4.0 INDIRECT LIQUEFACTION USING FIXED BED GASIFICATION OF ILLINOIS #6 COAL

4.1 Introduction

Two fixed-bed coal gasification systems are considered in this report. This follows the format of MTR-80W326[1] where the dry-ash Lurgi and the BGC Slagging Lurgi were used to process a Wyoming sub-bituminous coal. This study considers these two gasification systems, using a weakly-caking Interior province coal, Illinois #6, as a feedstock.

The coal feedstock analysis is shown in Table 4-1. Gasification of low rank sub-bituminous coals or lignites by the Lurgi dry-ash fixed bed process has been commercially proven for many years. However, high and medium volatile bituminous coals, because of their tendency to swell and agglomerate and their lower reactivities compared to low rank coals have been considered less suitable candidates for this type of fixed bed processing. Eastern and Interior province U.S. coals represent an important resource in this country and it is considered worthwhile to assess their suitability as feedstocks for indirect liquefaction processes. This is especially appropriate now because of the emerging second generation gasifier technology which represents a significant potential improvement over the current commercially available systems.

TABLE 4-1

ANALYSIS OF ILLINOIS #6 FEEDSTOCK

	Percent
<u>Proximate Analysis</u> (Results on "as sampled" basis)	
Moisture	10.23
Volatile Matter	34.70
Fixed Carbon	45.97
Ash	9.10
	100.00
Swelling Number	3
Caking Index	15
<u>Ultimate Analysis</u> (Results on dry basis)	
Carbon	71.47
Hydrogen	4.83
Nitrogen	1.35
Total Sulfur	3.13
Chloride as Cl	0.06
Ash	10.14
Oxygen by Diff.	9.02
	100.00
Pyritic Sulphur	1.50
Calorific Value	Btu/lb
Gross	12,770

4.2 Data Sources Used for Process Evaluation

4.2.1 Lurgi Dry-Ash

For the dry-ash Lurgi system, data were obtained from the gasification trials of American coals that were conducted in a Lurgi gasifier located at Westfield, Scotland.(2) These trials took place between 1972 and 1974 and were jointly organized by the AGA and OCR. Woodall-Duckham was awarded the contract for design, procurement and construction work on the project. In addition, they were responsible for data compilation and analyses during the tests. The British Gas Corporation and Lurgi were involved in modification of the Lurgi gasifier to render it suitable for use in processing American caking coals.

The Lurgi dry-ash gasifier used at Westfield was a mild steel vessel operating at 25 to 30 atmospheres pressure. Approximate overall height was 19 feet and the outer diameter was 10 feet. The internal diameter was 8'9". This gasifier is much smaller than the Mark IV Lurgis used at SASOL, which have diameters of approximately 13'6". From a coal throughput consideration the Westfield gasifier was about one-sixth the size of the Lurgi Mark IV (i.e., about 6½ tons per hour compared to approximately 38 tons per hour throughput). To prevent the formation of large gas tight agglomerates, a cooled stirring mechanism was situated inside the gasifier that could cut through the softening coal mass and break up the agglomerates.

This stirrer was designed, manufactured and installed by Lurgi and apparently performed to design specifications with no problems. It must be cautioned that the modification to the existing Lurgi at Westfield does not necessarily conform to the optimum design that Lurgi could produce for the gasification of caking American coals. Thus, the operating parameters used at Westfield, and the data obtained from the tests with Illinois #6 coal, could possibly be further optimized with proper design considerations.

Results of the Illinois #6 run used in this report were for coarse graded coal, size 1¼" to ¼". In addition, a run was performed with Illinois #6 coal to determine if run-of-mine coal could be used directly in the Lurgi during gasification. The so-called simulated run-of-mine test used coal that contained 32 percent fines that were smaller than ¼". Since modern mechanical mining techniques produce large quantities of fines, it is important to be able to utilize the proportion of this fine material in the gasifier that is in excess of steam coal requirements. The simulated run-of-mine coal test was completed successfully.

Coal throughput per hour for Illinois #6 was considerably less than for a subbituminous coal like Montana Rosebud, because of the lower reactivity of the bituminous coal. Relative throughputs were calculated from the data in Reference 2. The ratio of Rosebud throughput to Illinois #6 throughput was calculated to be 1.335.

This lower throughput means utilizing more gasifiers for the same plant coal input and this will be discussed later.

The trial run using Illinois #6 coarse graded coal gasified 1,960 short tons of feedstock in total. A run-up period of two days allowed for stable gasifier operation to be achieved. Steam and oxygen rates were adjusted enabling a throughput of 6½ short tons per hour of coal to be obtained. A tar injection rate of approximately 50 gallons per hour was initiated to lower the dust content of the recycled tar. Conditions during the 48-hour test period remained at steady state. The only alteration necessary was to change the grate speed to maintain a constant depth of ash at the bottom of the gasifier. Ash discharge was satisfactory and the ash contained an average carbon content of 3.2 percent. Mass balances showed a discrepancy of 1.6 percent. After completion of the trial the gasifier was inspected. The internal chamber of the gasifier was normal except for accumulation of tarry material around the stirrer and the internal walls. The wash cooler was found to have a 3 inch layer of firm tar that had lined the lowest section. This tarry material was also found in the drain pipe work and in the waste heat boiler sump.

Another study of the gasification of Illinois #6 coal, sponsored by EPRI, was performed by Fluor Engineers & Constructors.[6] In this study, a general processing scheme using an unnamed moving bed dry-ash reactor was analyzed. The source of the Fluor gasification data is not specified but considerable differences exist between their process conditions and those of the actual test run performed at Westfield with Illinois #6 coal. Table 4-2 illustrates these differences. This large discrepancy in the steam demand between the Fluor and Westfield data impacts directly on the net efficiency in these two cases. Since the Westfield data represent the practical results of a real test run, the Westfield data are used in this report.

4.2.2 BGC Slagging Lurgi

Data relating to the gasification of Illinois #6 coal in the BGC Slagging Lurgi gasifier were obtained from the process and project engineering design report of the Continental Oil Company.(7) This report details the design of a commercial plant for the production of pipeline gas.

In the CONOCO design, the coal and the flux-handling section prepares sized coal for gasification. The size range of the run-of-mine coal is 37 percent smaller than ¼", 48 percent between ¼ and 2", 15 percent between 2 and 5". The breaking and screening unit produces coal for feed to the Slagging Lurgi gasifiers of the following size consist: 3 percent smaller than ¼", 92 percent between ¼ and 1¼", 5 percent between 1¼ and 2". The already sized flux is blended with this sized coal to produce the final gasification feedstock.

TABLE 4-2

COMPARISON OF PARAMETERS USED IN GASIFICATION OF ILLINOIS #6 COAL USING THE DRY-ASH LURGI

	Fluor[16]	Westfield[2]
Steam/Coal	1.31	2.51
Oxygen/Coal	0.34	0.48
# Moles CO + H_2/lb Coal	0.042	0.045
# Moles CH_4/lb Coal	0.010	0.008
Net Efficiency	76.6	57.8

Oxygen input to the gasifier is provided by air separation units which provide 98 percent purity oxygen at 275°F and 500 psig for the gasifier. Steam is generated by using coal fines in a conventional pulverized fuel boiler. The 1,500 pounds steam at 900°F that is generated in these boilers is transferred to the 1,500 pounds steam header for distribution. This 1,500 pounds steam is used by the condensing turbines in the air-separation units for oxygen production for gasification, and by the extraction turbine turbogenerators for power production.

Since Illinois #6 coal was never actually run in the Westfield slagging gasifier unit, the composition of the raw gas from the BGC Slagger using Illinois #6 as feedstock, as tabulated in the CONOCO report,[7] was calculated by BGC and Lurgi from elemental balances based on Scottish Francis coal. The validity of the calculated data is substantiated to a degree by actual trial runs in the BGC Slagger using Ohio #9 and Pittsburgh #8 bituminous coals.

Coal throughput in the BGC Slagger apparently is not very sensitive to the coal feedstock as illustrated by the coal gasification rate data shown in Table 4-3, which were obtained from the Westfield slagging gasifier trials.[8]

In the present analysis the coal gasification rate was assumed to be independent of the coal feedstock and the gasifier capacity required was then only proportional to the total weight of DAF coal

TABLE 4-3

PERFORMANCE DATA FROM THE WESTFIELD SLAGGING GASIFIER ON A VARIETY OF COALS (2)

	Francis (Scotland)	Killoch (Scotland)	Pittsburgh #8 (USA)	Ohio #9 (USA)
Swelling Number	1½	3½	7½	6½
Caking Index (Gray-King)	B	E	G8	G6
Coal Gasification Rate (Tonnes/ m^2h)	4.16	4.13	4.25	3.27

processed. In the Wyoming coal case in Reference 1, 1,385 pounds per hour of DAF coal was processed. In the BGC Illinois #6 case considered in this study, 1,569 pounds per hour of DAF coal was gasified. Thus, the gasifier capacity would increase by a factor of 1.133 over the capacity in the Wyoming coal case. The number of gasifiers in the Wyoming coal case using the BGC was 14, including two spares.

The BGC Lurgi gasifier operates at high temperatures with the significant advantages of both higher coal throughput and lower steam requirement than the conventional dry-ash Lurgi. This high temperature operation causes the coal ash to melt and form a fluid slag that is drained from the bottom of the combustion zone through a small tap hole. Slag viscosity is very critical for continuous trouble free slag tapping operation and a viscosity of less than 50 poises seems necessary for a slagging fixed bed gasifier. BGC claims that the design of their slag tap system allows complete control over the rate of slag discharge by allowing sufficient time for fluxing reactions to occur to produce a homogeneous slag displaying uniform and consistent flow characteristics. In the CONOCO design, lime is added as a flux to provide the correct rheological behavior of the molten slag. The lime presumably lowers the temperature of critical viscosity of the slag,[9] thus maintaining the slag as a newtonian fluid to much lower temperatures. For high iron oxide containing ashes under the highly reducing conditions of gasification, about a quarter of the iron oxide can be converted to metallic iron. Iron oxide acts as a flux for coal ash slags, thus any reduction to iron represents a net loss of flux and hence an increase in slag viscosity.[10]

4.3 General Plant Assumptions

To allow for comparison with the previous report[1] on processing Wyoming coal in indirect liquefaction, the plant coal capacity in this report, using Illinois #6 coal, has been kept constant; 27,800 TPSD of as-received coal is the total plant coal input. Although this is the same absolute weight of Illinois #6 coal, the higher rank coal contains considerably more carbon than the corresponding weight of Wyoming coal. Thus, the synthesis gas production and hence total liquid products are greater in these Illinois #6 coal feedstock cases. Because of the larger overall plant, there is inherent in the analysis an advantage due to scale when using Illinois #6 coal. This point has been addressed by the sensitivity analysis of constant plant investment discussed in Section 2.0 of this report.

The overall engineering designs of the plants are identical to the Wyoming coal cases and the same system of analysis and unit scaling is applied. Since a comparative ranking of processes is still the main objective of these reports, rather than an exact prediction of product cost, no separate DCF analysis was performed in this report. Cost data from the MRDC study[3] were again used as a basis and therefore construction costs are still given in 1977 dollars. Product costs are updated to 1980 dollars by using inflators provided in the Neilson Refinery Cost Indices.

Plant steam coal requirements for both fixed bed-gasification systems have been calculated in Appendix A and it is assumed that the balance of the plant coal can be sent to gasification.

For the Lurgi dry-ash system, data from the Westfield trial[2] have been scaled to conform to the total plant input. Production of refined Fischer-Tropsch products for both Synthol and Kolbel synthesis have been calculated from the molar amount of clean synthesis gas generated in the plants using the same analysis procedure as in Reference 1. Likewise, for the BGC Slagging Lurgi system, the data from the CONOCO report[7] have been scaled to conform to total plant input. Again, refined product outputs for Synthol and Kolbel have been calculated using the procedure of Reference 1.

Construction costs have been calculated by scaling each unit in the plant on flow capacity using .7 exponent factor. Where this procedure is not used details of the alternative procedure have been given. Capital costs are still given as 1.59 times construction cost as in the MRDC report.[3] Product costs have been calculated both on thermal and market bases as detailed in the MRDC report.

4.4 System Description and Analyses

4.4.1 Dry-Ash Lurgi/Synthol

The indirect liquefaction plant using Illinois #6 coal as feedstock will process 27,800 tons per stream day of coal in 29 Lurgi Mark IV dry-ash gasifiers operated with water-cooled stirrers. There will be three spare gasifiers on standby. For the Wyoming coal case, because of the higher reactivity of coal, only 25 gasifiers were required with three spares.

Of the 27,800 tons per stream day of coal entering the plant, 19,974 tons are sized and sent to gasification while 7,826 tons of fine coal are sent to the steam plant. The computation of plant coal split is explained in detail in Appendix A. The steam plant provides sufficient steam for oxygen generation, coal gasification and downstream plant process unit requirements. Oxygen requirement for gasification is 755,661 lb/hr. An additional 885,117 lb/hr of saturated steam is also used. On a DAF coal basis, the steam-to-coal weight ratio is 3.11, the oxygen-to-coal weight ratio is 0.563, and the steam-to-oxygen mole ratio is 9.84.

Table 4-4 shows the material and energy fluxes at each stage of the production of clean synthesis gas for the Lurgi Dry Ash/Synthol system. The raw gasifier output contains tars, oils, phenols and naphtha as well as the synthesis gas, methane, C_2's and impurities. The naphtha is scrubbed from the gas in the Rectisol unit and

TABLE 4-4

MATERIAL AND ENERGY FLUX IN SYNTHESIS GAS PREPARATION UNITS
FOR LURGI DRY ASH/SYNTHOL SYSTEM
(ILLINOIS #6 COAL)

COMPONENT	INPUTS		RAW GASIFIER OUTPUT			SHIFTED RAW GAS			CLEAN GAS TO SYNTHESIS		
	Mlb/hr	MMBtu/hr	MlbMoles/hr	Mlbs/hr	MMBtu/hr	MlbMoles/hr	Mlbs/hr	MMBtu/hr	MlbMoles/hr	Mlbs/hr	MMBtu/hr
CO			23.3	653	2,837	21.0	588	2,556	20.9	586	2,547
H_2			51.0	102	5,307	55.3	107.5	5,548	53.1	107	5,525
CH_4			12.8	205	4,414	12.8	205	4,414	12.6	199	4,282
C_2			0.7	21	420	0.7	21	420	0.33	10	197
CO_2			42.8	1,885	0	45.1	1,985	0	2.9	129	0
H_2O			195.6	3,520	0	193.3	3,479	0			
H_2S			1.4	48	310	1.4	48	310			
NH_3			0.65	11	88	0.65	11	88			
INERTS	46,664		1.68	47	0	1.7	47	0			
TOTAL GASES			329.9	6,492	13,377	329.9	6,492	13,336	89.7	1,031	12,550
TARS				51	801						
OILS				5	84						
PHENOLS				10	78						
NAPHTHA				19	357						
TOTAL				6,577	14,696						
COAL/ASH (DAF) (AR)	1,342.854 1,664.612	18,280		150							
*STEAM/H_2O (HP)	3,[illegible]95.982										
OXYGEN	755.661										
FLUX/CARBON	0										
TOTAL	6,648.5	18,280	329.9	6,727	14,696	329.9	6,492	13,336	89.7	1,031	12,550
EFFICIENCY					.804 .578 (net)			.730 .524 (net)			.687 .493 (net)

*In addition, 885,114 saturated gasifier jacket steam used.

Note: All Btu values are in LHV (for coal on as-received basis).

sent for hydrotreatment to produce gasoline blending stock. The phenols separated by the Phenosolvan process are sent to the steam plant for incineration along with the tars and oils. For raw gas production the efficiency defined as the LHV of coal input to gasifier divided by the LHV of total output products is 80.4 percent. The net plant efficiency, defined as the ratio LHV of total plant output to LHV of total coal input, is 57.8 percent. This value represents the energy penalty incurred in the production of steam and oxygen for the gasification process and some of the plant steam requirement.

A portion of the raw gas is diverted to the raw shift conversion unit where the raw gas hydrogen-to-carbon monoxide mole ratio of 2.19 is adjusted to 10.0 so that on remixing with the bulk raw gas stream the overall hydrogen-to-carbon monoxide ratio is 2.54 which is the required ratio for feed to the Synthol Fischer-Tropsch unit. Appendix B describes the computation of the number of moles of raw gas required to be shifted. After shift the gasifier efficiency is 73 percent. The bulk and shifted raw gas stream is then sent to the Rectisol unit for gas purification, i.e., removal of acid gases (CO_2, H_2S, COS) and naphtha. The clean shifted synthesis gas and methane is then sent to the Synthol Fischer-Tropsch synthesis unit. The LHV efficiency at this point is 68.7 percent based on gasifier coal input. Total production of clean shifted synthesis gas by this system is 74 M lb moles/hour. Gasifier methane production is 12.4 M lb moles/hour. This is equivalent to 0.9? x 10^6 Nm^3 per hour of synthesis gas and methane (32.85 x 10^6 SCFH).

Products from Synthol synthesis are shown in Table 4-5. Two process configurations are shown as in Reference 1. The mixed output case co-produces SNG as well as refined liquid products. In the all-liquid case the SNG is autothermally reformed back to synthesis gas and the product spectrum is totally liquid.

As mentioned in Reference 1, the efficiency penalty incurred in reforming the C_1 and C_2 hydrocarbon gases is considerable, as can be realized by the difference in the HHV efficiency (defined as HHV of output divided by HHV of all coal to plant) between the two cases. For the mixed output case the efficiency is 46 percent. For the all-liquid case the efficiency is ?4.8 percent. The total barrels of C_4^+ liquids per ton of dry coal are doubled in the all-liquids case compared to the mixed output case, i.e., from .84 to 1.69. The high methane make in the dry-ash Lurgi, combined with the additional C_1 and C_2 make from the Synthol synthesis, makes this an unattractive combination for an all-liquid Fischer-Tropsch plant

Comparison with the Wyoming coal case, Table 4-6, shows that the gasifier efficiency is greater in the Wyoming coal case for all stages in the production of clean synthesis gas. In terms of net thermal efficiency, Wyoming coal case shows a considerable advantage

TABLE 5-5

PRODUCT SLATE FOR LURGI DRY-ASH/SYNTHOL SYSTEM
(ILLINOIS #6 COAL)

	Mixed			All-Liquid		
	MMBtu/hr Syngas	MMBtu/hr Total	B/SD MMSCF/SD*	MMBtu/hr Syngas	MMBtu/hr Total	B/SD
Gasoline	2,857	[illegible]	15,308	6,007	[illegible]	[illegible]
C_3		199	1,251		418	[illegible]
C_4		[illegible]	163		[illegible]	[illegible]
Diesel		381	2,608		[illegible]	[illegible]
Fuel Oil		166	702		[illegible]	[illegible]
Alcohol		328	2,070		689	[illegible]
Total Liquids		[illegible]	22,202		[illegible]	[illegible]
SNG	[illegible]	[illegible]	180.0*		--	--
Totals		12,047			[illegible]	
FOE			48,188			[illegible]
Efficiency (HHV %)		[illegible]			[illegible]	
B C_4^+/Ton Dry Coal			[illegible]			[illegible]

Syngas [illegible]

Gasifier Naphtha [illegible]

Gasifier Methane [illegible]

TABLE 4-6

COMPARISON OF GASIFICATION EFFICIENCIES WITH WYOMING AND ILLINOIS #6 COALS USING LURGI DRY-ASH GASIFIER

Feed Stock	Raw Gasifier Output	Shifted Raw Gas	Clean Gas
Gasifier Efficiency			
Wyoming	0.899	0.791	0.747
Illinois #6	0.804	0.730	0.687
Net Efficiency			
Wyoming	0.738	0.649	0.613
Illinois #6	0.578	0.524	0.493

over the Illinois #6 coal case, i.e., 61 percent compared to 49 percent. The causes of this efficiency disadvantage with the Illinois coal are attributable to greater steam and oxygen requirements for processing this less reactive bituminous coal.

4.4.2 Plant Construction Costs

For Illinois #6 coal, the plant construction and installation costs for mixed and all-liquid indirect liquefaction plants employing Lurgi dry-ash gasification technology in conjunction with Synthol Fischer-Tropsch synthesis are shown in Table 4-7. For comparison, the construction cost data for a Lurgi Synthol plant using a Wyoming coal are shown.[1] For mixed and all-liquid plants the construction costs for Illinois coal feedstock are 23 percent and 21 percent, respectively; higher than for the plants using Wyoming coal. Synthesis gas preparation costs are higher for the Illinois coal case because more gasifiers are needed to process the less reactive coal (28 for Wyoming compared to 32 gasifiers for Illinois). Gas cooling and shift units have to be larger to handle the greater raw gas volume in the Illinois case. The by-product recovery costs are higher because a larger sulfur recovery unit is required to handle the greater flow volume of acid gases. In addition, much larger gas liquor separation units, Phenosolvan and Chemie-Linz ammonia recovery units, are required to process over twice the quantity of water in the raw gas stream.

TABLE 4-7

CONSTRUCTION COST FOR LURGI DRY-ASH SYSTEMS[1]
(MM 1977 $)

Unit Description \ System	Lurgi Dry-Ash Synthol Wyoming Coal		Lurgi Dry-Ash Synthol Illinois #6 Coal	
	Mixed	All-Liquid	Mixed	All-Liquid
Coal & Ash Handling	71.4	71.4	71.4	71.4
Synthesis Gas Preparation	331.6	331.6	380.5	380.5
Gasifiers	200.7		229.4	
Gas Cooling	19.3		27.3	
Shift	12.8		15.1	
Gas Cleaning	98.8		108.7	
By-Product Recovery	110.4	110.4	161.8	161.8
Gasifier Naphtha Treatment	12.5	12.5	12.5	12.5
Synthesis	104.1	157.1	120.6	196.7
SNG Preparation or Reforming	24.3	40.7	24.9	41.7
F-T Liquid Recovery & Upgrading	92.1	159.6	106.3	173.6
Oxygen Plant	110.1	148.5	156.4	185.0
Steam Plant	195.3	212.9	280.2	293.8
Waste Water Treatment	26.3	30.0	42.0	45.5
Miscellaneous	108.0	108.0	108.0	108.0
TOTALS	1,186.1	1,382.7	1,464.5	1,670.5

(1) Costs are for plants processing 27.8 M tons/day as-received coal.

The synthesis units are larger in the Illinois case to handle the greater quantity of synthesis gas produced in the gasifier. Liquids refining operations are also larger in order to handle the greater liquid output. For example, in the Wyoming case for an all-liquid plant, the total liquid output of C_4^+ liquids is 38,884 BPSD; whereas for the Illinois case it is 44,743 BPSD, a 15 percent increase.

The larger steam and oxygen plants required in the Illinois case reflects the greater oxygen and steam-to-coal ratios required in the gasification of less reactive Illinois coal.

4.4.3 Product Costs

Product costs were computed for all the systems in this report by using the same procedure as in Reference 1. No separate DCF analysis was performed; instead the value of \$7.06 per MMBtu obtained in the MRDC report[3] based on equity financing and 12 percent DCF was used as the baseline capital derived cost. This figure was for a plant construction cost of \$1,186.1 and a HHV output of 11,238 million Btu's per hour. All costs, excluding coal costs, but including operating and maintenance costs, are considered to be proportional to construction costs. Since the MRDC report used last quarter 1977 dollars, the computed product costs will be in 1977 dollars. Revision of these costs to 1980 dollars by appropriate inflators has been inc[illegible] the Executive Summary. Coal-derived costs were determined on a raw coal cost of $20.38 per ton (first quarter 1978 dollars) entering the plant. Costs of products on a thermal basis were then computed as follows:

Thermal cost $/MM Btu output	=	7.06 x construction cost factor x HHV output factor + 1.96 Y
Where construction cost factor	=	plant construction cost/1,186.1
and HHV output factor	=	11,238/HHV plant output
and Y	=	12,047/HHV plant output

Market basis costs are computed in exactly the same way as in Reference 1. For details, see Reference 1, Appendix D.

Table 4-8 shows product cost data for the dry-ash Lurgi Synthol system using Wyoming and Illinois #6 coal as feedstocks.

The combination of high construction costs and expensive coal with low thermal efficiency and, hence output, for the Illinois coal case results in high gasoline product costs for these plants, compared to the plants using the lower cost, more reactive Wyoming coal. This higher cost results in spite of the advantage of scale obtained when using the Illinois coal. For the Illinois case mixed-output mode, coal feedstock cost accounts for 19 percent of the product cost and for Wyoming coal it accounts for 9 percent of the product cost.

TABLE 4-8

COST DATA FOR LURGI DRY-ASH/SYNTHOL SYSTEM*

Feedstock	Mode	Construction Cost (MM $)	HHV Output (MMBtu/hr)	Capital Derived Cost ($/MMBtu)	Total Cost ($/MMBtu)	Gasoline Cost ($/Gallon)	
						Thermal	Market
Illinois #6	M	1,464.5	12,047	8.13	10.09	1.21	2.14
	AL	1,670.5	9,114	12.26	14.85	1.78	1.91
Wyoming	M	1,186.1	11,238	7.06	7.78	0.93	1.33
	AL	1,382.7	8,413	10.99	11.95	1.43	1.51

M = Mixed Output

AL = All Liquid Output

*Costs are for plants processing 27.8 M tons/day as-received coal.

4.4.4 BGC Slagging Lurgi Systems

The indirect liquefaction plant using Illinois #6 coal as feedstock to the BGC slagging Lurgi gasifier will process 27,800 tons per stream day of coal in 14 gasifiers with two spares. In the Wyoming case, 12 gasifiers with two spares were required.

Of the 27,800 tons per day of coal entering the plant, 23,335 tons are sent to gasification along with 1,240 tons of flux while 4,465 tons are sent to the steam plant. Computation of plant coal split is provided in Appendix A. The steam plant provides steam for oxygen production, gasification steam and downstream plant unit steam requirements. In addition to the fine coal used for steam production, oils and phenols produced via gasification are also sent to the steam plant. Net tar production from gasification is zero since all the tar is gasified. The injection tar is fed through the tuyeres into the slagging zone of the gasifier and the recycled tar is fed to the top of the gasifier. Oxygen requirement for gasification is 968,540 pounds per hour, while the high pressure steam requirement is 689,740 pounds per hour. On a DAF coal basis, the steam-to-coal weight ratio is 0.44, the oxygen-to-coal weight ratio is 0.62, and the steam-to-oxygen mole ratio is 1.27. Notice the substantial savings in the steam required when using the BGC gasifier over the dry-ash Lurgi.

Table 4-9 shows the material and energy fluxes at each stage in the production of clean synthesis gas for the BGC Lurgi/Synthol system. The raw gas processing scheme is the same as for the dry-ash Lurgi/Synthol system except that, because of the low hydrogen-to-carbon monoxide ratio of the gasifier off-gas, all of the raw gas stream is sent to shift. This necessitates a very large shift unit to handle this volume of gas.

The gasification of 1,945 M pounds per hour of as-received coal produces 116 M pounds per hour of synthesis gas and 8 M pound moles per hour of methane. Gasifier and net efficiencies are shown in Table 4-9. Comparison of LHV efficiencies of the BGC and the Lurgi dry-ash systems shows that for gasifier efficiency the BGC production of clean syngas is 74 percent--for dry-ash it is 69 percent. Comparing net LHV efficiencies of the two processes for production of clean synthesis gas, the BGC efficiency is 62 percent, whereas for the dry-ash gasifier, the efficiency is only 49 percent. Comparison of LHV efficiencies for clean syngas production using the BGC with Wyoming coal as feedstock are shown in Table 4-10.

Clearly, even when the BGC Lurgi is used, higher efficiencies are achieved when using the more reactive, low rank Wyoming coal rather than the bituminous Illinois #6. The effect of the coal feedstock is not as pronounced with BGC as it is when using the dry-ash gasifier.

TABLE 4-9

MATERIAL AND ENERGY FLUX IN SYNTHESIS GAS PREPARATION UNITS
FOR BGC-LURGI/SYNTHOL SYSTEM
(ILLINOIS #6 COAL)

COMPONENT	INPUTS		RAW GASIFIER OUTPUT			SHIFTED RAW GAS			CLEAN GAS TO SYNTHESIS		
	Mlb/hr	MMBtu/hr	MlbMoles/hr	Mlbs/hr	MMBtu/hr	MlbMoles/hr	Mlbs/hr	MMBtu/hr	MlbMoles/hr	Mlbs/hr	MMBtu/hr
CO			80.675	2,260	9,823	32.793	919	3,993	32.7	915	3,977
H_2			35.411	71.4	3,685	83.293	168	8,668	82.86	167	8,633
CH_4			8.397	135	286	8.397	135	2,896	8.145	130.6	2,809
C_2			0.831	25	499	0.831	25	499	0.44	14.1	282
CO_2			8.883	391	0	56.765	2,498	0	3.69	162	0
H_2O			45.28	815	0	20.579	370	0			
H_2S			2.659	91	589	2.659	91	589			
NH_3											
INERTS	11		0.656	18	0	0.656	18	0			
TOTAL GASES			182.792	3,805.6	17,491	205.97	4,223	16,645	127.9	1,388.7	15,701
TARS				(a)	0						
OILS				32.7	546						
PHENOLS				9.1	71						
NAPHTHA				22.3	419						
TOTAL				3,869.7	18,528	205.97	4,233	16,645	127.9	1,388.7	15,701
COAL/ASH (DAF) (AR)	1,568.7 1,944.6	21,355									
STEAM/H_2O	689.74					23.0					
OXYGEN	968.54										
FLUX/CARBON	103.12										
TOTAL	3,753	21,355	182.792	3,869.7	18,523	205.97	4,233	16,645	127.9	1,388.7	15,701

EFFICIENCY	RAW GASIFIER OUTPUT	SHIFTED RAW GAS	CLEAN GAS TO SYNTHESIS
	.868 .728 (net)	.779 .652 (net)	.735 .617 (net)

(a) Tars recycled to gasifier. Material balance not good because of CO_2 used for lock-hopper compression and water quench streams.

TABLE 4-10

COMPARISON OF GASIFICATION EFFICIENCIES OF BGC-LURGI USING WYOMING AND ILLINOIS #6 COALS

	Wyoming	Illinois #6
Gasifier Efficiency (LHV)	0.774	0.735
Net Efficiency (LHV)	0.692	0.617

Where:

$$\text{Gasifier Efficiency} = \frac{\text{LHV of clean Syngas}}{\text{LHV of coal to gasifier}}$$

$$\text{Net Efficiency} = \frac{\text{LHV of clean Syngas}}{\text{LHV of coal to plant}}$$

The BGC Lurgi gasification system was coupled both with the Synthol units and the Kolbel[11] slurry phase synthesis units in this analysis. Kolbel synthesis was considered to be a potentially compatible candidate process for BGC since BGC produces a synthesis gas having a low hydrogen-to-carbon monoxide mole ratio in the order of 0.5. The Kolbel synthesis is capable of accepting low hydrogen-to-carbon monoxide mole ratios of about .67. Thus, less shift is required when combining BGC with Kolbel. For Synthol, the low hydrogen-to-carbon monoxide synthesis gas from BGC must be shifted extensively to achieve the 2.54 ratio requirement of the Synthol unit.

In the Synthol case, all of the raw gas stream from the gasifier is passed to the raw gas shift reactor to obtain the correct hydrogen-to-carbon monoxide ratio. For coupling with Kolbel, only 29 M pound moles per hour, or 16 percent of the total raw gas, needs to be shifted to a hydrogen-to-carbon monoxide ratio of 10 and then recombined with the stream to produce the required 0.67 ratio for the Kolbel system. This represents a considerable cost savings on the size of shift reactor and also a smaller sulfur recovery unit is required to process the reduced flow of acid gases.

Table 4-11 shows the product distribution and quantities of refined outputs for the BGC Synthol mixed output and all-liquid cases. In the all-liquid case total liquids produced per ton of dry coal

TABLE 4-11

PRODUCT SLATE FOR BGC-LURGI/SYNTHOL SYSTEM
(ILLINOIS #6 COAL)

	Mixed			All-Liquid		
	MMBtu/hr Syngas	MMBtu/hr Total	B/SD MMSCF/SD*	MMBtu/hr Syngas	MMBtu/hr Total	B/SD
Gasoline	4,478	4,910	23,460	7,574	8,006	38,255
C_3		312	1,961		527	3,317
C_4		46	256		78	453
Diesel		910	4,088		1,540	6,915
Fuel Oil		260	1,100		440	1,860
Alcohol		514	3,245		869	5,489
Total Liquids		6,953	34,111		11,461	56,269
SNG		7,391	176.9*		--	--
Totals		14,344			11,461	
FOE			57,375			45,844
Efficiency (HHV %)		54.8			43.8	
B C_4^+/Ton Dry Coal			1.29			2.12

Syngas = 116 1000# moles

Gasifier Naphtha = 432 Btu/hr

Gasifier Methane = 3,372 Btu/hr

exceeds two barrels. The HHV efficiencies are much higher than for the dry-ash Lurgi case where an all-liquid HHV of 35 percent was obtained compared to 44 percent using the slagging Lurgi system.

Table 4-12 shows the products obtained from the BGC Lurgi and Kolbel synthesis. Here, the mixed output HHV efficiency is almost identical to the Synthol system, but the all-liquid efficiency of the Kolbel plant is 8 percent higher than the corresponding Synthol case because less methane is produced in the Kolbel unit.

4.4.5 Plant Construction Cost

Table 4-13 shows the elements of plant construction costs for both the BGC Lurgi/Synthol and Kolbel systems. Coal handling costs have increased because of the use of flux with this slagging gasifier. Per stream day, 1,240 tons of lime are added to the coal in this coal handling unit.

For all of the BGC cases using Illinois coal, gasification costs are based on the use of 16 gasifiers in comparison to 14 for the Wyoming case. The larger oxygen plant requirement for the all-liquid cases results from the autothermal reforming of C_1 and C_2 gases; 1.4 M pounds of oxygen is required per MMSCF of methane that is reformed. Energy for this extra oxygen is provided by slip streaming 10 percent of the stream to be reformed to plant steam generation. The larger steam plant requirement in all-liquid cases is a result of this additional oxygen demand.

TABLE 4-12

PRODUCT SLATE FOR BGC-LURGI/KOLBEL SYSTEM
(ILLINOIS #6 COAL)

	Mixed			All-Liquid		
	MMBtu/hr Syngas	MMBtu/hr Total	B/SD MMSCF/SD*	MMBtu/hr Syngas	MMBtu/hr Total	B/SD
Gasoline	7,357	7,789	37,217	9,545	9,977	47,668
C_3		591	3,721		767	4,827
C_4		0	0		0	0
Diesel		1,026	4,609		1,332	5,979
Fuel Oil		114	480		147	623
Alcohol		118	745		153	967
Total Liquids		9,639	46,772		12,376	60,064
SNG		4,487	107.4*		--	--
Totals		14,126			12,376	
FOE			56,504			49,504
Efficiency (HHV %)		53.9			47.3	
B C_4^+/Ton Dry Coal			1.73			2.21

Syngas = 116 1000# moles

Gasifier Naphtha = 432 Btu/hr

Gasifier Methane = 3,372 Btu/hr

TABLE 4-13

CONSTRUCTION COST FOR BGC-LURGI SYSTEMS*
(ILLINOIS #6 COAL)
(MM 1977 $)

Unit Description \ System	BGC-Lurgi Synthol Illinois #6 Coal			BGC-Lurgi Kolbel Illinois #6 Coal		
	Mixed		All-Liquid	Mixed		All-Liquid
Coal & Ash Handling		79.2	79.2		79.2	79.2
Synthesis Gas Preparation		308.5	308.5		261.8	261.8
Gasifiers	114.7			114.7		
Gas Cooling	16.2			11.0		
Shift	40.1			18.4		
Gas Cleaning	137.5			117.7		
By-Product Recovery		101.7	101.7		67.2	67.2
Gasifier Naphtha Treatment		12.5	12.5		12.5	12.5
Synthesis		162.8	231.0		176.1	213.8
SNG Preparation or Reforming		24.7	41.3		17.4	29.1
F-T Liquid Recovery & Upgrading		143.6	203.8		172.4	205.4
Oxygen Plant		185.6	218.0		185.6	205.8
Steam Plant		189.9	215.1		189.9	201.7
Waste Water Treatment		18.8	23.1		9.9	7.8
Miscellaneous		108.0	108.0		108.0	108.0
TOTALS		1,335.2	1,542.2		1,279.0	1,392.3

*Costs are for plants processing 27.8 M tons/day as-received coal.

For the BGC Kolbel combination, the smaller shift requirement means a smaller shift reactor and Rectisol gas purification unit since less carbon dioxide gas is produced. In addition, the by-product recovery systems can be reduced in size because of smaller acid gas streams. These changes represent a cost savings over Synthol plants of 11 percent in the all-liquid cases.

4.4.6 Product Costs

Product costs from the BGC plants with Synthol and Kolbel systems are shown in Table 4-14. For an all-liquid output, Kolbel synthesis could potentially reduce gasoline cost over the Synthol case on a market basis by 19 percent. Raw material costs of coal and flux (purchased at $13/ton) for the four plant configurations shown in Table 4-14 account for, on an average, 21 percent of the product cost. For Wyoming coal feedstock using BGC gasification, coal costs account for, on average, 10 percent of product cost. On a market basis, for an all-liquid plant, gasoline cost in $/gallon is, on average, 13 percent higher when using Illinois #6 coal than for a Wyoming coal system processing the same tonnage of as-received coal.

TABLE 4-14

COST DATA FOR BGC-LURGI SYSTEMS*
(ILLINOIS #6 COAL)
(Last Quarter 1977 $)

System	Mode	Construction Cost (MM $)	HHV Output (MMBtu/hr)	Capital Derived Cost ($/MMBtu)	Total Cost ($/MMBtu)	Gasoline Cost ($/Gallon)	
						Thermal	Market
BGC-Lurgi Synthol	M	1,335.2	14,344	6.2[illegible]	7.90	0.95	1.20
BGC-Lurgi Synthol	AL	1,542.2	11,461	9.00	11.10	1.33	1.39
BGC-Lurgi Kolbel	M	1,279.0	14,126	6.06	7.76	0.93	1.05
BGC-Lurgi Kolbel	AL	1,392.3	12,376	7.53	9.47	1.13	1.17

*Costs are for plants processing 27.8 M tons/day as-received coal.

Notes: M = Mixed output
AL = All-Liquid output