CASE STUDIES ON DIRECT LIQUEFACTION OF LOW RANK WYOMING COAL

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ABSTRACT

Previous Studies have developed process designs, costs, and economics for the direct liquefaction of Illinois No. 6 and Wyoming Black Thunder coals at mine-mouth plants. This investigation concerns two case studies related to the liquefaction of Wyoming Black Thunder coal. The first study showed that reducing the coal liquefaction reactor design pressure from 3300 to 1000 psig could reduce the crude oil equivalent price by 2.1 \$/bbl provided equivalent performing catalysts can be developed. The second one showed that incentives may exist for locating a facility that liquefies Wyoming coal on the Gulf Coast because of lower construction costs and higher labor productivity. These incentives are dependent upon the relative values of the cost of shipping the coal to the Gulf Coast and the increased product revenues that may be obtained by distributing the liquid products among several nearby refineries.

INTRODUCTION

This study is an extension of DOE Contract Number DE-AC22-90PC89857 (which started in 1990) in which several process designs and economics were developed for direct coal liquefaction facilities processing Illinois No. 6 coal at a mine-mouth location in southern Illinois[1-5]. ASPEN process flowsheet simulation (PFS) models also were developed for the Baseline design case and the following optional cases:

- Additional coal cleaning by heavy media separation
- Additional coal cleaning by heavy media separation and spherical agglomeration
- Thermal-catalytic liquefaction reactor configuration
- Catalytic-catalytic liquefaction reaction configuration with vent gas separation
- Fluid coking as a alternate vacuum bottoms processing option
- Hydrogen production by steam reforming of natural gas
- Addition of a naphtha reforming plant

As the above designs were being developed, additional pilot plant data became available from the advanced coal liquefaction facility at Wilsonville, Alabama. Thus, the contract was extended to develop a design, economics and an ASPEN PFS model for an Improved Baseline case based on this higher space velocity data[4,5]. An additional case based on this Improved Baseline design also was developed which produced the required hydrogen by steam reforming of natural gas rather than by coal gasification.

In April, 1993, this contract was modified to develop two designs, economics and ASPEN PLUS simulation models for a direct liquefaction facility processing a low rank, sub-bituminous Black

Thunder coal at a mine-mouth plant located near Gillette Wyoming. The results for the base case design with hydrogen production by coal gasification and for one where hydrogen is produced by steam reforming of natural gas have been reported[6-8].

This paper reports on the results of two case studies based on the low rank coal liquefaction design. The first study considers the hypothetical effects on the design and economics of reducing the design pressure of the coal liquefaction reactors from 3300 psig to 2000 psig and to 1000 psig. The second one considers the effects of relocating the plant from Gillette, Wyoming to the Louisiana Gulf Coast.

LOW RANK COAL PROCESS DESIGN

Table I shows the properties of the low rank coal from the Black Thunder mine which was used in this study. HRI Inc. developed the process design for the high pressure coal liquefaction reactor section utilizing a two-stage ebulated-bed catalytic reactor system[9]. This design is based on data from Wilsonville pilot plant runs 262E and 263J[7-9]. The coal liquefaction reactors have a design pressure of 3300 psig although the first stage reactors operate at 3100 psig.

Figure 1 is a simplified block flow diagram of the main processing area for the entire base case design. Not shown in this figure are the air separation plant, sulfur recovery plant, phenol recovery plant, and other offsite plants. Run of Mine coal enters the complex through the coal screening, crushing and grinding plant (Plant 1) where partial drying is achieved in the presence of a nitrogen purge. Coal containing 6.3 wt% ash (MF) is further dried to a moisture level of 2.0 wt% in a slurry drier (Plant 1.4) before entering the coal liquefaction plant (Plant 2).

The light products from the coal liquefaction plant are sent to the gas plant (Plant 3) to separate the fuel gas, propane and mixed butanes. The C5-350 F stream goes to the naphtha hydrotreater (Plant 4). The 350-850 F fraction from Plant 2 goes to the gas oil hydrotreater (Plant 5). Hydrogen is produced by coal gasification in Plant 9. The bottoms from the coal liquefaction plant goes to a Kerr-McGee Rose-SR unit (Plant 8) which produces an extract stream that is recycled back to Plant 2 and an ash concentrate stream that goes to the gasification section of Plant 9. The oxygen required by the Texaco gasification section of Plant 9 is produced in the air separation plant (Plant 10).

The hydrogen purge from Plant 2 is recovered by the hydrogen purification plant (Plant 6) which is a combination of membrane and PSA units. The purified hydrogen goes to Plants 2 and 5. Sulfur is produced in a sulfur recovery plant (Plant 11). The sour water collected from the various plants is sent through an ammonia recovery plant (Plant 38) before going either to the coal gasification plant or to the phenol recovery plant (Plant 39) followed by a waste water treatment plant (Plant 34).

Table II shows the major input and output flows from the plant on a stream day basis. From 24,952 tons/day of dry ROM coal from the Black Thunder mine, this plant produces 67,312 bbls/day of C4+ hydrocarbons. The naphtha is only 10% of the C5+ liquids production (naphtha, distillates and gas oil). In addition, this plant produces significant amount of byproducts; 167 tons/day of ammonia, 127 tons/day of liquid sulfur, and 45 tons/day of mixed phenols.

EFFECT OF LIQUEFACTION REACTOR DESIGN PRESSURE

Direct coal liquefaction is a capital intensive process as has been demonstrated in numerous previous studies. For the above described low rank coal liquefaction design, the costs associated with the coal liquefaction plant are about 28% of the installed field costs (total ISBL and OSBL costs) of the liquefaction facility. A good portion of these costs can be attributed to the cost of the large thick-walled high pressure reactors, other high pressure vessels, and the compressors required to supply the high pressure hydrogen-rich gas. Therefore, this study was made to determine the incentives for conducting catalyst and process research to develop coal liquefaction catalysts which will operate at lower pressures.

Since these catalysts do not yet exist and insufficient thermodynamic, kinetic and hydrodynamic data are available for a complete and detailed study, this is a somewhat hypothetical study. Therefore, the following assumptions were made.

- The kinetics of the coal liquefaction reaction remains unchanged as a function of pressure.
- The coal liquefaction reaction product distribution and product composition are independent of the coal liquefaction pressure.
- Most stream compositions are independent of the coal liquefaction pressure.
- The cost of the new coal liquefaction catalysts which will operate at low pressure are the same as those which are used in the base low rank coal design case.
- There is no effect of coal liquefaction pressure on either the design or cost of the OSBL plants.
- There is no net effect of coal liquefaction pressure on the total utilities consumed by the entire facility. This is a conservative assumption that will underestimate the cost savings of lower liquefaction reactor pressures because less power is required to compress the make-up hydrogen and pump the liquids into the liquefaction reactors.

Besides the major effect of reaction pressure on the cost of the coal liquefaction plant, the coal liquefaction pressure may have a small effect on the cost of some of the other processing plants.

<u>Plants 1 and 1.4 - The Coal Crushing, Grinding and Drying Plants</u> -- The coal liquefaction pressure has no effect on these plants.

<u>Plant 3 - The Gas Plant</u> -- This plant uses lean oil absorption at 200 psig to recover light hydrocarbons from several gas streams. Since this pressure is well below the lowest coal liquefaction reactor design pressure of 1000 psig which is being studied, the coal liquefaction pressure has no effect on this plant.

<u>Plant 4 - The Naphtha Hydrotreater</u> -- The naphtha hydrotreater reactor operates at about 1000 psig. The naphtha feed comes from either intermediate storage or the low pressure fractionation section of Plant 2 and has to be pumped up to reactor pressure. The make-up hydrogen comes from either the Plant 3 (gas plant) or Plant 9 (coal gasification plant), and consequently, is independent of the coal liquefaction reactor pressure. In either case, these make-up hydrogen streams have to be compressed to the naphtha hydrotreater reactor pressure. Therefore, there is no effect of liquefaction reactor pressure on this plant.

<u>Plant 5 - The Gas Oil Hydrotreater</u> -- The gas-oil hydrotreater reactor operates at about 2600 psig. The distillate and gas oil feeds come from either intermediate storage or the low pressure fractionation section of Plant 2 and have to be pumped up to reactor pressure. The make-up hydrogen comes from either Plant 2 (coal liquefaction plant), Plant 3 (gas plant), or Plant 9 (hydrogen production by coal

gasification plant). The make-up hydrogen streams from Plants 3 and 9 always have to be compressed to the gas oil hydrotreater reactor pressure. Only when the coal liquefaction reactor pressure is below the gas oil hydrotreater reactor pressure does the hydrogen-rich stream from the coal liquefaction plant require compression. Thus, for the coal liquefaction reactor pressure cases at 1000 and 2000 psig, an additional make-up hydrogen-rich gas compressor is needed. Consequently, the ISBL cost of Plant 5 increases as the liquefaction reaction pressure decreases.

<u>Plant 6 - The Hydrogen Purification Plant</u> -- The hydrogen purification plant has two major sections. The first section is a high pressure section which recovers hydrogen by membrane permeation from the high pressure purge gas from the coal liquefaction and gas oil hydrotreating plants. In this section, the key parameter for the hydrogen separation is the pressure difference across the membranes. The second section is a low pressure section which recovers hydrogen by pressure swing adsorption (PSA) from the lower pressure purge gas from the coal liquefaction, naphtha hydrotreating, gas oil hydrotreating, and gas plants. The cost of this section of the plant is independent of the liquefaction reaction pressure. However, the overall cost of Plant 6 does vary with liquefaction pressure. As the liquefaction reaction pressure decreases, so does the ISBL cost of this plant.

<u>Plant 8 - The ROSE-SR Critical Solvent Deashing Plant</u> - The feed to this plant comes from the bottom of the vacuum tower of Plant 2. Consequently, the coal liquefaction pressure has no effect on this plant.

<u>Plant 9 - The Hydrogen Production by Coal Gasification Plant</u> -- The cost of the hydrogen production by coal gasification plant is independent of the liquefaction reaction pressure except for the final hydrogen compressor. This machine compresses the product hydrogen going to Plants 2 and 6. Thus, the cost of this plant decreases as the liquefaction reaction pressure decreases only because a smaller compressor is required to supply that portion going to Plant 2.

Table III compares the fourth quarter 1993 capital costs of an Nth plant for the entire low rank coal liquefaction reactor complex at three liquefaction reactor design pressures; 3300, 2000 and 1000 psig. As the liquefaction reactor design pressure drops from 3300 psig to 1000 psig, the field cost of Plant 2 drops by 245 MM$ to 547 MM$; a decrease of about 31%. This corresponds to a drop in the total installed cost of the entire complex of 324 MM$ to 3368 MM$. However, this is only a drop of 8.8% in the cost of the entire coal liquefaction complex because so much of the cost of the facility is in either the other ISBL plants or the OSBL plants.

Figure 2 graphically shows the amount of COE reduction for the three cases. At a coal liquefaction reactor design pressure of 2000 psig, the crude oil equivalent (COE) price is about 1.3 $/bbl less than that at the 3300 psig liquefaction reactor design pressure. At a coal liquefaction reactor design pressure of 1000 psig, the COE price is about 2.1 $/bbl below that at the 3300 psig liquefaction reactor design pressure of the base case. This corresponds to about a 6.5% reduction of the COE when going from a 3300 psig to a 1000 psig liquefaction reactor design pressure.

Thus, there is a significant economic incentive for continuing catalyst and process research to lower the reaction pressure for direct coal liquefaction. Because of the large costs associated with all the other plants in the complex, the cost reduction is not as great as would be expected when only the coal liquefaction plant is considered. However, the savings could be more significant when considering the integration of a coal liquefaction facility into an existing petroleum refinery because the other processing plants and offsite facilities could already be available.

EFFECT OF PLANT LOCATION

A study was undertaken to determine the effect of relocating the base case coal liquefaction plant from the Wyoming mine-mouth location to a Gulf Coast location because of lower labor and operating costs. Construction costs are lower on the Gulf Coast as a consequence of the lower labor costs and higher productivity. Furthermore, there are more petroleum refineries in the area which allows the distribution of the products among several of them so that each of them can take maximum advantage of the coal liquids. Also, the option of several potential customers creates a more competitive market which, over the long term, could result in higher sales prices.

Table IV compares some cost differences between the Wyoming mine-mouth location and the Gulf Coast location. The average 1994 operator wage rate on the Gulf Coast is 12.00 $/hr which is significantly less than the 18.50 $/hr in Wyoming. Also, the overhead factor for worker benefits on the Gulf Coast is only 1.25 compared to a 1.40 factor in Wyoming. Construction and supervisory labor costs also are lower by a similar ratio.

Utility costs also are lower on the Gulf Coast. Fuel gas is about 10% less expensive. Raw water is significantly less expensive. It is only 20 cents per thousand gallons compared to $2.50 per thousand gallons at the Wyoming mine-mouth location.

The above lower labor and utility costs and a more skilled labor force on the Gulf Coast results in a lower installed plant cost at the Gulf Coast than in Wyoming. The design of the Gulf Coast plant essentially is the same as that of the base Wyoming plant with the 3300 psig coal liquefaction reactor design pressure. However, in order to save on coal transportation costs, Plant 1 still is located in Wyoming to crush and dry the ROM Black Thunder coal from 27.0 wt% moisture to 12.9 wt% moisture. It was assumed that drying the coal to 12.9 wt% moisture prior to shipping has no effect on its reactivity. Table V compares the costs of the Gulf Coast and Wyoming coal liquefaction plants in fourth quarter 1993 dollars. The installed cost of the Gulf Coast plant is 369 MM$ or 10% less than that of the Wyoming location.

The Bechtel transportation department estimated the cost of transporting the Black Thunder coal from the mine to the plant located on the Louisiana Gulf Coast along the Mississippi River to be 15 $/ton in large rail cars carrying over 100 tons per car. However, in actual practice this will be a negotiated price, and can vary. Therefore, in the subsequent economic analysis, a range of coal transportation costs will be studied.

In the Gulf Coast case, the ash will be returned to the mine for disposal in the same manner as is done in the base Wyoming case. Since the ash from the gasifier is in a vitrified state, it does not have to be shipped in closed containers and can be returned to the mine for disposal in the same railroad cars that are used to transport the coal to the Gulf Coast. Hence, the ash transportation cost will be the same as the coal transportation cost. In the Gulf Coast case, the ash disposal cost was assumed to be half that of the Wyoming case because the ash already is in the transportation mode from being shipped to the mine and will be handled by the same equipment used to ship the coal.

PIMS LP models were used to determine values for the coal liquefaction products. For the original study, a LP model of a typical mid-west refinery located in PADD II processing 150,000 bbls/day of crude was developed[5]. This model was developed before all the ramifications of the 1990 Clean Air Act Amendments were known and understood. Studies using this model showed that Wilsonville coal liquids were, on the average, 1.07 times more valuable to the refinery than the crude oil mix when the refinery was constrained to make the same product slate. Thus, by definition, the coal liquefaction products were said to have a syncrude premium factor (SCP) of 1.07. This 1.07 SCP was used for the economic evaluation of the base low rank coal case at the Wyoming location[6-8].

Petroleum refineries located along the Gulf Coast are in PADD III and are configured differently than those in the mid-west. These refineries generally contain more heavy ends processing, such as gas oil hydrocracking and delayed coking units. Another LP model was constructed to represent a typical PADD III refinery processing 150,000 bbls/day of an average PADD III crude mix [8]. Unlike the LP model of the PADD II refinery, this model contains processing units that are capable of producing reformulated gasoline and low sulfur diesel fuel. Six cases were studied, a base case that processed no coal liquids and five cases that processed various amounts of coal liquids. When the refinery was forced to process all the coal liquids produced by the liquefaction plant, the same SCP value of 1.07 was obtained. However, as the amount of available coal liquids was reduced, the SCP value increased. When the refinery processed 50% of the total coal liquids production, the SCP increased to 1.15; at 25% of the coal liquids production, it increased to 1.20; and when the refinery only processed 1,000 bbls/day of the coal liquids, the SCP increased to 1.25.

Because there is a concentration of petroleum refineries along the Gulf coast, the coal liquefaction facility has a larger possible customer base with an increased likelihood of distributing the products among several refineries. Thus, the liquefaction plant could increase its product revenue by distributing the coal liquids among several customers who would be willing to pay more for a portion of the production than a single buyer would pay for the entire amount. Economic evaluations were made at both the low and high SCP values of 1,07 and 1.25 to bracket the effect of the SCP value.

Although the LP models used to determine the SCP values are rigorous models, the calculated SCP values are much less rigorous because of several uncertainties. For example, the behavior of the coal liquids in the PIMS LP model was estimated based on limited published properties and processing data of similar coal liquids. Additional data are needed to better determine the behavior of these materials when processed in mixtures with conventional petroleum derived materials. Currently, a "Refining and End Use Study of Coal Liquids" is in progress to obtain such processing data for both direct and indirect coal liquids. Coal liquids from the POC facility are being upgraded mixed with conventional petroleum intermediates, and the results will be used to improve the LP models. Secondly, the LP models used in this study represented typical PADD II and PADD III petroleum refineries. Markedly different results could be obtained when models of specific refineries are used because each one has different unit capacities and processing constraints.

Finally, in the following economic calculations for both cases, no transportation costs for the coal liquids were assumed. Inclusion of liquids transportation coast will lower the product revenues, and correspondingly, the calculated COEs. However, if the liquids transportation costs are different, then this omission will favor that case with the lower costs. This omission should favor the Gulf Coast location because of the proximity of more refineries and the availability of water transport. However, the magnitude of this omission cannot be calculated without knowing specific locations, refineries and means of transportation.

Figure 3 shows the effect of the coal transportation cost on the COE difference between the Gulf Coast and Wyoming locations for the lowest and highest syncrude premium values. At the lowest syncrude premium value of 1.07, the Gulf Coast location is more favorable when the coal transportation cost is less than about 9.0 $/ton. Above a coal transportation cost of 9.0 $/ton, it is better to locate the plant in Wyoming. However, when the syncrude premium has a value of 1.25, the Gulf Coast location is more favorable when the coal transportation cost is less than 20 $/ton. At the estimated 15 $/ton coal transportation cost and a 1.25 SCP value, the COE is lower by 2.4 $/bbl at the

CONCLUSIONS AND RECOMMENDATIONS

An engineering study to determine the effect lowering the design pressure of the coal liquefaction reactors from 3300 to 1000 psig showed that the crude oil equivalent price could be reduced by about 2.1 $/bbl in a grass-roots plant. However, larger savings could be realized when integrating a coal liquefaction facility into an existing petroleum refinery. Thus, there is a significant economic incentive for continuing catalyst and process research to lower the reaction pressure for direct coal liquefaction.

Another study showed that incentives may exist for constructing a facility to liquefy Black Thunder coal on the Gulf Coast rather than at a mine-mouth location in Wyoming. The amount of which depends on the relative cost of shipping the coal to the Gulf Coast and the increased product revenues that may be obtained by distributing the liquid products among several nearby refineries. LP studies suggest that the coal liquids are more valuable when distributed among several refineries. However, additional data are needed to better determine the behavior of coal liquids when they are processed in mixtures with conventional petroleum derived materials

ACKNOWLEDGMENT

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Figure 1
SIMPLIFIED BLOCK FLOW DIAGRAM

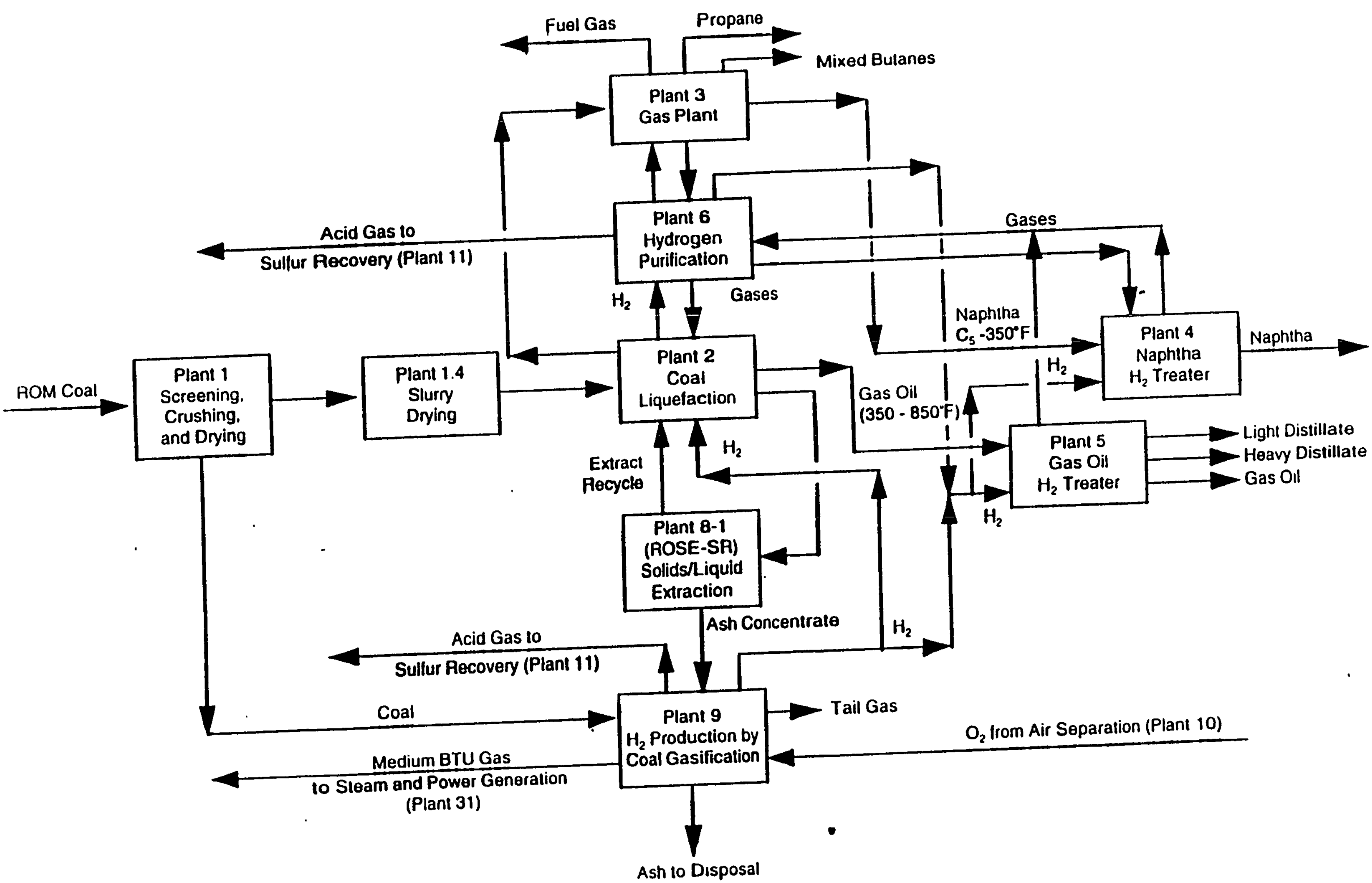

Figure 2
COE Reduction vs. Liquefaction Reactor Pressure

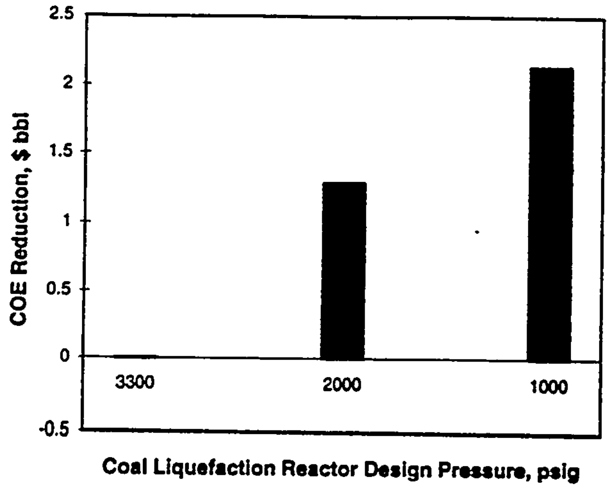

Figure 3
Effect of Coal Transportation Cost on the COE
Difference Between the Gulf Coast and Wyoming Locations

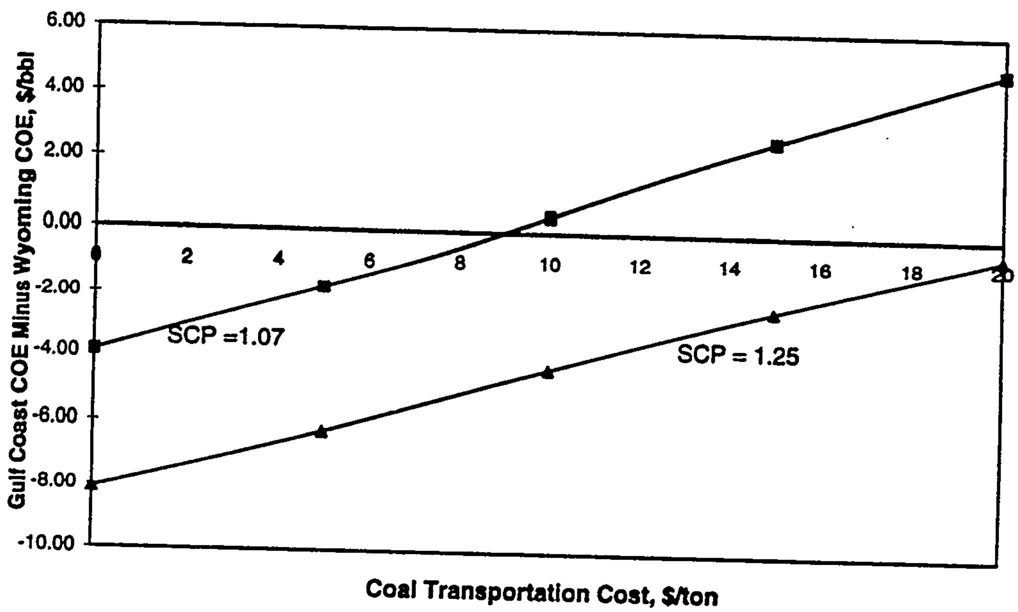

TABLE I

Properties of the Low Rank Coal from the Black Thunder Mine

Proximate Analysis (wt%)	
Volatile Matter	32.6
Fixed Carbon	35.8
Ash	4.6
Moisture	27.0
Total	100.0
Ultimate Analysis (wt% Dry Basis)	
Carbon	68.5
Hydrogen	4.8
Sulfur	0.5
Nitrogen	1.0
Ash	6.3
Oxygen (by difference)	18.9
Total	100.0

File: DOE895/Tab1

TABLE II

Major Plant Inputs, Outputs and Economic Parameters for the Base Low Rank Coal Design Case

ROM Coal Feed Rate, TSD (MF)	24952
Slag Production Rate, TSD (MF)	1659
Natural Gas Rate, MMMBTUs/SD	109.6
Raw Water Make-up, MMGSD	12.641
Naphtha Production, BSD	13063
Light Distillate Production, BSD	6610
Heavy Distillate Production, BSD	24167
Gas Oil Production, BSD	21221
Liquid Propane Production, BSD	4268
Mixed Butanes Production, BSD	2251
Total C3+ Hydrocarbons, BSD	71580
Total C4+ Hydrocarbons, BSD	67312
Ammonia Production, TSD	167
Phenol Production, TSD	45
Sulfur Production, TSD	127
No. of Operators and Boardmen	430
Total Installed Capital, $MM (4th Quarter 1993)	3692.1
Catalyst and Chemicals, $MM	80.0

File. DOE895/Tab2

TABLE III

Capital Costs for the Low Rank Coal Liquefaction Complex as a Function of Liquefaction Reactor Design Pressure

Liquefaction Reactor Design Pressure, psig	3300	2000	1000
Plant Field Costs			
Plant 2 - Coal Liquefaction Plant	792	664	547
Other ISBL Plants	1206	1187	1175
Total ISBL Plants	1998	1851	1722
OSBL Plant Costs	841	841	841
Total ISBL and OSBL Plant Costs	2839	2692	2563
Home Office, Fees and Contigency	853	805	805
Total Installed Plant Cost	3692	3498	3368

All costs are fourth quarter 1993 costs for the Nth plant.

File: DOE895/Tab3

TABLE IV

Comparison Between the Wyoming and the Gulf Coast Locations

	Wyoming	Gulf Coast
Labor		
Operator Pay Rate, $/hr	18.50	12.00
Labor Overhead Factor	1.40	1.25
Utility Costs		
Fuel Gas, $/MMBTUs	2.00	1.82
Raw Water, $/Mgals	2.50	0.20
Transportation Costs		
Coal, $/ton	0.00	Variable
Ash, $/ton	0.00	Variable
Ash Disposal Cost, $/ton	5.00	2.50
Plant Located in PADD	II	III
Syncrude Premium	1.07	1.07 to 1.25

All costs are first quarter 1994 costs.

File DOE895/Tab4&5

TABLE V

Comparison of Plant Costs Between the Wyoming and the Gulf Coast Locations

	Wyoming	Gulf Coast
ISBL Plant Field Costs, MM$	1998	1769
OSBL Plant Field Costs, MM$	841	776
Total ISBL and OSBL, MM$	2839	2545
Home Office, Fees and Contingency, MM$	853	778
Total Installed Plant Cost, MM$	3692	3323

All costs are fourth quarter 1993 costs for the Nth plant.

File:DOE895/Tab4&5