

4.5 Cost of Products Produced

MRDC⁽¹⁾ has calculated costs of products produced by the SASOL-U.S. plant on the basis of \$/MMBtu of thermal product, and on the basis of product costs for the multiple products produced. In the later case, the cost of gasoline has been derived by estimating prices for all other products, and subtracting this amount from the total product income required to meet DCF requirements. Non-gasoline product prices, and the basis of their estimation, are shown in Table IV-7.

The value assigned to SNG is of particular importance since this product is the dominant output of the SASOL-U.S. Base Case. Relatively small changes in SNG value will result in relatively large changes in the required price for the liquid products. The price of \$6.17/MMBtu of SNG is an estimate of the required selling price of SNG produced by a plant employing Lurgi gasification to produce SNG only. It thus represents a reasonable upper limit on SNG value; the actual value could be dramatically less.

Gasoline costs were computed by Mobil on an equity basis assuming a 12 percent DCF and on a utility cost basis assuming a debt/equity ratio of 75/25, 9 percent interest on the debt capital, and 15 percent return after taxes on equity capital. A six year construction period and a 20 year useful life are assumed in both cases. Table IV-8 shows the gasoline prices for the assumptions of equity financing.

TABLE IV-7
 PRICES ASSUMED FOR PRODUCTS OTHER THAN GASOLINE⁽¹⁾

<u>PRODUCT</u>	<u>UNIT</u>	<u>\$/UNIT</u>
SNG & LPG*	MMBtu	6.17
Butanes	MMBtu	Gasoline - 30¢/MMBtu
Diesel	BBL	Gasoline - \$1.70/B
Fuel Oil	BBL	Gasoline - \$3.50/B
Alcohols	lb	15¢/lb
Sulfur	Ton	\$25/Ton
Ammonia	Ton	\$155/Ton

*Based on production cost at plant designed to product SNG ONLY

TABLE IV-8
 COST OF GASOLINE FROM SASOL-U.S. PLANT⁽¹⁾
 (Oct. 1977 Dollars)

INVESTMENT, MM\$	
Onsite	826
Offsite	713
Other	348
TOTAL	1,887
WORKING CAPITAL, MM \$	60
COAL COST, MM\$/YR	65
OPERATING COST, MM\$/YR	152
BYPRODUCT CREDIT, MM\$/YR	6
UNIT COST	
Thermal Product	
\$/MMBtu	7.78
Gasoline, ¢/gal*	93
Multiple Products	
Gasoline, ¢/gal	133

*At 5.0 MMBtu/B

On a thermal product basis (e.g., all products valued in terms of cost per Btu) gasoline prices are seen to be 93¢ per gallon on an equity basis. Note, however, that the product cost of \$7.78 per 10^6 Btu is substantially higher than the \$6.17 cost of a plant designed to produce only SNG. If the SNG, which made up nearly two-thirds of the Btu output of the SASOL-U.S. plant, must be sold for these lower values, then a substantial premium must be added to the price of the liquid produced in order to meet DCF requirements. Thus, when products other than gasoline are valued in accordance with the schedule given in Table IV-7, then gasoline prices are \$1.33 per gallon.

The elements which make up the product cost for the case of equity financing are shown in Table IV-9. Coal costs are seen to account for only 9% of the final cost of the output products. The assumed coal cost of \$7/ton (or received) could more than double without increasing the product cost 10 percent. The largest component, capital charges, account for 65%. The start-up penalty of 5% is equivalent to unrecovered capital charges during the first year of operation during which only 50% of rated plant output is assumed to be achieved.

The largest operating expense other than coal (maintenance and local taxes) are directly proportioned to the capital cost of the plant. Thus to a high degree of accuracy, all elements of product cost other than coal cost may be assumed to be proportional to the plant investment. This assumption has been made in the computation

TABLE IV-9

ELEMENTS OF PRODUCT COST

(%)

Operating Expense	21
Coal Cost	9
Start-Up Penalty	5
Capital Charges	65
	<hr/>
	100

of product costs for the all-liquid output case discussed subsequently, and for all other plant modifications considered in this report.

4.6 All-Liquid Output

The results presented previously concluded that a SASOL-U.S. plant can produce a mixed output for total product cost of the order of \$7.78/MMBtu. The resulting product price of \$.93 per gallon for gasoline (1977\$) is attractive, however, two thirds of the output is SNG for which the \$7.78/MMBtu price is extremely high. Even if the SNG can be sold for the relatively high price of \$6.17/MMBtu, the gasoline price required to meet DCF requirements increases to \$1.33 per gallon. As an alternative to marketing the SNG, it may be advantageous to structure the plant in a manner which would reform the SNG to produce an all-liquid output.

The referenced MDRC study did not investigate a SASOL-type plant producing an all-liquid product. They did, however, investigate an alternative MTG (methanol to gasoline) plant in which all methane produced was reformed to produce additional synthesis gas, resulting in an all-liquid output (MRDC Case 1B in Reference 1). The methane was assumed to be reformed by partial oxidation in an "Autothermal Reformer" of Lurgi design.

Mobil states that the system resulted in "about 61%" of the thermal contents of the methane formed in the mixed output case being recovered as liquid fuel. Detailed analysis of results show that the actual recovery in HHV was 62.3% in Mobil Case 1B.

As an alternative to a detailed analysis of reforming and recycle flows, we have assumed that a SASOL-U.S. plant incorporating an Autothermal reformer would result in 61% of thermal content of the SNG produced in the mixed output case being recovered as liquids. This assumption is believed to be reasonable since the efficiency of the Autothermal unit (about 83%) would be similar in both cases and since the efficiency of F-T synthesis and upgrading is only about 1% less than the alternative MTG system.

Liquids produced in the all liquid case occur in the same proportions as in the mixed product case, since the basic synthesis and refining system is unchanged. Figure 4-7 shows the revised process flow and gives the thermal efficiency for each major process step.

The major difference from the mixed output case are in the Synthesis section, which has an overall efficiency of 67.4 percent vice 86.8 for the mixed product case. The low value occurs because of losses in the autothermal reformer, and because all output is subject to Synthol synthesis losses.

A purge gas equivalent to 9% of the output of the autothermal reformer is removed to prevent build up of contaminants. This gas is burned in the steam plant and permits the additional steam required to produce oxygen for the autothermal reformer to be generated without increasing coal usage.

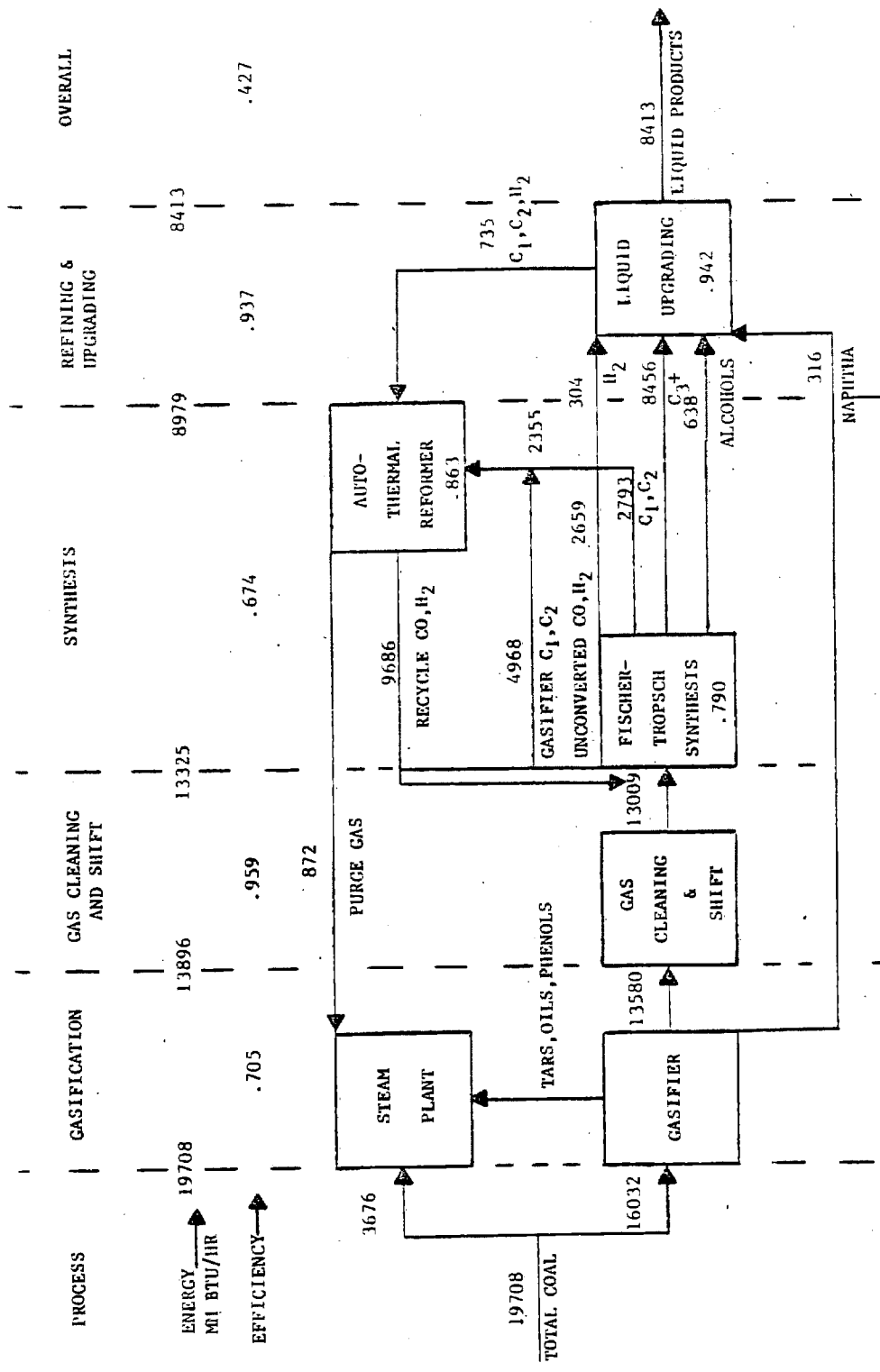


FIGURE 4-7
SYSTEM ENERGETICS SASOL-U.S. (ALL LIQUID)
(UNITS MM BTU/HR)

The plant modifications required to produce an all-liquid product in this manner are:

1. Add the Autothermal Reformer
2. Increase the capacity of the oxygen plant to supply the Autothermal Reformer
3. Increase the capacity of the steam plant to provide power for the enlarged oxygen plant
4. Delete methanation and SNG preparation
5. Enlarge the F-T synthesis and upgrading units to handle the increased flow
6. Enlarge waste water treatment facilities to accommodate increase waste water from enlarged Synthol reactors

Table IV-10 shows the estimated construction costs for the required changes. Estimates for individual process units are scaled from the base case using a scale factor of 0.7 based on capacity required. Details are shown in Appendix C.

Table IV-11 compares the output of the all-liquid plant with the base case. Since about two thirds of the thermal output of the mixed output case was in the form of SNG, a substantial increase in liquid output is to be expected if the SNG is converted to liquid. However, since only 61% of the energy in the SNG is recovered, the overall efficiency is seen to drop markedly from 57.3 in the mixed output case to 42.7 in the all-liquid case.

The gasoline prices for equity financing are given in Table IV-12. Because of the higher capital cost and lower efficiency of

TABLE IV-10
 IMPACT ON CONSTRUCTION COST FOR ALL LIQUID OUTPUT
 (MM \$1977)

REFERENCE CONSTRUCTION COST (SASOL-U.S. Mixed Output)	Relative to <u>Mixed Output Case</u>	
Synthesis	+53.0	
Add Autothermal Reformer	+40.7	
Delete SNG Preparation and Methanator	-24.3	
F-T Liquid Product Upgrading	+67.5	
Oxygen Plant	+38.4	
Steam Plant	+17.6	
Waste Water Treatment	+ 3.7	
TOTAL CHANGE		196.6
ALL-LIQUID OUTPUT		1,382.7
CAPITAL COST		2,198.5
		1,186.1

TABLE IV-11

COMPARISON OF ALL LIQUID AND MIXED OUTPUT PLANTS

	<u>Mixed Output</u>	<u>All-Liquid Output</u>
SNG MM SCF/D	173.3	-
Gasoline B/D	13,580	28,090
C ₃ B/D	1,107	2,436
C ₄ B/D	146	321
Diesel B/D	2,307	5,078
Fuel Oil B/D	622	1,369
Alcohol B/D	1,829	4,026
Total Liquids B/D	19,591	41,320
FOE B/D	44,950	33,652
Efficiency (HHV)	57	42.7
Liquid Fuels/Ton Dry Coal (B)	.98	2.07
Liquid Fuels C ₄ ⁺ /Ton Dry Coal (B)	.92	1.94

TABLE IV-12
 GASOLINE COSTS FOR MIXED PRODUCT
 AND ALL LIQUID SASOL U.S. PLANTS

	<u>Mixed Output</u>	<u>All Liquid Output</u>
THERMAL PRODUCT BASIS		
\$/MMBtu	7.78	11.95
\$/Gal	.93	1.43
MARKET BASIS*		
\$/Gal	1.33	1.51

* Assumes products other than gasoline sold at prices given in Table IV-7.

the all-liquid plant, the gasoline price for the all-liquid plant is higher than must be charged in the mixed output case if the SNG produced can be sold for \$6.17/10⁶ Btu.

4.7 Conclusions

The SASOL-U.S. plants for mixed product and all-liquid production of synthetic fuels by indirect processes are believed to be representative of low risk plants which could be designed using present commercial technology. Lurgi, who have had extensive experience at SASOL, participated in the MRDC effort on which the designs are based. The MRDC report provides highly detailed material balances which are very useful. The unit-by-unit estimates of plant construction costs provide an excellent basis for estimating comparative costs for alternative plants. However, detailed heat balances (other than for overall input/output) were not provided.

The SASOL-U.S. designs follow the conservative design approach employed at SASOL and may not represent the highest level of technical or economic performance that could be achieved with currently proven technologies. Cost estimates for coal handling, gasification, gas purification and by-product recovery are substantially higher than have been estimated in other recent studies of U.S. based coal gasification complexes of similar scale. (7,8) Steam requirements also appear high presumably indicating that waste heat management has not been optimized. The overall steam requirement for the mixed product SASOL-U.S. plant is 36 percent greater than for

the corresponding Mobil methanol to gasoline (MTG) process described in Reference 1. This difference appears anomalous since the dominant steam requirement (for gasification and oxygen separation) are identical. In point of fact, one might anticipate a lower steam requirement for the SASOL-U.S. plant since the two processes are similarly exothermic and the MTG process requires substantial compression of gases to achieve the 1160 psig requirement of the methanol synthesis process. None the less, in the designs produced by MRDC, steam produced by the MTG system is of substantially higher quality than that produced in F-T synthesis. This had the effect of reducing steam plant requirements from 18 percent of total coal to only 13 percent. This reduction in steam coal requirement is responsible for virtually all of the overall thermal efficiency advantage of the MTG system shown in MRDC analysis. It is likely that comparable steam coal savings would be possible in the SASOL-U.S. plant if optimum waste heat recovery systems were employed.

The possible deficiencies in the SASOL-U.S. designs do not detract from their suitability for use in the present study. The objective is to show the relative advantages of substituting advanced gasifier and synthesis systems in the SASOL-U.S. design. The level of overall design conservatism will not influence the relative results if care is taken to maintain a similar level of conservation in the alternative designs.

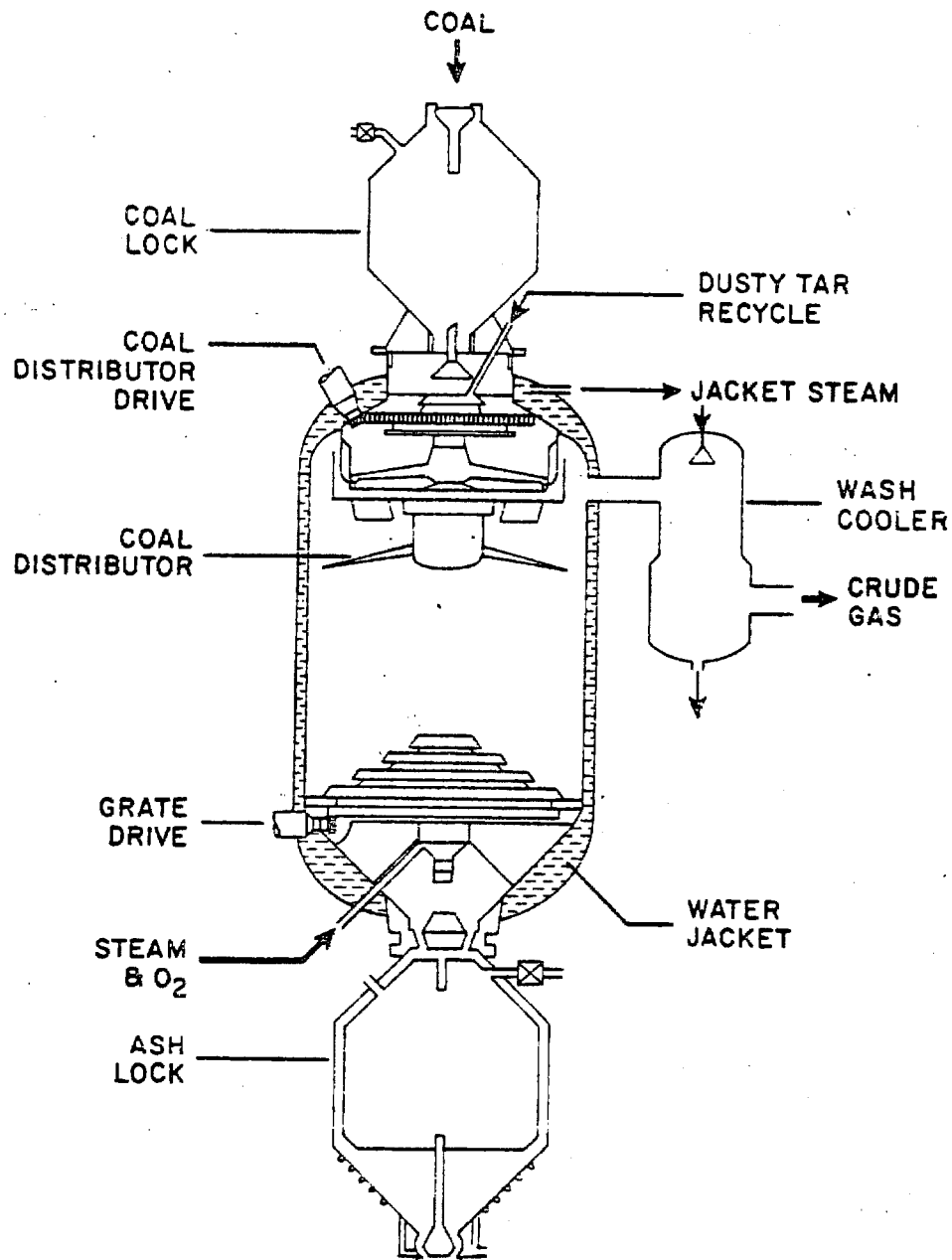
5.0 GASIFICATION FOR INDIRECT LIQUEFACTION

5.1 Introduction

The dry bottom Lurgi gasifier used at SASOL II has several significant limitations. The gasifier, shown schematically in Figure 5-1, is the fixed-bed type. Coal is crushed to approximately 1/4" to 2" in size and introduced to the gasifier through a lock hopper at the top. The reactants, steam and oxygen, are introduced at the bottom. The coal is reacted while moving downward under gravity through the reaction zone.

The unreacted ash is collected on a rotating grate at the bottom of the gasifier, from which it spills to a lock hopper for removal. Coal particles smaller than about 1/4 inch cannot be used because they are swept away with the product gas without reacting. Conventional grinding of run of mine coal usually results in 20 percent or more of the ground coal being too fine for Lurgi gasification. At SASOL these fines are used for steam generation. A second limitation is that the coal must be non-caking. Caking coals will agglomerate when heated to their softening point, and choke the reaction zone with an unreacted "clinker."

The dry bottom Lurgi requires a quantity of steam substantially in excess of the quantity required for efficient reaction. The excess steam is required to prevent overheating of the grate and/or fusion of the ash. This results in a wet product gas, and a relatively low reaction temperature which encourages the formation of



**FIGURE 5-1
LURGI GASIFIER**

methane.⁽¹⁰⁾ In the SASOL-U.S. plant described in Section 4.0, about 19% of the carbon in the coal is converted to methane. Methane formation is a disadvantage if a liquid product is desired, since the reformation of the methane to produce additional synthesis gas increases equipment costs and results in energy losses.

A final limitation of the dry bottom Lurgi is the production of tars, oils and phenols. These are produced because the counter-flow of coal and product gases results in the coal being heated to devolatilization temperature above the reaction zone. The volatiles, which contain the tars, oils and phenols, never reach the reaction zone. These products are undesirable because they are expensive to recover from other products and liquor, and have limited commercial value. They also foul heat exchanger surfaces and thus prevent efficient recovery of the sensible heat in the product gas in waste heat boilers.

The excess steam used in the Lurgi gasifier results in a product gas having a molar H_2/CO ratio of about 2:1. This substantially reduces the amount of shift (e.g., $H_2O + CO = H_2 + CO_2$) which must be performed in order to produce the H_2/CO ratio of 2.5 required by conventional Fischer-Tropsch or Methanol synthesis reactors. As was noted in the discussion of Figure 4-2, only about 20% of the product gas need be shifted in order to produce the required H_2/CO ratio when the shifted and unshifted streams are combined. Furthermore, no additional steam need be added to accommodate

the shift. Indeed, only a small percentage of the excess steam in the product gas is required for the shift reaction. Most is removed when the gas is cooled. The recovery of useful products contained in the condensate, and the subsequent treatment of the large quantity of wastewater, add to plant capital and operating costs.

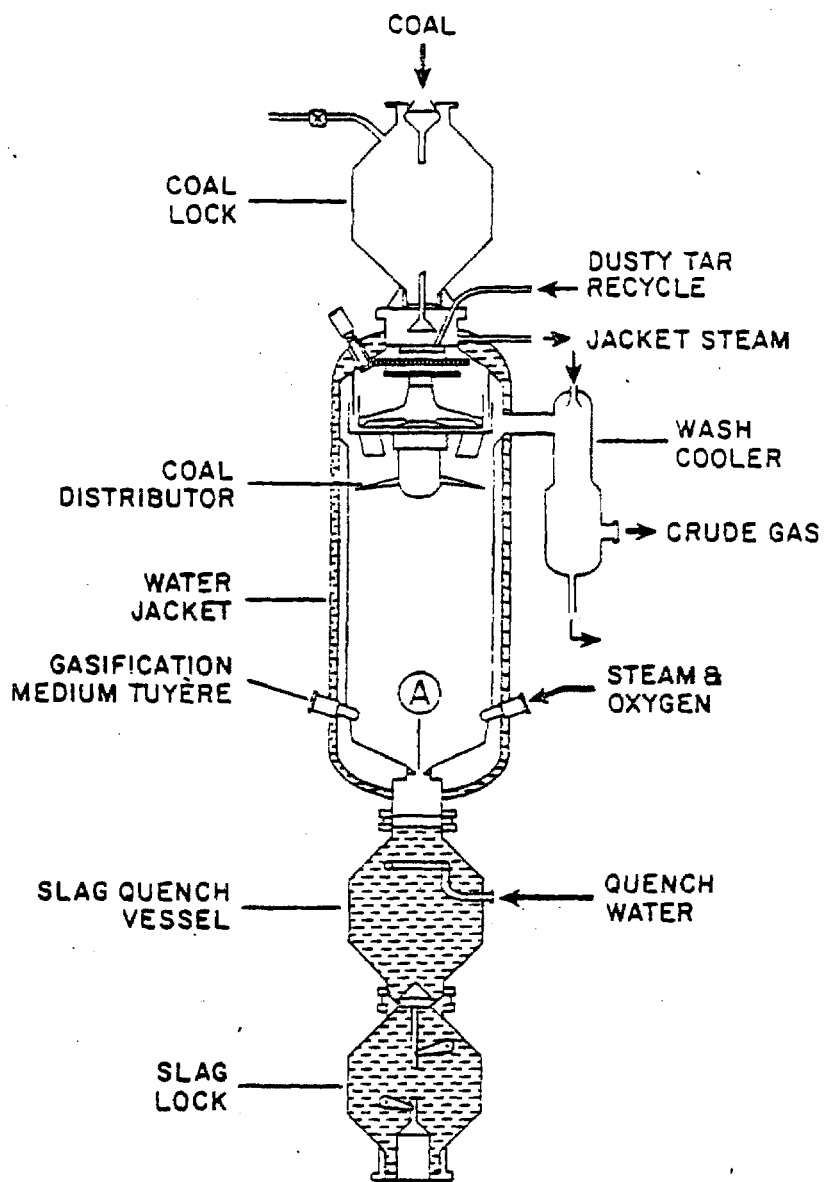
The primary advantage of the Dry Bottom Lurgi gasifier is that it is the only gasifier which has had extensive commercial experience on a large scale. Gasifiers of later design which circumvent some or all of the limitations of the Dry Ash Lurgi are discussed below.

5.2 The BGC-Lurgi Gasifier

5.2.1 Description

The British Gas Corporation (BGC) has conducted research directed toward improving the Lurgi gasifier intermittently over the past 40 years. (11,12,13) By the late 1960's, when work was terminated because of the discovery of large natural gas reserves in the North Sea, BGC had successfully demonstrated a modified Lurgi having several advantages over the original model. This gasifier is shown schematically in Figure 5-2.

The most significant modification is in the manner in which the ash is handled. The BGC design, frequently referred to as the "slagging Lurgi" permits the ash to reach melting temperature and removes it as a molten slag. The slag is removed through a lock hopper incorporating a water quench. The excess steam which the dry ash Lurgi required to keep the ash below fusion temperature



**FIGURE 5-2
SLAGGING GASIFIER**

is no longer required. The reduced steam requirement results in a gasification reaction which approaches the thermochemical optimum for efficiency.

The BGC design incorporates a stirring blade which permits mildly agglomerating coal to be processed. This device could presumably be adapted to the Dry Ash Lurgi.

The gas composition and other operational parameters of the BGC-Lurgi are compared to the Dry Ash Lurgi in Table V-1. The BGC design requires .36 pounds of steam per pound of DAF coal compared to 1.34 for the Dry Ash Lurgi. As a result, net thermal efficiency is improved from 70.1% to 77.7%, and a much dryer gas is produced. Virtually all steam supplied is reacted. The moisture in the product gas comes from the moisture in the coal which is removed when the coal is heated and devolatilized prior to reaction. For this reason, moisture in product gas from 28% moisture Wyoming Sub-bituminous coal assumed herein is somewhat higher than was obtained in the tests with dryer coals. (10,13) Appendix A gives details of how the gas compositions shown in Table V-1 were derived.

The percent of carbon which is converted to methane is reduced from about 19% to about 10%. None the less, methane is high compared to entrained flow gasifiers.

The mole ratio of H_2 to CO for the BGC design is .5 compared to 2.0 for the Dry Bottom Lurgi. However, the steam which must be added to shift the BGC output to a H_2/CO ratio of 2.5 is small

TABLE V-1

LURGI AND BGC/LURGI OPERATIONAL PARAMETERS
(Wyoming Subbituminous Coal)

RAW GAS COMPOSITION (MOLE %)	LURGI	BGC/LURGI
H ₂	23.0	21.9
CO	11.1	43.8
CO ₂	17.6	2.1
CH ₄	6.7	5.2
H ₂ O	41.0	26.3
Miscellaneous Gases	0.3	
C ₂ H ₆	0.3	0.3
H ₂ /CO	2.06	0.5
Steam/Coal	1.34	0.36
Oxygen/Coal	0.36	0.39
THERMAL EFFICIENCY (LHV)*		
Clean Gas	81.5	87.7
Clean Fuel	89.4	95.6
Net	70.1	77.7

* As defined by Shinnar⁽¹³⁾

compared to the excess steam in the Dry Ash Lurgi output. Because the temperature in the reaction zone of the BGC is higher, excess coal fines, as well as tars, oils and phenols, can be gasified by introducing them into the bottom of the gasifier through tuyeres. Experiments funded by EPRI⁽¹⁴⁾ have shown that bottom injections as high as 25% of total coal throughput can be used without compromising the operability or controllability of the gasifier.

A final and very significant advantage of the BGC design is the higher rate of coal throughput which can be processed. The slagging design will process about four tons per hour per square meter of cross section, vs. about .1 to 1.0 tons per hour for early Dry Ash Lurgi, and about 2.0 for the most modern units at SASOL II.

Proponents of this gasifier at British Gas Corporation claim that there are no known disadvantages to the BGC Slagger relative to the Dry Ash model other than the reduced methane make. In this particular case the lower methane make is in fact an advantage.

Although half scale tests of the BGC Slagger have been highly successful, operation in a truly commercial application has not yet been demonstrated. This lack of commercial experience is the primary disadvantage of the BGC relative to the proven Dry Ash Lurgi design.