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ENVIRONMENTAL CONTROL TECHNOLOGY FOR THE FLASH HYDROLYSIS OF COAL

PROGRESS REPORT NO. 3
APRIL 1, 1978 TO DECEMBER 31, 1978

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U.S. Department of Energy
Washington, D.C. 20545

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ABSTRACT

Based on experimental results and analytical studies, preliminary environmental and economic assessments of a conceptual, industrial scale FHP-Refinery Complex have been made. Three cases were considered, one for the production of pipeline gas, the second for the production of liquid hydrocarbons such as motor gasoline and LPG, and the third for the production of both liquid and gaseous products. Solid, liquid and gaseous effluents for each case have been evaluated, and required effluent treatment facilities and potential problem areas identified. Thermal efficiencies and investment costs have also been estimated. Net thermal efficiency ranges from 50% to 72%, depending upon the mode of operation of the Complex, and investment costs from 834 to 936 million dollars. The installed cost of environmental equipment is estimated at about 5% of total plant investment for all three cases.

INTRODUCTION

Intensive efforts to produce gaseous and liquid fuels from coal, a plentiful domestic resource, are currently in progress. The Flash Hydro-pyrolysis of Coal Process (FHP) is one of several related processes for producing gaseous and liquid fuels by gas-phase hydrogenation of coal. The liquid product is suitable for conversion to distillate fuels and chemical feedstocks and consists primarily of aromatic hydrocarbons (BTX and other aromatics); the gaseous products are mainly methane and ethane. The gas to liquid product ratio is dependent upon reactor operating conditions, and generally, higher liquid yields are obtained as operating pressure is increased, while higher gaseous yields are obtained as operating temperature is increased. Reactor residence time also has an important effect on conversion.

In order for any developing coal conversion process to reach industrial fruition, we must have reasonable assurance that it will operate in an environmentally acceptable manner. Accordingly, all significant input, product and effluent streams must be defined and examined to determine the necessity for treatment and the suitability of existing control technology or the need for new technologies. By performing environmental reviews and assessments concurrent with process development, commercialization will be realized sooner than if done sequentially and, the comparative merits of related processes, from an environmental standpoint, will be identified at an early date. Thermal efficiency is also important inasmuch as it is an indication of the quantity of raw fuel consumed in producing the clean fuel product.

Considerable help in preparing this study was obtained from Burns and Roe Industrial Services Corporation (BRISC), and in particular from Mr. Quinton Lee.

A preliminary environmental assessment of an FHP-Chemical Complex appears in Progress Report No. 2, BNL 50906, May 1978.

PROCESS DESCRIPTION

In this study, three cases were investigated. One for the production of pipeline gas, referred to later as gasification; one for the production of motor fuel and LPG, referred to as liquefaction; and one for the production of both gaseous and liquid products, referred to as mixed products. In all three cases, the process design is based on data obtained from the BNL experimental FHP unit (Tables 1 and 2).

The block diagram of the gasification plant is shown in Figure 1. The plant combines the FHP process with cryogenic separation of methane and recycle hydrogen. FHP operating conditions for gasification are 1517°F and 2500 psig. The plant is capable of producing 395 MMSCFD of pipeline gas for export.

The block diagram of the liquefaction plant is shown in Figure 11, and the mixed-product case diagram in Figure 24. These plants consist of the FHP process, cryogenic separation of gaseous products and recycle hydrogen and a conventional refinery scheme. The FHP operating conditions are 1382°F and 2000 psig. Liquid products are recovered in a quench tower and processed through the refinery complex. Both plants produce 47,700 BPD of 90 RON (Research Octane Number) motor gasoline and LPG. For the liquefaction case, hydrogen feedstock is generated by reforming the methane produced by

the process, whereas, the mixed products plant generates hydrogen by gasification of residual FHP char. Thus, the mixed-products plant produces methane as well as motor gasoline and LPG. The overall process diagram of the three cases described above (Figures 1, 12 and 24), have been broken down into more detailed schematics in order to clarify and further identify process effluent streams. The bracketed numbers appearing on the overall process diagrams correspond to the figure numbers of the more detailed schematics. The effluents of the three processes are summarized in Tables 3 through 11.

GASIFICATION

Coal Preparation and Feeding (Figure 2)

The principal effluents generated in preparing run-of-mine coal for FHP process use are: (1) a coal refuse stream produced by the crushing, screening and grinding of the raw coal, S_1 , (2) a gaseous effluent stream from the coal drying system, A_1 , and (3) a rain run-off stream from the coal storage pile, W_{13} . The coal feed system will also require and emit (stream A_2) a substantial quantity of inert gas. CO_2 generated within the complex, e.g., CO_2 produced in the hydrogen furnace by the combustion of CO, and CO_2 recovered from acid gas removal (Figure 4), can be used for both drying and feeding purposes. However, effluent CO_2 must be cleaned of entrained coal dust before being discharged to the atmosphere. Either an electrostatic precipitator or bag filter can be used for this purpose. If a bag filter is used, the gas must be maintained at a temperature above its dew point to prevent condensation of water on the filter bags. In addition, some volatiles may be present in the CO_2 drying effluent.

The pollution aspects of coal refuse disposal and of rain run-off are generally common to all coal preparation facilities and have been investigated by others; therefore, they will not be treated in detail in this study. However, it should be mentioned that significant amounts of trace elements are contained in coal refuse and that this refuse will have a higher concentration of these elements than was present in the raw coal before processing.³

Hydropyrolysis (Figure 2)

The coal feed to the hydropyrolyzer (FHP unit) is converted to gaseous and solid products by reaction with preheated hydrogen gas. The solid product (effluent) of the hydropyrolyzer is char which is fed to a BCR (BiGas) gasifier for hydrogen production. The raw gaseous FHP product is passed through a bank of multistage cyclones to remove entrained dust and then fed to an iron oxide unit for removal of H_2S . The gaseous stream is then split into a recycle stream to the FHP unit and a product recovery stream. The product gas is indirectly cooled by water, which is vaporized to steam, and then passed through a quench tower where it is contacted with recycled quench water. The condensables (BTX and oils) from the FHP product gas and the quench water are fed to an oil separator (Figure 3) while the non-condensed gases (e.g. CH_4 , C_2H_6 , CO , CO_2 , and H_2) are directed to a gas purification system (Figure 4). In the gasification mode, only a small amount of liquid hydrocarbons are produced.

Oil Recovery (Figure 3)

The mixture of FHP condensables and quench water is separated into an aqueous and an organic phase in an oil separator unit. The organic

phase (raw oil) is fed to the BCR gasifier (Figure 6). The bulk of the aqueous phase is returned to the quench tower for reuse and about 10% directed to a Phenosolvan unit where NH_3 , phenols and oxygenated compounds are recovered. The waste water effluent from the Phenosolvan unit, W_4 is directed to a Biox unit (Figure 8) for processing and subsequent reuse within the plant. The gaseous effluent of the Phenosolvan unit, A_5 , may require further treatment to remove H_2S and NH_3 .

Gas Purification (Figure 4)

The raw gas from the Quench Tower (Figure 2) is cooled in a series of heat exchangers to remove moisture, W_8 , and then fed to an MEA absorption unit for removal of H_2S and CO . Once concentrated, the $\text{H}_2\text{S}/\text{CO}_2$ stream is fed to a Claus Plant (Figure 7) for sulfur recovery. The liquid effluent stream from the MEA system, W_5 , containing degraded materials produced during MEA regeneration, is exhausted to a Biox unit (Figure 8) while stream W_8 is fed to a waste water treatment system (Figure 8). The clean raw gas is then passed through a dryer for further moisture removal, W_6 , before being treated in a copper wash system (Cosorb). Stream W_6 is also exhausted to the Biox unit. The purpose of the dryer is to prevent the formation of hydrochloric acid in the Cosorb system which would otherwise occur by hydrolysis of CuAlCl_4 . The function of the Cosorb system is to remove CO from the raw gas before it is fed to the cryogenic separation system (Figure 5), otherwise (because of its low boiling point), it would remain in the system. CO recovered by the copper wash system is fed to the hydrogen furnace (Figure 1) for use as fuel. The liquid effluent of the Cosorb system, W_7 , is a mixture of toluene and CuAlCl_4 .

Gas Separation (Figure 5)

Purified gas from the Cosorb system (Figure 4) is fed to the gas separation system via a dryer. Alumina contained in the dryer reduces the water dew point of the feed gas to about -73°C . The gas separation system is essentially a combination of low-temperature condensation and absorption processes. In the gasification mode the bulk of the H_2 is separated from C_1 by condensing the C_1 at about -131°F . The purified hydrogen is recycled to the FHP unit while the product C_1 stream is compressed and sold as pipeline gas. The effluent streams of the gas drying and separation systems are spent alumina absorbent, S_3 , and vaporized nitrogen, A_6 , supplied as liquid refrigerant to the gas separation system by an on-site oxygen plant (Figure 1).

Hydrogen Production (Figure 6)

The char product of the FHP unit (Figure 2) is fed to a BCR (BiGas) gasifier along with water, steam, oxygen and coal, to produce hydrogen. The oxygen is supplied to the gasifier by an adjacent air separation plant. Liquid nitrogen coproduced by the plant is utilized in the gas separation plant (Figure 5), as described above.

The BCR (BiGas) gasifier consists of three separate stages: a slag quench zone; Stage 1, where char burners are located; and Stage 2, where coal feed is injected. The slag quench zone is located at the bottom of the gasifier and contains cooled recirculating water. As the molten slag from Stage 1 falls into the water, it is fractured to a granular form. The granular slag settles to the bottom of the zone where high pressure water jets agitate and move it to outlet nozzles.

The slag, S_2 , is then sent to lock hoppers and finally to disposal. Water used in quenching the slag is recirculated within the process except for a bleed stream which is drained to a waste pond following pressure let down. Dissolved gases contained in the slurry bleed are released during pressure let down and vented to a thermal oxidizer.⁴ The slag produced by the gasifier would be similar to blast furnace slag, but lower in sulfur, and may be suitable for fill, road aggregate or for addition to cinder blocks. Since the slag stream was initially a molten pool, it should contain a minimum amount of dust and leachable materials. However, a complete slag analysis and evaluation would be required before any firm conclusions can be reached.

The cooled raw gaseous product of the BCR gasifier is passed through a sand filter and then fed to a shift converter where additional high pressure steam is injected to adjust the steam-to-dry gas ratio to unity. The shift reaction is exothermic, therefore, the gases are heated in passing through the converter. On exiting the converter, the hot gases are cooled, fed to a knockout drum for removal of condensate (gas liquor), W_1 , and sent on to a Selexol solvent system for purification. The collected condensate is let down to a degasifier and condensate cooler (Figure 8).

Many of the trace elements contained in the coal and char feed to the gasifier will have appreciable volatilities under the operating conditions of gasification, and some compounds may be reduced to metals in the presence of carbon. These volatile trace elements are expected to concentrate in the gas liquor condensate; non-volatiles will concentrate in the BCR ash.

Following shift conversion and quenching, the hydrogen rich gas is fed to a Selexol solvent unit for acid gas removal. The acid gas (H_2S , CO_2 , COS , and H_2O) effluent of the Selexol process is combined with the acid gas effluent of the MEA unit (Figure 4) and fed to a Claus plant (Figure 7). Recovered CO_2 , stream A_3 , is vented to the atmosphere or used for coal drying (Figure 2).

The Selexol solvent process was chosen for the subject service because it is selective enough to simultaneously provide a product gas (H_2 rich stream) low in H_2S content (less than 1 ppm), a rich H_2S stream for Claus processing, and an essentially sulfur free CO_2 vent stream.

Minor emissions from the hydrogen production system include raw product gas leakage, for example, leakage through the coal feeding device, and a gasifier start up vent stream. The latter would normally be flared.

Sulfur Removal (Figure 7)

As described above, the combined stream of acid gas effluents from the Selexol and MEA units is directed to a Claus plant for sulfur recovery. Based on the composition of the Claus feed gas, an H_2S conversion efficiency of about 93% is expected. The sulfur recovered by the process would be quite pure and marketable. The tail gas from the Claus process, A_4 , is expected to contain SO_2 , COS , CS_2 and sulfur as well as unconverted H_2S . Depending upon air quality regulations, it may be necessary to treat the tail gas in an additional facility before it is discharged to the atmosphere. One available treatment option is to convert the tail gas to sulfur dioxide in an incinerator, and then direct the gas stream to the flue gas desulfurization (FGD) system

servicing the plant boiler; another, is to incinerate the tail gas in the plant boiler itself, in which case the treated tail gas would exit with the stack gases. FGD economics may or may not be affected.

Sulfur compounds in the Claus tail gas can also be eliminated by use of an absorption/oxidation type process such as Stretford, which converts H_2S and other sulfur compounds to free sulfur. The degree to which sulfur emissions are reduced will depend largely on investment and pollution abatement considerations. The equivalent SO_2 discharge attributed to COS and CS_2 ranges from 4,000 ppm for a Claus plant operating at 96% recovery to as high as 10,000 ppm for a Claus unit operating at 90% recovery. To reduce sulfur emissions to 250 ppm SO_2 or less, a substantial increase in the overall gasification plant investment would be required.

Water Purification (Figure 8)

Almost all process modules reject a liquid waste stream, however, all waste streams do not have the same physical, chemical, and biological properties. Thus, in some cases control may be more or less extensive than in others. Waste streams from some modules may be combined prior to treatment, or pretreated separately and then combined for further treatment and discharge.

The facilities required to treat the liquid effluents of the complex are shown in Figure 8. The treatment facilities include: (1) a degasifier (not shown) and condensate cooler, (2) a Phenosolvan unit, (3) a Biox unit, and (4) a waste water treatment facility. At present, the system is rather speculative and further development and evaluation are

required, however, the following general statements can be made.

Aqueous effluents such as cooling tower blowdown, W_{12} (Figure 9), the blowdown from the utility boilers, W_{11} (Figure 9), and process sewer effluents should not present any unusual problems, and would be routed to a waste water treatment system.

The gas liquor, W_1 , formed on condensing the volatiles contained in the shift conversion product gases (Figure 6), will probably contain a great many contaminants including some oil, phenols, cresols and nitrogen compounds - not only primary ammonia, but also cyanides, amines, thiocyanate, pyridines, etc. Trace elements vaporized during gasification as well as H_2S and other sulfur compounds may also be present. Specific methodologies will have to be developed to adequately recover and dispose of all of these in an environmentally acceptable manner. Furthermore, the leaching of salts and trace metals from such disposal residues must also be carefully evaluated. The situation bears some resemblance to the problem of acid mines water and such background experience may be helpful in evaluating and solving the problem.

The pollution aspects of rain run-off, W_{13} , were discussed briefly under coal preparation and feeding and have been treated in detail elsewhere.

Phenolic compounds can be extracted with an organic solvent such as isopropyl ether or oil. A sour water stripper can be used to remove H_2S , ammonia and light hydrocarbons. Other clean-up steps such as filtration to remove suspended solids and lime treatment to remove fluoride ions may also be required. A Biox unit is effective in treating

organics, nitrogen, phosphorous and sulfur compounds, however, trace metals are not effectively consumed or destroyed by biological action, and, heavy metals such as copper and chromium are highly toxic to the culture. In addition, Biox will not sufficiently clean-up refractory compounds such as benzene, alkylated benzene and naphthalene, therefore, follow-up treatment with activated carbon may be needed. The cellular material produced in the Biox unit, S_5 , can be disposed of by incineration.

Recovery and or disposal of the toluene- CuAlCl_4 effluent, W_7 , of the Cosorb system, Figure 4, must also be evaluated.

Utilities (Figures 9 and 10)

The anticipated utility modules required to service the subject complex (excluding drinking water, etc.) are shown in Figures 9 and 10, along with anticipated effluents. The utilities consist of two make-up water filtration and demineralization systems, utility boilers and cooling towers. The units would be standard items currently in use today and their effluents should pose no unusual problems. Utility boilers would require dust collection and SO_2 scrubbing systems (Figure 10) typical of those presently used by fossil-fuel power plants. Cooling towers would also be of standard design, however, cooling tower emission rates (of air and mist) and the potential problem of cooling tower drift loss may impose constraints on site selection. The air and mist flow from the cooling towers (2.5×10^6 tons/day), A_7 , is by far the largest stream released to the environment by the complex. It may be advantageous to use air cooling whenever possible to minimize the use of cooling water and

thereby reduce cooling tower load and emissions.

The production rates of effluent streams A_8 , A_9 , W_{11} , S_4 , S_8 , and S_9 have not been specified because the steam system for the complex has not been optimized as yet.

Energy Balance and Utility Requirements

A preliminary energy balance and a utility summary have been prepared for the subject Gasification Complex. Material and energy balance summaries appear in Table 12; summaries of the cooling water, electricity, steam and fuel (generated internally by the complex) requirements are presented in Table 13. The overall steam rate required by the complex is based on the nominal steam usage for each of the process units indicated in Figure 1. No attempt has been made to minimize steam utilization and neither heat pumps nor other mechanical energy recovery devices have been considered. Coal for steam and power generation has not been included in the material balance.

Economic Evaluation

The installed cost of the Gasification Complex is summarized in Table 14. The cost estimate is based on published modularized cost method and data and also on Burns and Roe's in-house cost files. While the estimate is only order-of-magnitude, it gives an indication of the relative cost of pollution abatement equipment and also of the relative costs between gasification, liquefaction and mix-products plants. The latter two are discussed below. An economic evaluation of the gasification plant is presented in Table 15.

LIQUEFACTION

A process block diagram of the liquefaction complex is shown in Figure 11 and a schematic of the ancillary refinery in Figure 12. The key to the process is the Flash Hydropyrolysis step. Here, 28,000 TPD of cleaned, sized lignite is fed into the FHP unit. The hydropyrolyzer operates at 2,000 psi and 1382°F, and produces about 14% wt hydrocarbon liquids, 8% C₂, and 22% C₁ based on MAF lignite. Hydrocarbon liquids are removed on quenching the product gas stream and light hydrocarbons are separated into C₁ and C₂ cryogenically.

BTX and light oil recovered from the oil stabilizer unit are sent directly to the oil refinery complex for motor gasoline production. C₂ is steam cracked in an olefin plant to produce ethylene and further processed to C₄ in a dimerization module. The bulk of the light oil produced and the excess BTX are sent to a hydrocracker unit to produce gas oil which is fed to a blending plant. Hydrogen is generated by methane-steam reforming.

Coal Preparation and Feeding (Figure 13)

The principal effluents that would be generated in preparing run-of-mine coal for FHP process use are: (1) a coal refuse stream produced by the crushing, screening and grinding of the raw coal, S₁, (2) a gaseous effluent stream from the coal drying system, A₁, (3) a gaseous effluent stream from the coal feed system, A₂, and (4) a rain run-off stream from the coal storage pile, W₁₃. The coal feed system will also require a substantial quantity of gas. Flue gas from the H₂ furnace and steam reformer is used for both coal drying and feeding purposes. The flue gas is discharged to a flue gas desulfurization system (Figure 23). Char and methane are used to fuel the H₂ furnace.

and reformer producing ash stream S_2 . Other environmental considerations are similar to those discussed under Gasification, above.

Hydropyrolysis (Figure 13)

The coal feed to the hydropyrolyzer (FHP unit) is converted to gaseous and solid products by reaction with preheated hydrogen gas. The solid product (effluent) of the hydropyrolyzer is char which is fed to the H_2 furnace and methane reformer for use as fuel. The raw gaseous FHP product is passed through a bank of multistage cyclones to remove entrained dust and then fed to an iron oxide unit for removal of H_2S . The gaseous stream is then split into a recycle stream to the FHP unit and a product recovery stream. The product gas is indirectly cooled by water, which is vaporized to steam, and then passed through a quench tower where it is contacted with recycled oily quench water. The condensables (BTX and oils) from the FHP product gas and the quench water are fed to an oil separator (Figure 14) while the non-condensed gases (e.g. CH_4 , C_2H_6 , CO , CO_2 , and H_2) are directed to a gas purification system (Figure 15).

Oil Recovery (Figure 14)

The mixture of FHP condensables and quench water is separated into an aqueous and an organic phase in an oil separator unit. The bulk of the aqueous phase is returned to the quench tower for reuse and about 10% directed to a Phenosolvan unit where NH_3 , phenols and oxygenated compounds are recovered. The waste water effluent from the Phenosolvan unit, W_4 , is directed to a Biox unit (Figure 21) for processing and subsequent reuse within the plant. The gaseous effluent of the Phenosolvan system, A_5 , may require further treatment to remove H_2S and NH_3 .

The raw oil from the oil separator is fed to an oil stabilizer unit which consists of a Udex or Sulfolane system to selectively separate BTX from product oils. Some heavy ends, W_3 , are discharged by the unit which may possibly be used as a source of fuel for the plant's utility boiler.

Gas Purification (Figure 15)

The raw gas from the Quench Tower (Figure 13) is cooled in a series of heat exchangers to remove moisture, W_8 , and fed to a Selexol unit for selective removal of H_2S and CO_2 . The H_2S stream is fed to a Claus plant (Figure 19) for sulfur recovery; the essentially sulfur free CO_2 stream is vented to the atmosphere. Stream W_8 is directed to a waste treatment unit (Figure 21) for treatment and reuse. The Selexol solvent (dimethyl ether of polyethylene glycol) is physically and chemically stable and is not degraded by impurities such as COS , NH_3 , HCN and aromatics which may be present in the raw gaseous product of the FHP unit.

The treated raw gas is then passed through a dryer for further moisture removal, W_6 , before being treated by a copper wash system (Cosorb). Stream, W_6 , is fed to a Biox unit (Figure 21). The purpose of the dryer is to prevent the formation of hydrochloric acid in the Cosorb system which would otherwise occur by hydrolysis of $CuAlCl_4$. The function of the Cosorb system is to remove CO from the raw gas before it is fed to the cryogenic separation system (Figure 16), otherwise (because of its low boiling point), it would remain in the system. The CO recovered by the copper wash system is fed to a shift converter (Figure 18), for conversion of steam to hydrogen. The liquid effluent of the Cosorb system, W_7 is a mixture of toluene and $CuAlCl_4$.

Gas Separation (Figure 16)

Purified gas from the Cosorb system (Figure 15) is fed to the gas separation system via a dryer. Alumina contained in the dryer reduces the water dew point of the feed gas to -73°C . The gas separation system is essentially a combination of low-temperature condensation and absorption processes in which the bulk of the C_2 and heavier are condensed out of the gas stream as the product gas is cooled. The C_2 is fed to an ethylene plant (Figure 17) along with some of the C_1 (for fuel); the remainder of the C_1 is fed to a steam reformer (Figure 13). The cooling medium for the gas separation system is provided by a separate refrigeration unit since the Liquefaction Complex does not contain an air separation plant as in the case of Gasification. The only effluent of the gas drying and separation system is spent alumina, S_3 .

Ethylene Plant (Figure 17)

The clean C_2 and heavier components recovered in the gas separation plant (Figure 16) are directed to an ethylene plant where they are cracked by steam to produce a host of products. Better than 83% of the C_2 feed is converted to ethylene, the balance to H_2 , C_1 , C_3 , C_4 and aromatics.

With ethane as the principal feedstock, the operation of the ethylene plant would be relatively clean and the environmental impact of the system would be low as compared with heavier feedstocks. The only anticipated effluent of the plant is sour water, W_9 , generated by blowdown of the ethylene plant quench tower. The sour water is directed to a waste treatment system (Figure 21) for processing and subsequent reuse.

Steam Reforming (Figures 11 and 13)

Part of the methane recovered in the gas separation unit (Figure 16) is fed to a steam reformer. The function of the reformer is to crack methane in the presence of steam to produce hydrogen. This is accomplished in a number of reforming tubes that are basically heat transfer tubes filled with catalyst. The tubes are oriented vertically in a direct-fired furnace box. On passing through the tubes the steam and methane react endothermally producing hydrogen and carbon monoxide. Methane is combusted in the furnace box supplying heat to the process. The flue gas (CO_2 , H_2O , O_2 , and N_2) can be vented to the atmosphere or used for coal preparation and feeding (Figure 13). The product gas from the secondary reformer (upper section of the reformer) is fed to a waste heat boiler for high pressure steam generation and then to a shift converter (Figure 18).

In the liquefaction mode, the C_1 produced in the FHP is utilized in the following order of priority: (1) for steam reformer feed and fuel, (2) for ethylene plant fuel, (3) for H_2 furnace fuel, and (4) for reformer steam generation.

Shift Conversion (Figure 18)

The gaseous product of the steam reformer is directed to a two-stage shift conversion system where the carbon monoxide produced in the reformer is utilized to generate hydrogen by the shift reaction. CO recovered by the Cosorb system (Figure 15) is also fed to the converter.

In the first-stage converter, since the concentration of CO is relatively high, a fairly large quantity of heat is released as the CO reacts with the feed stream. The heat of reaction is removed by indirect heat exchange with boiler feed water. The second stage converter is

operated adiabatically with an approximate inlet temperature of 630°F and an outlet temperature of 653°F. These conditions have been proven in commercial operation, and result in good conversion of CO without the need for excessive steam.

The H₂ rich stream which exits the converter is cooled producing gas liquor, W₁, and then fed to an MEA system for CO₂ removal. Stream W₁, is directed to condensate cooler and then fed to a Biox unit (Figure 21). Carbon Dioxide Removal (Figure 18).

Carbon dioxide is removed from the hydrogen synthesis gas (from shift conversion) by means of an MEA absorption system. Use of an MEA system for the subject service results in a simpler operation and requires less equipment than if a multiple adsorbent system, such as hot potassium carbonate followed by MEA, is used. The use of a single absorbent enables the CO₂ removal to be carried out at different stages, but with a common regenerator. The second-stage absorber has a relatively high liquid rate, and since the pickup is quite small in this tower, the solution is still adequately "lean". It can, therefore, be combined with the primary lean solution flow and fed to the first-stage MEA absorber for bulk CO₂ removal.

This split system results in a substantial steam economy, since the stripping steam requirement is essentially only that required for the first-stage stripper. The MEA system is highly efficient, and the stripping steam requirements are supplied by heat recovery from the process gas.

After purification, a portion of the purified H_2 stream is fed to the refinery complex; the remaining portion is recycled to the FHP unit. The main effluents of the MEA system are CO_2 , A_3 , which is discharged to the atmosphere and some degraded MEA, W_5 , which is blown-down periodically to a Biox unit (Figure 21).

Sulfur Removal (Figure 19)

The sour gas effluent of the Selexol system (Figure 15) is fed to a Claus plant for conversion of H_2S to saleable elemental sulfur. The concentration of CO_2 in the Claus plant feed has a dramatic effect on sulfur conversion, however, for the subject case very little CO_2 is present, therefore, the CO_2 should not play a significant role with respect to sulfur conversion. The tail gas from the Claus plant, A_4 , is expected to contain COS, SO_2 , CS_2 , and other sulfur compounds and may require further treatment such as incineration, followed by flue gas desulfurization.

Refinery

The refinery complex produces 90 RON motor gasoline and LPG. The design basis and philosophy for the refinery complex can be summarized as follows:

- 47,700 BPD of coal liquids (BTX, light oil, and ethylene dimers) as feedstock to a petroleum refinery are produced.
- No hydrogen is produced for sale. The refinery is designed to utilize methane which is produced in situ, and generate only sufficient hydrogen to satisfy the needs of the complex.
- Hydrogen production by the steam reforming scheme eliminates the need for an oxygen plant.
- No methane is produced for sale. It is utilized within the system to generate hydrogen and fuel the needs of the processing units.

- o Carbon monoxide produced in the FHP is used for hydrogen production.
- o No attempt has been made to optimize process steps to attain low capital investment and/or to maximize ROI (return on investment).

A conceptual refinery configuration for processing the coal liquids is schematically shown in Figure 12. Basically, three cuts would be fed to the refinery unit.

- o 2/3 of the BTX cuts are directly blended into the gasoline pool. The balance is sent to a hydrocracking unit to make up a sufficient amount of the hydrocrackate. This split is necessary to limit aromatics to 20% by volume in the gasoline pool.
- o Light oil is sent to hydrocracking which converts low quality middle and heavy distillates into gasoline, jet fuel and higher quality middle distillates. From the laboratory analysis of the coal-derived liquids there are very few (less than .03 wt%) heteroatoms present in the feed, thus hydrotreating steps can be eliminated. There has been some experimental work done by UOP in a pilot scale Platforming unit utilizing a feedstock obtained from H-coal derived naphthas.¹ The properties of the H-coal feedstock are comparable to the properties of the 425°F to 688°F cut obtained from the light oil liquids. UOP's results have been used as a guide in estimating yields for the hydrocracking unit.
- o Ethylene is directed to a dimerization unit to produce a sufficient amount of C₄ for the gasoline pool. This process is followed by three additional processing steps, i.e., hydrogenation of the unsaturates, isomerization of the C₄, and alkylation of C₆ - C₈. It is anticipated that a gasoline having 10% C₄ and better than 26% alkylate in the blend stock will be produced.

The anticipated effluents of the refinery complex are shown in Figure 20. They consist of spent catalyst from the various processing units, S₈, S₉, S₁₀, and S₁₂, and spent sludge from the alkylation unit, S₁₁. Tar produced in the hydrocracking unit and off-gases from the hydrogenation and isomerization units can probably be used as plant fuel. Obviously, a much more detailed refinery design and analysis must be prepared before

any firm conclusions regarding effluents can be made.

Water Purification (Figure 21)

The anticipated facilities required to treat the liquid effluents of the complex are shown in Figure 21. The equipment is similar to that described under Gasification, above, and the same general statements concerning effluents can be made here.

Utilities (Figures 22 and 23)

The anticipated utility modules required to service the subject complex (excluding drinking water, etc.) are shown in Figures 22 and 23, along with anticipated effluents. The utilities consist of two make-up water filtration and demineralization systems, utility boilers and cooling towers. The units would be standard items currently in use today and their effluents should pose no unusual problems. Utility boilers would require dust collection and SO_2 scrubbing systems (Figure 23) typical of those presently used by fossil-fuel power plants. Cooling towers would also be of standard design, however, cooling tower emission rates (of air and mist) and the potential problem of cooling tower drift loss may impose constraints on site selection. The air and mist flow from the cooling towers (1.7×10^6 tons/day) is by far the largest stream released to the environment by the complex. It may be advantageous to use air cooling whenever possible to minimize the use of cooling water and thereby reduce cooling tower load and emissions.

The production rates of effluent streams A_9 , A_{10} , W_{11} , S_{14} , and S_{15} have not been specified because the steam system of the complex has not been optimized as yet.

Energy Balance and Utility Requirements

A preliminary energy balance and a utility summary have been prepared for the subject Liquefaction Complex. Material and energy balance summaries appear in Table 16; summaries of the cooling water, electricity, steam and fuel (generated internally by the complex) requirements are presented in Table 17. The overall steam rate required by the complex is based on the nominal steam usage for each of the process units indicated in Figure 1. No attempt has been made to minimize steam utilization and neither heat pumps nor other mechanical energy recovery devices have been considered. Coal for steam and power generation has not been included in the material balance.

Economic Evaluation

The installed costs for the liquefaction complex is summarized in Table 18. An economic evaluation of the liquefaction plant is presented in Table 19.

MIXED-PRODUCTS

A process block diagram of the Mixed-Products Complex is shown in Figure 24 and a schematic of the ancillary refinery in Figure 25. The complex produces pipeline gas, motor gasoline, LPG, and byproduct sulfur, ammonia, phenols and tar.

28,000 TPD of screened, dried lignite is fed to the FHP unit which operates at the same conditions, 1382°F and 2,000 psi, as the Liquefaction Case. FHP products and yields are based on data obtained from the BNL experimental unit (Table 2). Char from the FHP is utilized to produce hydrogen by BCR gasification. A small amount of coal is also fed to the

gasifier to produce the balance of the hydrogen required for the FHP. An oxygen plant is included in the processing scheme for supplying oxygen to the BCR gasifier and liquid nitrogen byproduct to the gas separation unit. Other processing units are similar to those described under Liquefaction or Gasification, above.

The effluents of the Mixed-Products Complex are similar to those described under Liquefaction or Gasification. Processing modules and effluents are shown in Figures 26 through 36, material and energy balances in Table 20, utility requirements in Table 21 and installed costs in Table 22. An economic evaluation summary appears in Table 23.

SUMMARY AND OVERVIEW

A vital national goal is to augment our supplies of gaseous and liquid fuels while simultaneously limiting the emission of effluents to the environment. In this study, the environmental impacts of coal gasification and liquefaction by means of the Flash Hydropyrolysis Process have been preliminarily assessed. Three modes of operation were considered, one for the production of pipeline gas (Gasification case), one for the production of motor fuel (Liquefaction case), and one for the production of both pipeline gas and motor fuel (Mixed-Products case). The effluents from each plant were discussed individually and in as much detail as possible considering the availability of experimental data.

It must be recognized that the subject analyses were only preliminary and that more detailed analyses would be required before commercialization would be possible. This is especially true with regard to the possible trace materials that may be contained in the various

product and effluent streams of the processing modules considered. It is intended to update the subject analyses and review the suggested control methods as more experimental data becomes available.

The design basis for the gasification complex was taken at 395 million standard cubic feet per day of pipeline gas product. Using a similar coal throughout as for gasification, the liquefaction complex would produce about 50,000 BPD of fuel oil product. It would take 32 of the commercial gasification plants to provide about 20% of the current U. S. natural gas consumption, about 40 of the liquefaction plants to provide about 25% of the current motor fuel consumption.

The estimated thermal efficiencies of the three cases under consideration are presented in Tables 12, 16, and 20. Efficiencies were calculated by dividing the heat content (heating value times quantity) of the net plant products by the heating value of the coal feed. The thermal efficiency of the gasification complex is estimated at 62%, the liquefaction plant at 50%, and the mixed products complex at 72%. Thus, it appears, from a first approximation at least, that the mixed-products complex would be the most efficient from a resource conservation viewpoint.

While the primary purpose of the subject study is to assess the environmental impacts of the various modes of operation of the FHP, we have chosen to also include some economics in the analysis. Cost estimates and economics for the three cases appear in Tables 14 and 15, 18 and 19, and in 22 and 23. The cost estimates are based on Guthrie's modularized cost method and data² and also on Burns and Roe's in-house cost files. Accuracy of the estimates is projected at not better than

± 25% and the economic analyses are based on highly debt-leveraged regulated utility financing. Using the above approach, the total investment for the gasification plant would be \$936 million, for the liquefaction plant \$834 million, and for the mixed-products plant \$892 million. Considering the accuracy of the estimates, the costs of the three plants are comparable. The investment cost attributable to environmental services is estimated at about 5% of the total investment for all three cases. On the basis of FOE (Fuel Oil Equivalent), the cost for all products is \$30.17/BBL for gasification, \$32.34/BBL for the liquefaction case and \$22.98/BBL for the mixed-products case.

WORK FORECAST FOR NEXT REPORTING PERIOD

1. Analyze available data on coal liquefaction processes which are related to the FHP process, such as those under development at Rocketdyne, Cities Service and IGT. Common problem areas will be identified as well as significant differences in process effluents and required control technologies.

2. Continue to analyze the process effluents from BNL's experimental FHP unit. The data will be used as input in developing the conceptual process schemes for the commercial plants.

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2. K. Guthrie, Process Plant Estimating Evaluation and Control, Craftsman Book of America, 1974.
3. E. M. Wewerka, et al., Environmental Contamination from Trace Elements in Coal Preparation Wastes, Los Alamos Scientific Laboratory, ERDA LA-6600-MS, EPA-600/7-76-007, August 1976.
4. Handbook of Gasifiers and Gas Treatment Systems, Dravo Corporation, FE-1772-11, February 1976.
5. A. S. Albanese and Meyer Steinberg, Environmental Control for the Flash Hydrolysis of Coal, Progress Report No. 1, Brookhaven National Laboratory, BNL 50793, October 1977.
6. A. S. Albanese and Meyer Steinberg, Environmental Control for the Flash Hydrolysis of Coal, Progress Report No. 2, Brookhaven National Laboratory, BNL 50906, May 1978.

Table 1

GASIFICATION

Coal Type - North Dakota Lignite

Coal residence time - measured --
 calculated - 9.500 sec

<u>Nominal Conditions</u>			<u>Total Accumulated Wts. (gm)</u>			
Coal feed rate	1.020	lb/hr	Coal feed	1202.0		
Hydrogen flow rate	.7600	lb/hr	Char	391.3		
Hydrogen velocity	.1641	ft/sec	Light oils	-0-		
Reactor temperature	825.0	°C	BTX	-0-		
Preheat temperature	20.00	°C	Water	-0-		
Hydrogen pressure	2500	psi				
	<u>Coal Analysis</u>	<u>Char Analysis</u>				
	(wt%)	(wt%)				
Carbon	59.10	69.00				
Hydrogen	4.200	2.530				
Oxygen	26.40	-0-				
Nitrogen	.9000	.5500				
Sulfur	.9000	1.800				
Ash	8.500	26.12				
	<u>Product Conc. in</u>	<u>% Carbon Conv.</u>	<u>% MAF Coal Conv.</u>			
	<u>Gas, lb/C. ft</u>	<u>to Product</u>	<u>to Product</u>			
Carbon Monoxide	.4896E-03	5.00	7.532			
Carbon Dioxide	-0-	-0-	-0-			
Methane	.3085E-02	55.0	47.46			
Ethylene	-0-	-0-	-0-			
Ethane	.5256E-04	1.00	.8086			
Benzene	.2276E-04	.500	.3501			
Toluene	-0-	-0-	-0-			
Xylene	-0-	-0-	-0-			
Light oil	.2274E-04	.500	.3498			
Water	.1798E-02	-0-	27.66			
	Total	62.0	84.16			
	<u>Total Material Balance</u>					
	<u>Carbon</u>	<u>Hydrogen</u>	<u>Oxygen</u>	<u>Nitrogen</u>	<u>Sulfur</u>	<u>MAF Coal</u>
Liquids	1.00	1.18	-0-	-0-	-0-	.700
Hydrocarbon gases	56.0	263.0	-0-	-0-	-0-	48.3
Carbon oxides	5.00	-0-	14.9	-0-	-0-	7.53
Water	-0-	66.9	85.2	-0-	-0-	27.7
Ammonia	-0-	3.68	-0-	80.1	-0-	.957
Hydrogen Sulfide	-0-	.467	-0-	-0-	34.9	.365
Char	38.0	19.6	-0-	19.9	65.1	26.3
Total	100	355	100	100	100	112

Hydrogen Consumption - .1072 lb hydrogen/lb coal; 2.177 mole hydrogen/mole carbon
 Effective carbon conversion - 58.37%
 Effective Energy efficiency - 82.72%

Table 2
LIQUEFACTION

Coal Type - North Dakota Lignite

Coal residence time - measured --
calculated - 8.700 sec

<u>Nominal Conditions</u>			<u>Total Accumulated Wts. (gm)</u>	
Coal feed rate	1.521	lb/hr	Coal feed	631.0
Hydrogen flow rate	.8000	lb/hr	Char	205.4
Hydrogen velocity	.2011	ft/sec	Light oils	-0-
Reactor temperature	750.0	°C	BTX	-0-
Preheat temperature	750.0	°C	Water	-0-
Hydrogen pressure	2000	psi		

	<u>Coal Analysis</u> <u>(wt%)</u>	<u>Char Analysis</u> <u>(wt%)</u>	
Carbon	59.10	69.00	
Hydrogen	4.200	2.530	
Oxygen	26.40	-0-	
Nitrogen	.9000	.5500	
Sulfur	.9000	1.800	
Ash	8.500	26.12	
	<u>Product Conc. in</u> <u>Gas, lb/C. ft</u>	<u>% Carbon Conv.</u> <u>to Product</u>	<u>% MAF Coal Conv.</u> <u>to Product</u>
Carbon Monoxide	.8325E-03	6.00	9.038
Carbon Dioxide	.2180E-03	1.00	2.367
Methane	.1987E-02	25.0	21.57
Ethylene	-0-	-0-	-0-
Ethane	.7447E-03	10.0	8.085
Benzene	.6449E-03	10.0	7.001
Toluene	-0-	-0-	-0-
Xylene	-0-	-0-	-0-
Light oil	.6447E-03	10.0	6.999
Water	.2279E-02	-0-	24.74
Total		62.0	79.80

	<u>Total Material Balance, %</u>					
	<u>Carbon</u>	<u>Hydrogen</u>	<u>Oxygen</u>	<u>Nitrogen</u>	<u>Sulfur</u>	<u>MAF Coal</u>
Liquids	20.0	23.5	-0-	-0-	-0-	14.0
Hydrocarbon gases	35.0	154.0	-0-	-0-	-0-	29.7
Carbon oxides	7.00	-0-	23.9	-0-	-0-	11.4
Water	-0-	59.9	76.2	-0-	-0-	24.7
Ammonia	-0-	3.68	-0-	80.1	-0-	.957
Hydrogen sulfide	-0-	.467	-0-	-0-	34.9	.365
Char	38.0	19.6	-0-	19.9	65.1	26.3
Total	100	261	100	100	100	107

Hydrogen Consumption -.6749E-01 lb hydrogen/lb coal; 1.370 mole hydrogen/mole carbon

Effective carbon conversion - 58.37%

Effective energy efficiency - 69.26%

Table 3

SUMMARY OF GASEOUS EFFLUENTS
GASIFICATION

<u>Stream No.</u>	<u>Ton/Day</u>	<u>Description</u>
A ₁	3,290	Coal preparation vent gas, CO ₂
A ₂	2,590	Feed system vent gas CO ₂
A ₃	31,275	Selexol off-gas CO ₂ and mist
A ₄	798	Tail gas
A ₅	25	Off-gas from Phenosolvan unit
A ₆	31,000	N ₂ from gas separation
A ₇	2,530,000	Air and mist from cooling tower
A ₈	--	Flue gas from utility boilers
A ₉	--	Desulfurized flue gas to atmosphere

Table 4

SUMMARY OF LIQUID EFFLUENTS
GASIFICATION

<u>Stream No.</u>	<u>Ton/Day</u>	<u>Description</u>
W ₁ *	24,157	Condensate from shift conversion
W ₄ *	7,535	Phenosolvan effluent
W ₅ *	1.6	MEA waste
W ₆ *	18	Dryer effluent
W ₇	0.142	COSORB effluent
W ₈ *	234	Condensate from the quench tower
W ₁₀ *	648	Blowdown or purge from condensate cooler
W ₁₁ *	--	Utility boiler blowdown
W ₁₂ *	4,800	Cooling tower blowdown
W ₁₃	--	Rain runoff

* To water purification system (Figure 8).

Table 5
SUMMARY OF SOLID EFFLUENTS
GASIFICATION

<u>Stream No.</u>	<u>Ton/Day</u>	<u>Description</u>
S ₁	100	Refuse and losses from coal preparation
S ₂	2,474	Slag and ash from BCR gasifier
S ₃	1	Adsorbent loss from gas separation unit
S ₄	--	Suspended solids from demineralizer
S ₅	4.5	Sludge from Biox unit
S ₆	18	Sludge from wastewater treatment
S ₇	21	Suspended solids from demineralizer
S ₈	--	Ash from utility boilers
S ₉	--	Sludge from the flue gas desulfurization

Table 6
SUMMARY OF GASEOUS EFFLUENTS
LIQUEFACTION

<u>Stream No.</u>	<u>Ton/Day</u>	<u>Description</u>
A ₁	3,160	Coal preparation vent gas
A ₂	183,000	Feed system vent gas and flue gas
A ₃	12,617	MEA off-gas CO ₂
A ₄	265	Tail gas from sulfur plant
A ₅	25	Off-gas from phenosolvan
A ₇	1,703,000	Air and mist from cooling tower
A ₈	600	Selexol off-gas CO ₂
A ₉	--	Flue gas from utility boilers
A ₁₀	--	Desulfurized flue gas to atmosphere

Table 7

SUMMARY OF LIQUID EFFLUENTS
LIQUEFACTION

<u>Stream No.</u>	<u>Ton/Day</u>	<u>Description</u>
W ₁ *	30,009	Condensate from shift conversion
W ₃	359	BTX heavy ends
W ₄ *	7,650	Phenosolvan effluent
W ₅ *	7.4	MEA waste
W ₆ *	18	Dryer effluent
W ₇	0.142	COSORB effluent
W ₈ *	237	Condensate from quench tower
W ₉ *	2,820	Sour water from ethylene plant
W ₁₁ *	--	Utility boiler blowdown
W ₁₂ *	3,230	Blowdown from cooling tower
W ₁₃	--	Rain runoff

* To water purification system (Figure 21).

Table 8
SUMMARY OF SOLID EFFLUENTS
LIQUEFACTION

<u>Stream No.</u>	<u>Ton/Day</u>	<u>Description</u>
S ₁	96	Refuse and losses from coal preparation
S ₂	2,380	Ash from H ₂ furnace and reformer steam generator
S ₃	1	Adsorbent loss from gas separation
S ₄	17	Suspended solids from demineralizer
S ₅	5	Sludge from Biox unit
S ₆	20.9	Sludge from wastewater treatment
S ₇	14	Suspended solids from demineralizer
S ₈	34.3	Spent catalyst from dimerization
S ₉	11.5	Spent catalyst from hydrogenation
S ₁₀	11.5	Spent catalyst from isomerization
S ₁₁	0.5	Spent sludge from alkylation
S ₁₂	15.7	Spent catalyst from hydrocracking
S ₁₄	--	Ash from utility boilers
S ₁₅	--	Flue gas desulfurization sludge

Table 9

SUMMARY OF GASEOUS EFFLUENTS
MIXED-PRODUCTS

<u>Stream No.</u>	<u>Ton/Day</u>	<u>Description</u>
A ₁	3,290	Coal preparation vent gas
A ₂	2,590	Feed system vent gas and flue gas
A ₃	24,857	Selexol off-gas, CO ₂
A ₄	753	Tail gas from sulfur plant
A ₅	25	Off-gas from phenosolvan
A ₇	2,260,420	Air and mist from cooling tower
A ₈	600	Selexol off-gas, CO ₂
A ₉	--	Flue gas from utility boilers
A ₁₀	--	Desulfurized flue gas to atmosphere

Table 10

SUMMARY OF LIQUID EFFLUENTS
MIXED-PRODUCTS

<u>Stream No.</u>	<u>Ton/Day</u>	<u>Description</u>
W ₁ *	18,709	Condensate from shift conversion
W ₃	359	BTX heavy ends
W ₄ *	7,650	Phenosolvan effluent
W ₆ *	18	Dryer effluent
W ₇	0.142	COSORB effluent
W ₈ *	237	Condensate from quench tower
W ₉ *	2,820	Sour water from ethylene plant
W ₁₁ *	--	Blowdown from utility boilers
W ₁₂ *	4,287	Blowdown from cooling tower
W ₁₃	--	Rain runoff

* To water purification system (Figure 34).

Table 11

SUMMARY OF SOLID EFFLUENTS
MIXED-PRODUCTS

<u>Stream No.</u>	<u>Ton/Day</u>	<u>Description</u>
S ₁	100	Refuse and losses from coal preparation
S ₂	2,477	Slag and ash from BCR gasifier
S ₃	1	Adsorbent loss from gas separation
S ₄	--	Suspended solids from demineralizer
S ₅	4	Sludge from Biox unit
S ₆	20.9	Sludge from wastewater treatment
S ₇	19	Suspended solids from demineralizer
S ₈	34.3	Spent catalyst from demineralization
S ₉	11.5	Spent catalyst from hydrogenation
S ₁₀	11.5	Spent catalyst from isomerization
S ₁₁	0.5	Spent sludge from alkylation
S ₁₂	15.7	Spent catalyst from hydrocracking
S ₁₄	--	Ash from utility boilers
S ₁₅	--	Sludge from flue gas desulfurization

Table 12

MATERIAL AND ENERGY BALANCES
GASIFICATION

<u>Feed</u>	<u>Ton/Day</u>	<u>MM Btu/Day</u>	
Lignite to Flash Hydrolyzer	25,000	568,400	
Lignite to BCR Gasifier	3,946	89,600	
H ₂ O for Plant (net)	4,604		
O ₂ for H ₂ Generation	11,464		
Total	45,014	658,000	
<u>Products from Plant</u>			
Benzene	80	2,882	
Light Oil	80	3,192	
Pipeline Gas*	11,020	493,300	395,780 (net production)
CO*	1,706	14,800	
CO ₂	29,256		
NH ₃	192	3,710	
Sulfur	203	1,620	
Ash	2,474		
Phenols	3	84	
Total	45,014	519,588	407,268

Thermal Efficiency = 62% (net products/coal feed)

* Used internally for fuel

Table 13

UTILITY REQUIREMENTS
GASIFICATION

<u>Unit</u>	<u>C. W. gpm</u>	<u>Power kw</u>	<u>Steam lb/hr</u>			<u>Fuel MM Btu/hr</u>
			<u>2500 psia</u>	<u>650 psia</u>	<u>50 psia</u>	
Coal Preparation & & Feeding System	97,800	46,400	58,300			
Char Separation		171				
Quench Tower	55,000	335	(918,000)	(332,000)		
Decanter		50				
O ₂ Plant	250,000	24,000	248,500			
BCR Gasifier		8,370	35,000			
Shift Conversion			1,850,000	(725,000)		
COSORB	12,900	4,825			214,500	
Selexol	29,250	54,500			366,000	
Sulfur Plant	86	148			(51,700)	
H ₂ Compression	10,600		202,000			
Gas Separation	11,400	30				
MEA Unit	25,600	46			343,000	
Hydrogen Furnace			(454,000)			4,680
Recycle Gas Stm Generator			(619,000)			
Total	493,410	138,830	402,800	(1,057,000)	871,800	4,680
Cooling System		<u>29,450</u> 168,280				

Table 14
MODULARIZED INSTALLED COST
GASIFICATION

<u>Unit</u>	<u>\$ MM</u>	<u>Percentage of Investment</u>
Lignite Preparation, Storage, Transport Pressurizing and Feeding Systems	94	10
Hydropyrolyzer	44.25	4.72
Char and Solid Separation System	10.00	1.1
Iron Oxide Tower	3.35	0.35
Quench Tower	11.65	1.24
Decanter	1.0	0.1
MEA and COSORB System	31.1	3.32
Dryer	2.76	0.30
Gas Separation Unit	1.62	0.17
H ₂ Furnace	63.6	6.80
Recycle Gas/Superheat Steam Generator	7.0	0.75
Feeding Gas Compressor	8.1	0.86
H ₂ Recycle Compression	37.80	4.04
H ₂ Compression	39.67	4.24
O ₂ Plant	89.7	9.59
BCR H ₂ Generator	11.6	1.24
Shift Converter	19.54	2.09
Selaxol	59.96	6.40
Claus Plant	6.5	*
Pre-Quench Steam Generator	25.67	2.74

* Included in Environmental

Table 14 (Continued)

MODULARIZED INSTALLED COST
GASIFICATION

<u>Unit</u>	<u>\$ MM</u>	<u>Percentage of Investment</u>
Recycle Gas/H ₂ Exchanger	33.0	3.53
H ₂ Furnace Steam Generator	96.25	10.29
Post-BCR Gasifier Steam Generator	6.23	0.66
Post Shift Steam Generator	6.23	0.66
Quench Water Coolers	1.22	0.13
Services	61.27	6.55
Environmental	<u>37.0</u>	4.64
Total	810.07	
Land	10.0	1.1
Working Capital	34.8	3.71
Mining Equipment, Facility & Land Development	<u>80.7</u>	8.63
Total	935.57	

Table 15

ECONOMIC EVALUATION
GASIFICATION

<u>Operating Expense</u>	<u>\$MM/yr</u>
Lignite @ \$20/Ton	212.47
Catalyst and Chemicals	5.00
Power @ 15 mills/kwhr	8.3
Ash Disposal	2.7
Insurance, Maintenance, G. A. 8% of Capital Cost	74.85
Operating Labor	<u>8.0</u>
Subtotal	\$311.32
Mortgage @ 10%	109.89
Depreciation @ 5% (20 yrs)	44.77
10% Return on Investment and Income Tax	<u>187.11</u>
Total	\$653.09
Pipeline Gas 395 MMSCFD	
Selling Price	\$ 4.53/MM Btu
FOE	\$ 30.17/BBL

Table 16
MATERIAL AND ENERGY BALANCES
LIQUEFACTION

<u>Feed</u>	<u>Ton/Day</u>	<u>MM Btu/Day</u>
Lignite to Flash Hydropyrolizer	28,000	636,000
H ₂ O for Plant (net)	<u>2,115</u>	<u> </u>
Total	30,115	636,000
<u>Products from Plant</u>		
C ₁ ¹	2,562	114,600
LPG	134	24,650
Motor Gasoline	5,481	286,200
Tar	55	1,650
Sulfur	186	1,485
NH ₃	184	3,555
Phenols, Cresols	3	84
Char ²	9,114	164,052
CO ₂	<u>12,396</u>	<u> </u>
Total	30,115	596,276

Thermal Efficiency - 50%

- 1) Used internally for fuel.
- 2) Used internally for fuel and for steam generation.

Table 17

UTILITY REQUIREMENTS
LIQUEFACTION

Unit	C. W. gpm	Power kw	Steam lb/hr			Fuel MM Btu/hr
			HP	MP	LP	
Coal Preparation & Feeding System	94,200	44,700	56,150			
Char Separation		165				
Quench Tower	50,200	340	(832,000)	(337,000)		
Light Oil Unit	2,175	181		3,630		61
BTX Extraction	4,556	316			7,563	
Reformer				230,000		4,075.5
Shift Conversion			840,000	(514,000)		
MEA Unit	48,973	87			655,893	
COSORB	15,638	5,849		39,789	220,824	
Selexol	664	1,278			8,580	
Sulfur Plant	78	135			(4,745)	
H ₂ Compression	7,350		140,300			
Gas Separation	10,300	31				
Hydrogen Furnace			(438,000)			2,200
Recycle Gas Stm. Generator			(529,200)			
Dimerization	7,910	296		26,344		
Olefin Hydrogenation		1,365		5,278		5.5
Isomerization		1,800		6,950		7.2
Alkylation	9,460	715		15,070		77

Table 17 (Continued)

UTILITY REQUIREMENTS
LIQUEFACTION

<u>Unit</u>	<u>C. W. gpm</u>	<u>Power kw</u>	<u>Steam lb/hr</u>			<u>Fuel MM Btu/hr</u>
			<u>HP</u>	<u>MP</u>	<u>LP</u>	
Hydrocracker	3,327	10,090		(27,597)		175.6
Gasoline Blending		124				
Ethylene Unit	37,139	9,660	(224,000)			2,130
Refrigeration for Gas Separation	40,440		272,000			
Total	332,401	75,902	(714,750)	(551,536)	888,115	8,731.8
Cooling System		<u>19,837</u> 95,739				

Table 18

MODULARIZED INSTALLED COST
LIQUEFACTION

<u>Unit</u>	<u>\$ MM</u>	<u>Percentage of Investment</u>
Lignite Preparation, Storage, Transport and Feeding System	92	10.97
Hydropyrolyzer	44.25	5.27
Char and Solid Separation System	10.00	1.19
Iron Oxide Tower	3.36	0.40
Quench Tower	11.65	1.39
BTX Extraction Unit	20.5	2.44
Dryer	2.1	0.25
Selexol and COSORB System	44.6	5.32
Gas Separation Unit	32	3.82
Ethylene Plant	95	11.32
H ₂ Furnace	30	3.57
Recycle Gas/Superheat Steam Generator	7	0.83
H ₂ Recycle Compression	4.6	0.55
H ₂ Feed Compression	4.7	0.56
Steam Reformer	37	4.41
Shift Conversion	13.3	1.59
MEA Unit	31.5	3.75
Claus Plant	5.9	*
Refrigeration Unit	11	1.31
Pre Quench Steam Generator	14.7	1.75

* Included in Environmental

Table 18 (Continued)

MODULARIZED INSTALLED COST
LIQUEFACTION

<u>Unit</u>	<u>\$ MM</u>	<u>Percentage of Investment</u>
Recycle Gas/H ₂ Heat Exchanger	2.57	1.75
H ₂ Furnace Steam Generator	11.85	1.41
Post Reformer Steam Generator	14.8	1.76
Post Shift Steam Generator	8.0	0.95
Quench Cooler	0.4	0.05
Post Quench Cooler	0.1	0.01
Feed Compressor	2.02	0.24
Reformer Waste Heat Steam Generator	30	3.58
Dimerization	6.1	0.73
Olefin Hydrogeneration	1.8	0.21
Isomerization	8.1	0.96
Alkylation	8.9	1.06
Hydrocracker	33	3.93
Gasoline Blending	2.78	0.33
Services	37.7	4.49
Environmental	<u>31.21</u>	4.42
Subtotal	714.49	
Land	10.0	1.19
Working Capital	33.6	4.00
Mining Equipment, Facility & Land Development	<u>80.7</u>	9.62
Total	838.79	

Table 19

ECONOMIC EVALUATION
LIQUEFACTION

<u>Operating Expense</u>	<u>\$MM/yr</u>
Lignite @ \$20/Ton	204.4
Catalyst and Chemicals	5.5
Power @ 15 mills/kwhr	1.1
Ash Disposal	2.6
Insurance, Maintenance, G. A. 8% of Capital Cost	57.16
Operating Labor	<u>8.0</u>
Subtotal	\$278.76
Mortgage @ 10%	93.40
Depreciation @ 5% (20 yrs)	41.94
10% Return on Investment and Income Tax	<u>167.76</u>
Total	\$581.86
Gasoline 47,700 BPD	
LPG 134 Ton/Day	
Selling Price	\$ 5.13/MM Btu
	\$.77/gal
FOE	\$ 32.34/BBL

Table 20

MATERIAL AND ENERGY BALANCES
MIXED-PRODUCTS

<u>Feed</u>	<u>Ton/Day</u>	<u>MM Btu/Day</u>
Lignite to Flash Hydropyrolyzer & BCR Gasifier	29,140	661,900
O ₂	9,729	
H ₂ O for Plant (net)	<u>864</u>	<u> </u>
Total	39,773	661,900
<u>Products from Plant</u>		
CO*	2,315	19,600
C ₁ *	5,583	250,537 (159,850 net production)
LPG	134	24,650
Motor Gasoline	5,481	286,200
Tar	55	1,650
Sulfur	201	1,605
NH ₃	184	3,555
Phenols, Cresols	3	84
Ash	2,477	
CO ₂	<u>23,300</u>	<u> </u>
Total	39,733	587,881

Thermal Efficiency - 72%

* Used internally for fuel.

Table 21

UTILITY REQUIREMENTS
MIXED-PRODUCTS

Unit	C. W. gpm	Power kw	Steam lb/hr			Fuel MM Btu/hr
			HP	MP	LP	
Coal Preparation & Feeding System	94,200	44,700	56,150			
Char Separation		165				
Quench Tower	50,200	340	(832,000)	(337,000)		
Light Oil Unit	2,175	181		3,630		
O ₂ Plant	181,567	15,510	160,597			
BCR Gasifier		5,409	22,619			
Shift Conversion			1,195,596	(468,544)		
COSORB	12,900	4,825			214,500	
Selexol	18,803	35,221			236,534	
Sulfur Plant	92	112			(39,034)	
H ₂ Compression	7,350		140,300			
Gas Separation	10,300	31				
Selexol Unit	25,600	46			343,000	
Hydrogen Furnace			(438,000)			2,200
Recycle Gas Steam Generator			(529,200)			
Dimerization	7,910	296		26,344		
Olefin Hydrogenation		1,365		5,278		5.5
Isomerization		1,800		6,950		7.2
Alkylation	9,460	715		15,070		77
Hydrocracker	3,327	10,090		(27,597)		175.6
Gasoline Blending		124				
Ethylene Unit	37,139	9,660	(224,000)			2,130
Total	441,203	130,590	(447,938)	(775,869)	755,000	4,595.3
Cooling System		31,276				
		161,866				

Table 22

MODULARIZED INSTALLED COST
MIXED-PRODUCTS

<u>Unit</u>	<u>\$ MM</u>	<u>Percentage of Investment</u>
Lignite Preparation, Storage, Transport and Feeding System	94	10.54
Hydropyrolyzer	44.25	4.96
Char and Solid Separation System	10.00	1.12
Iron Oxide Tower	3.36	0.37
Quench Tower	11.65	1.30
BTX Extraction Unit	20.5	2.30
Dryer	2.1	0.24
Selexol and COSORB System	44.6	5.0
Gas Separation Unit	32	3.58
Ethylene Plant	95	10.65
H ₂ Furnace	30	3.36
Recycle Gas/Superheat Steam Generator	7	0.78
H ₂ Recycle Compression	4.6	0.52
H ₂ Feed Compression	4.7	0.53
BCR H ₂ Generator	11.6	1.3
Shift Conversion	13.3	1.49
Selexol Unit	46	5.16
Claus Plant	6.0	*
Post-BCR Gasifier Steam Generator	6.23	0.70
Oxygen Plant	80	8.97

* Included in Environmental

Table 22 (Continued)

MODULARIZED INSTALLED COST
MIXED-PRODUCTS

<u>Unit</u>	<u>\$ MM</u>	<u>Percentage of Investment</u>
Pre Quench Steam Generator	14.7	1.65
Recycle Gas/H ₂ Exchanger	2.57	0.29
H ₂ Furnace Steam Generator	11.85	1.33
Post Shift Steam Generator	8.0	0.9
Quench Cooler	0.4	0.04
Post Quench Cooler	0.1	0.01
Feed Compressor	2.02	0.22
Dimerization	6.1	0.68
Olefin Hydrogeneration	1.8	0.20
Isomerization	8.1	0.91
Alkylation	8.9	1.00
Hydrocracker	33	3.70
Gasoline Blending	2.78	0.31
Services	61.3	6.87
Environmental	<u>38</u>	4.93
Subtotal	766.51	
Land	10.0	1.12
Working Capital	35	3.92
Mining Equipment, Facility & Land Development	<u>80.7</u>	9.04
Total	892.21	

Table 23

ECONOMIC EVALUATION
MIXED-PRODUCTS

<u>Operating Expense</u>	<u>\$MM/yr</u>
Lignite @ \$20/ton	212.7
Catalyst and Chemicals	5.5
Power @ 15 mills/kwhr	6.8
Ash Disposal	2.8
Insurance, Maintenance, G. A. 8% of Capital Cost	71.38
Operating Labor	8.0
Subtotal	\$309.16
Mortgage @ 10%	104.8
Depreciation @ 5% (20 yrs)	44.61
10% Return on Investment and Income Tax	178.42
Total	\$635.03
Gasoline 47,700 BBL	
LPG 134 Ton/Day	
Pipeline Gas: 3550 Ton/Day ((159) MIXED)	
Selling Price for all Products:	\$ 3.83/MM Btu
	\$.54/gall
FOE	\$ 22.98/HBL

Figure 1. Flash hydropyrolysis of coal to pipeline gas.

GASEOUS EFFLUENT	TONS/DAY
A ₁	3290
A ₂	2590
LIQUID EFFLUENT	
W ₁₃	-
SOLID EFFLUENT	
S ₁	100

UNITS: TONS / DAY

Figure 2. Coal preparation and hydropyrolysis.

GASEOUS EFFLUENT	TONS/DAY
A ₅	25
LIQUID EFFLUENT	
W ₄	7535

Figure 3. Oil recovery.

Figure 4. Gas purification.

Figure 5. Gas separation.

Figure 6. Hydrogen production.

Figure 7. Sulfur recovery.

Figure 8. Water purification.

Figure 9. Utilities.

Figure 10. Flue gas desulfurization.

Figure 11. Flash hydrolypyrolysis of coal to motor gasoline.

Figure 12. Refinery schematic.

Figure 13. Coal preparation, hydropyrolysis and methane reforming.

Figure 14. Oil recovery.

Figure 15. Gas purification.

NOTE: THE COOLING MEDIUM FOR THE GAS SEPARATION SYSTEM IS SUPPLIED BY A SEPARATE REFRIGERATION UNIT SINCE THE LIQUEFACTION COMPLEX DOES NOT CONTAIN AN AIR SEPARATION PLANT.

SOLID EFFLUENT	TONS/DAY
S ₃	1

Figure 16. Gas separation.

LIQUID EFFLUENT	TONS/DAY
W ₉	2820

Figure 17. Ethylene plant.

Figure 18. Shift conversion and CO₂ removal.

Figure 19. Sulfur recovery.

Figure 20. Refinery effluents.

Figure 21. Water purification.

Figure 22. Utilities.

Figure 23. Flue gas desulfurization.

Figure 24. Flash hydrolysis of coal to mixed products.

Figure 25. Refinery schematic.

Figure 26. Coal preparation and hydropyrolysis.

Figure 29. Gas separation.

Figure 30. Ethylene plant.

Figure 31. Hydrogen production.

Figure 12. Sulfur recovery.

Figure 33. Refinery effluents.

Figure 34. Water purification.

Figure 35. Utilities.

Figure 36. Flue gas desulfurization.

Distribution

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