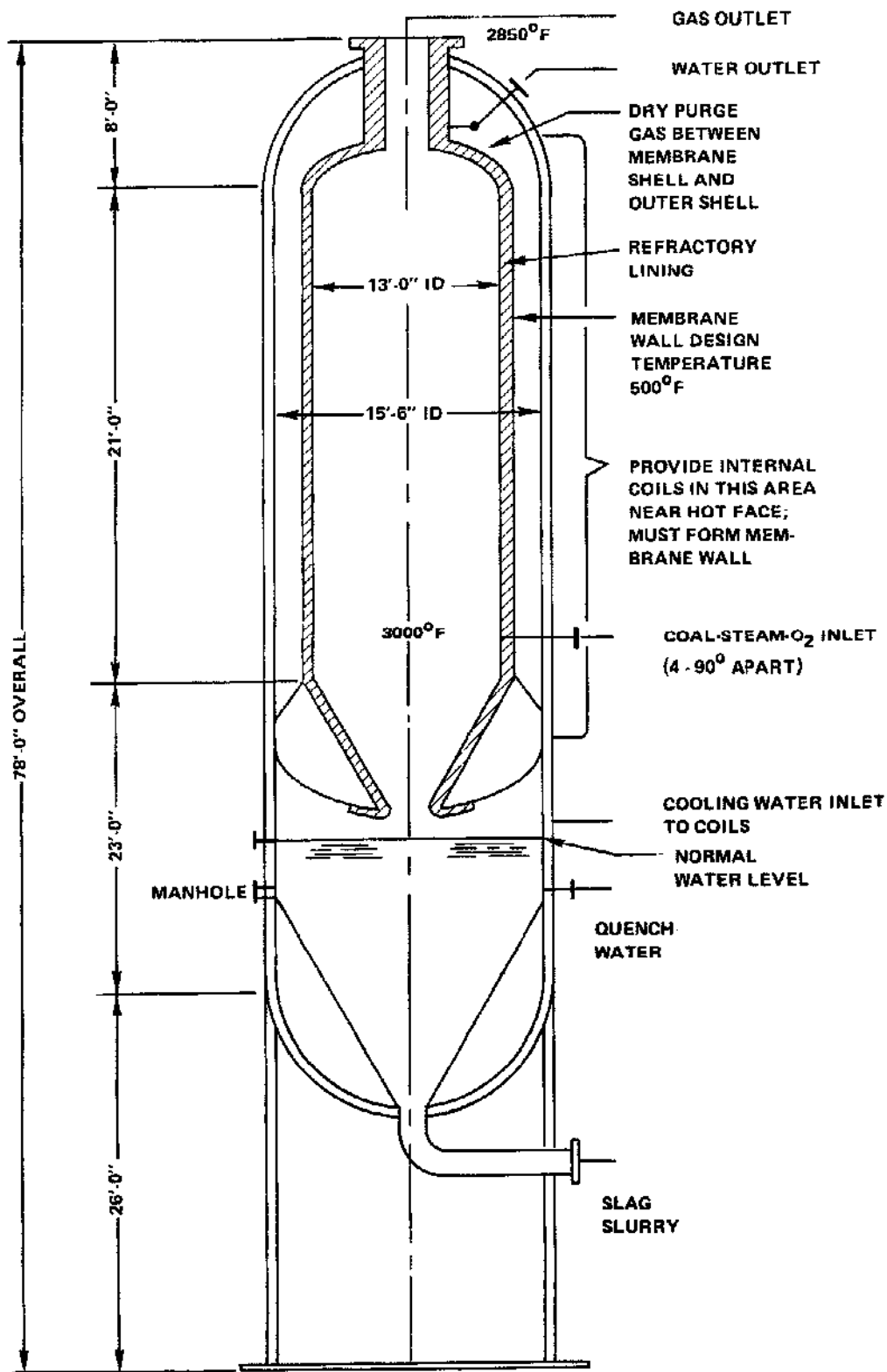
18-18

Figure 18-3 - Two-Stage Gasifier Design Concept

Labels in figure:

1700°F — GAS OUTLET

DRY PURGE-GAS BETWEEN MEMBRANE SHELL AND OUTER SHELL

(2) COMPONENT CASTABLE LINING DENSE ON HOT FACE INSULATING ON INNER FACE

COOLING WATER OUTLET FROM COILS

MEMBRANE WALL DESIGN TEMPERATURE 500°F

COAL-STEAM INLET (4 - 90° APART)

PROVIDE INTERNAL COILS IN THIS AREA NEAR HOT FACE; MUST FORM MEMBRANE WALL

CHAR-OXYGEN-STEAM INLET (4 - 90° APART)

COOLING WATER INLET TO COILS

NORMAL WATER LEVEL

MANHOLE

QUENCH WATER

SLAG SLURRY

11'-9"

94'-9"

42'-0"

22'-9"

28'-9"

200'-0" OVERALL

16'-9" ID

20'-6" ID

SECOND STAGE

FIRST STAGE

10'-8"

3000°F

MANHOLE

Figure 18-4 - Single-Stage Gasifier Design Concept

Figure 18-5 - Reactor

SECTION 19

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APPENDIX A

FISCHER-TROPSCH PLANT

CONCEPTUAL 500 MMM BTU/DAY

COMMERCIAL PLANT DESIGN BASIS

I. DESIGN OBJECTIVES

The objective is a conceptual commercial U.S.A. Fischer-Tropsch plant preliminary design, requiring a minimum capital investment, and resulting in minimum production costs per unit of product. The preliminary design is to be in sufficient detail for a preliminary fixed capital investment with target accuracy of -5 + 20% estimate and profitability analysis.

Major products will consist of liquid motor-fuel range hydrocarbons and SNG. The ratio of liquid products to SNG will approximate 1 to 1, based on product Btu values.

II. DESIGN PARAMETERS

1. Design Capacity: approximately 500 billion (MMM) Btu/day in liquid and gas products.

   - The plant will be comprised of two parallel trains with capacity of approximately 250 MMM Btu/day each.

2. Site Location: Eastern region of U.S. Interior Coal Province.

3. Coal Feed: See Section IV-A for typical analysis.

   - Requirements of cleaned coal are expected to be of the order of 40,000 tons per day.

4. Coal will be produced in a captive mine.

5. Products:

   - Liquid products will consist of LPG, naphtha, premium fuel and oxygenates.

   - SNG of pipeline quality with heating value in the range of 925 to 975 Btu/scf.

   - By-product sulfur will be produced.

● Product quantities are expected to be of the order of:

LPG	13,000 bbls/day
Naphtha	31,000 bbls/day
Premium Fuel	11,000 bbls/day
Oxygenates	3,000 bbls/day
SNG	250,000 scf/day
Sulfur	1,000 tons/day

III. FISCHER-TROPSCH PROCESS PLANT DESCRIPTION

A. General

This section describes current thinking regarding major equipment and process elements of the plant. Changes will be made during the design development as appropriate to achieve the stated objectives. A block flow diagram, Figure A-1, depicts tentative processing sequence. All effluents will meet environmental standards.

B. Dryer

Dryers will be used to remove free water and a small portion of the inherent moisture.

C. Grinders

Dried 1-1/4" x 0 coal will be ground to produce a gasifier feed.

D. Gasifier

This design will use the following features:

1. A steam-oxygen entrainment slagging-type gasifier will be used.

2. Coal will be fed to the gasifier by a dry feeder system.

3. Char will be recycled.

E. Gas-Solids Removal

1. The char solids entrained in the gasifier product gas are to be partially removed by such means as a heat recovery superheater and boiler deentrainment sections, high-efficiency cyclones, venturi scrubbers, electrostatic precipitators, wash columns, and final polishing filters.

F. Acid Gas Clean-Up

CO$_2$ and H$_2$S are to be removed by scrubbing with a physical solvent in a gas-treating unit. A clean CO$_2$ stream is discharged to the atmosphere. An H$_2$S/CO$_2$ mixture is directed to the sulfur recovery unit.

G. Shift Conversion

A shift conversion unit will be used to adjust the hydrogen to carbon monoxide ration.

H. Fischer-Tropsch Synthesis

The proposed reactors will consist of a pressure vessel expected to operate at about 400 psig with internal fintube coils. Exact pressure will be established during the design. Catalyst will be applied to the fins and external tube walls. Heat developed by the exothermic reaction will be transferred to boiler feed water inside the tubes to produce high-pressure steam.

I. Fischer-Tropsch Products Recovery

This will provide water scrubbing to remove oxygenates and carbonate wash to remove organic acids. The liquid hydrocarbon is to be fractionated to separate premium fuel oil and naphtha. The gas product is scrubbed to remove hydrocarbons and carbon dioxide. A mixed LPG cut is made. Alcohols and oxygenates are removed by water and recovered by distillation. The organic acids will be produced as salts and dried. All products contain nil sulfur, particulates, and nitrogen contaminates.

J. Methanation

The gas will be methanated, dried, and compressed for delivery at 1000 psig.

K. Sulfur Recovery

Sulfur will be recovered from the process and utility treating units' H$_2$S gas effluent.

L. Waste Heat Recoveries

Waste heat will be recovered in the form of steam for use within the plant as process feed, for machinery turbine drives and for power generation. Waste heat recoveries will be maximized in order to achieve optimum overall efficiency.

M. Utilities

All utilities will be generated on site.

N. Oxygen

Oxygen will be captively produced.

IV. FISCHER-TROPSCH PROCESS DESIGN CRITERIA

A. Coal Properties

Cleaned, sized and dried coal with the following typical analysis:

a. Proximate analysis (wt. %) (Clean dry coal)

Moisture	2.7
Ash	7.1
Volatile matter	38.5
Fixed carbon	51.7

Heating value 12,600 Btu/lb.

b. Ultimate analysis (wt. %)

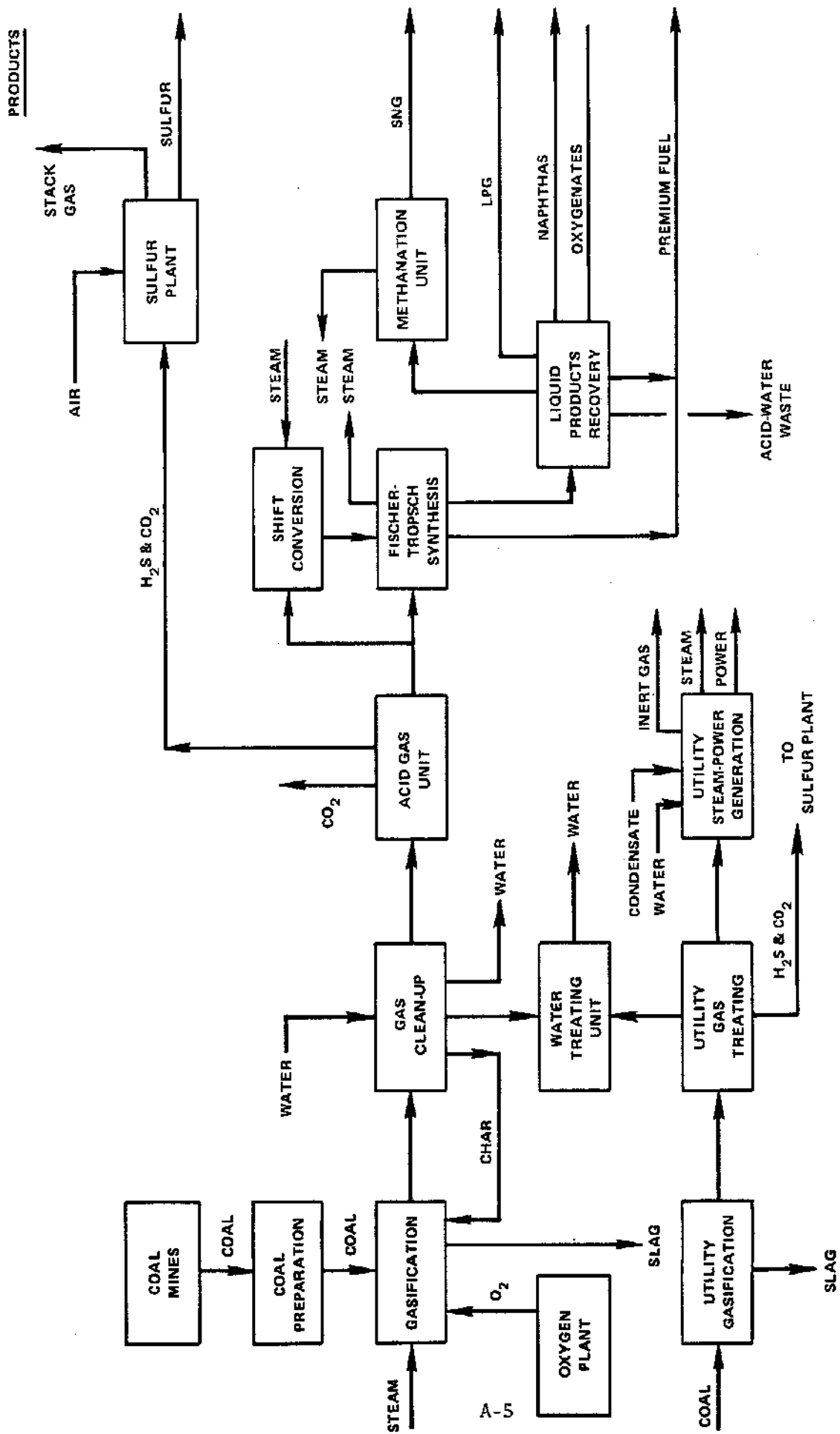|	
Carbon	70.7
Hydrogen	4.7
Nitrogen	1.1
Sulfur	3.4
Oxygen	10.3

Figure 1 - Flow Diagram, Fischer-Tropsch Process

A-5