Figure 29b

NET THERMAL EFFICIENCY (APPROXIMATE WESTERN COAL)

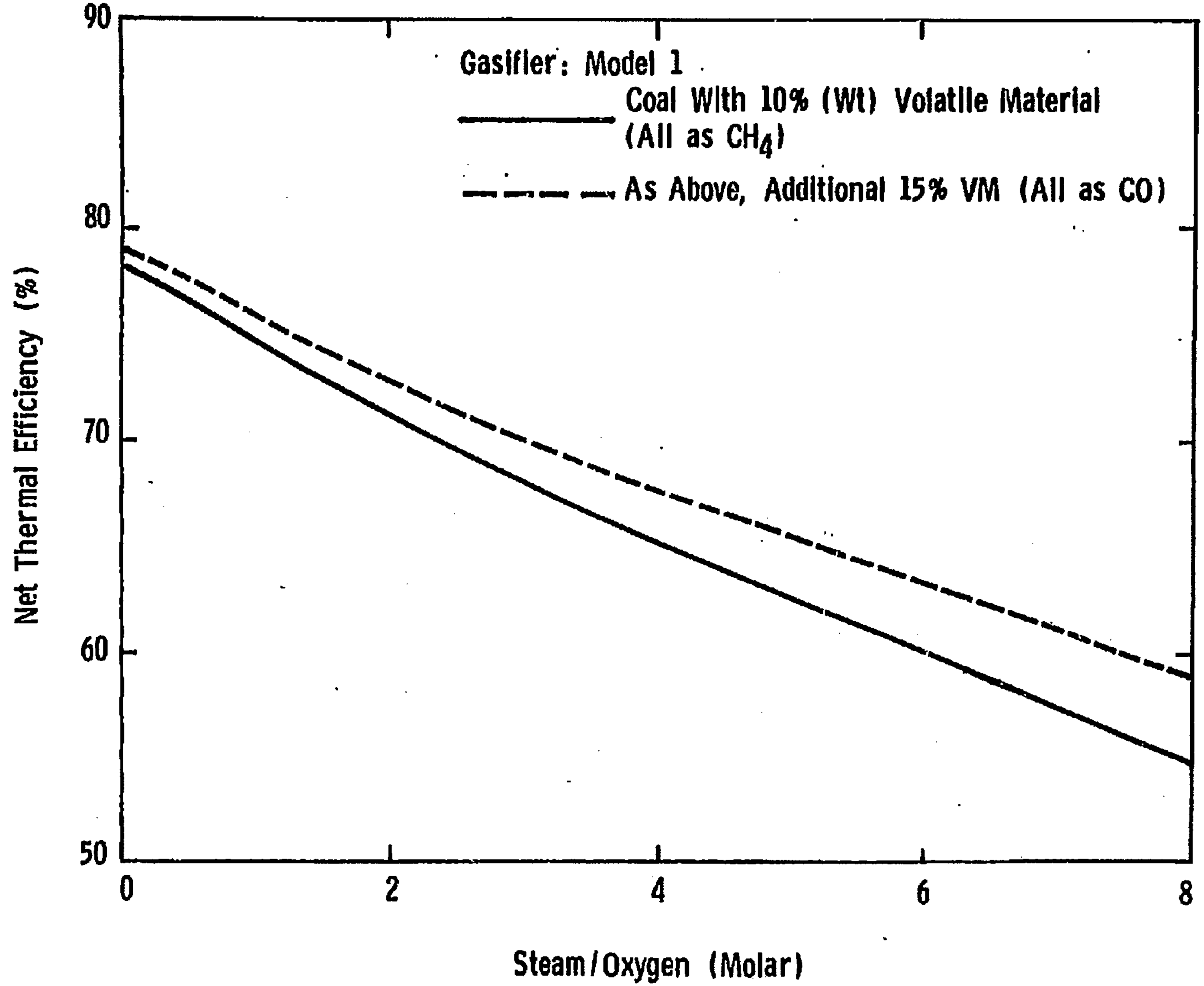


Figure 29c

OXYGEN FEED (APPROXIMATE WESTERN COAL)

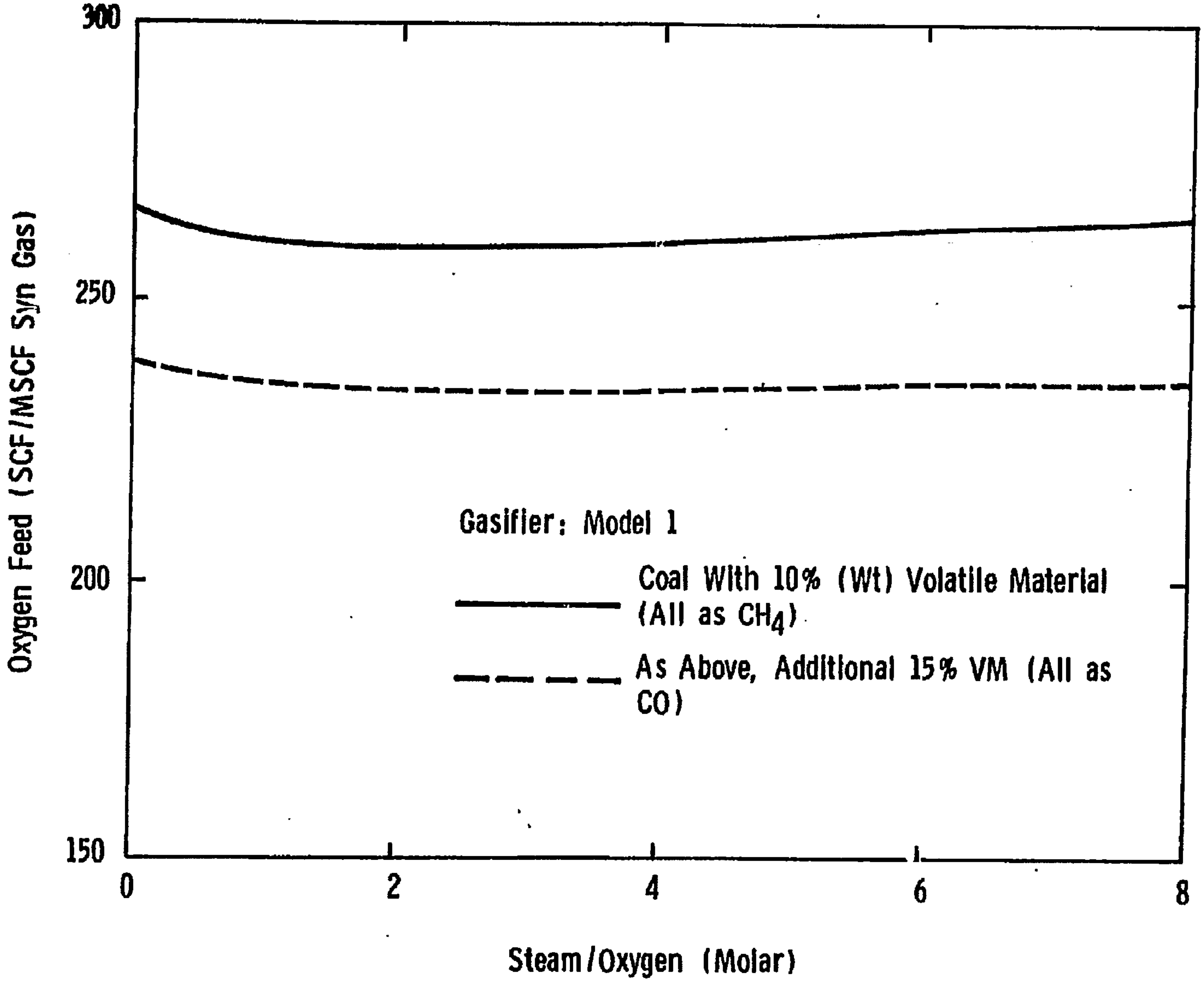


Figure 29d

GAS EXIT TEMPERATURE (APPROXIMATE WESTERN COAL)

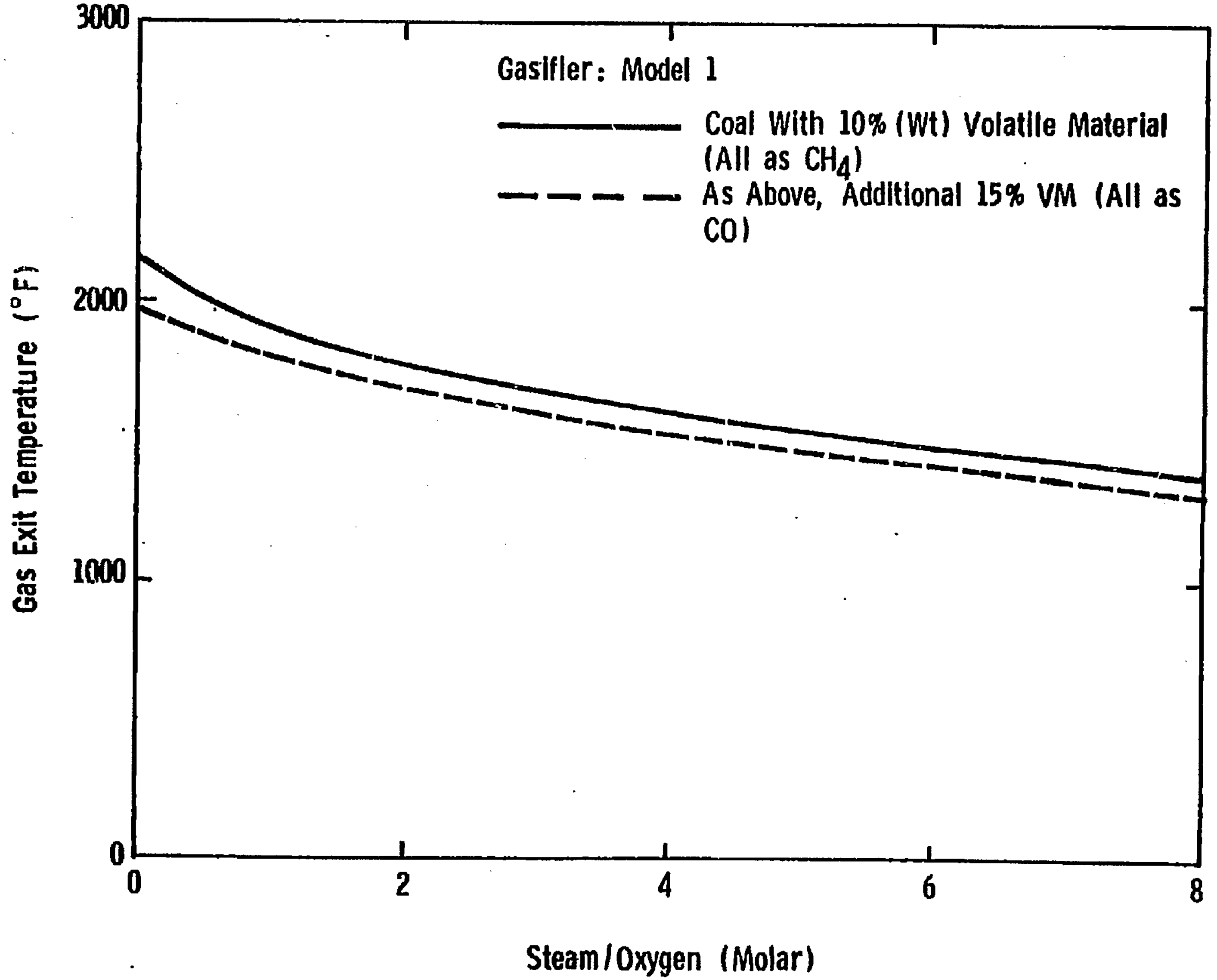


Figure 30a

COLD GAS THERMAL EFFICIENCY (APPROXIMATE WESTERN COAL)

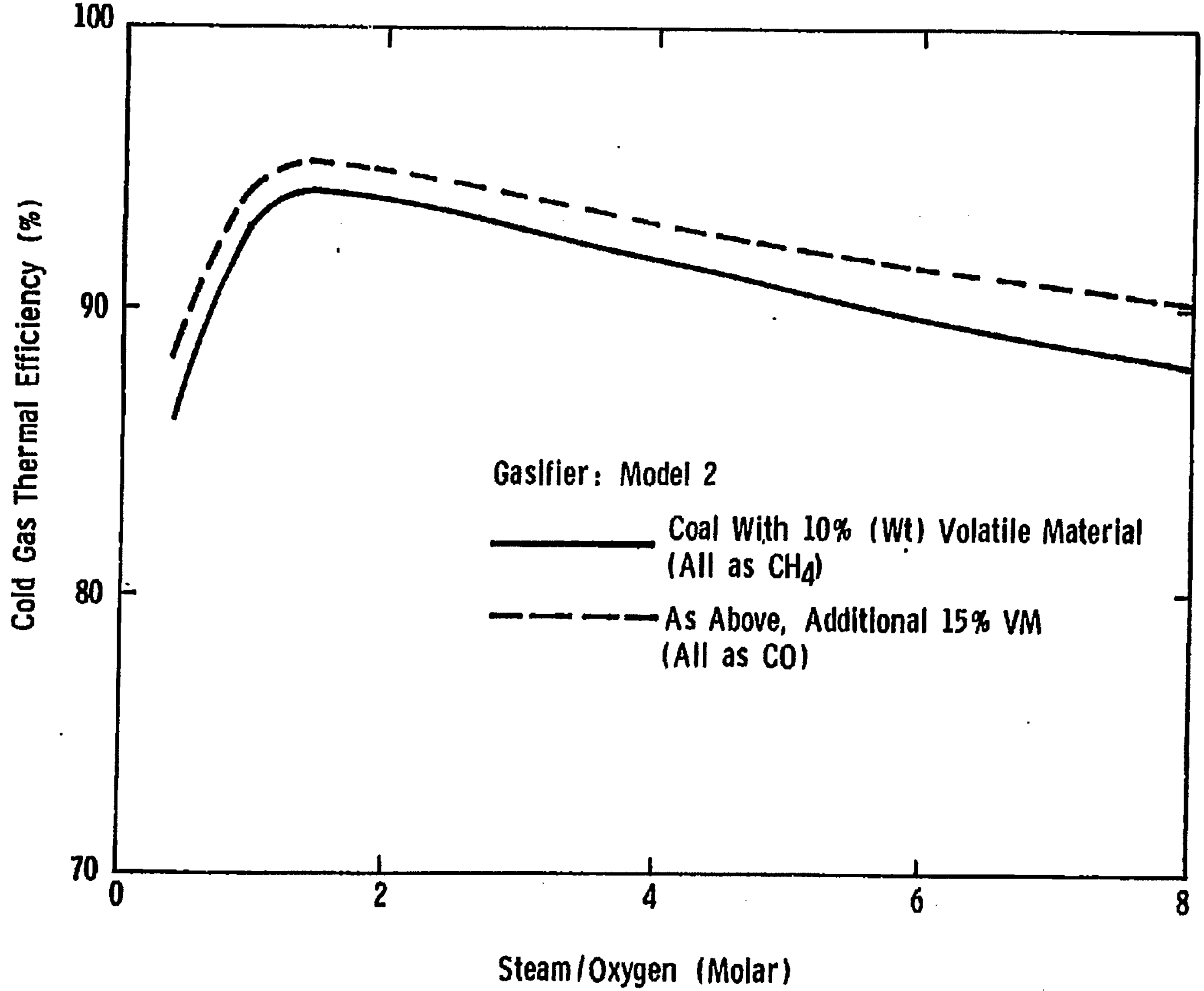


Figure 30b

NET THERMAL EFFICIENCY (APPROXIMATE WESTERN COAL)

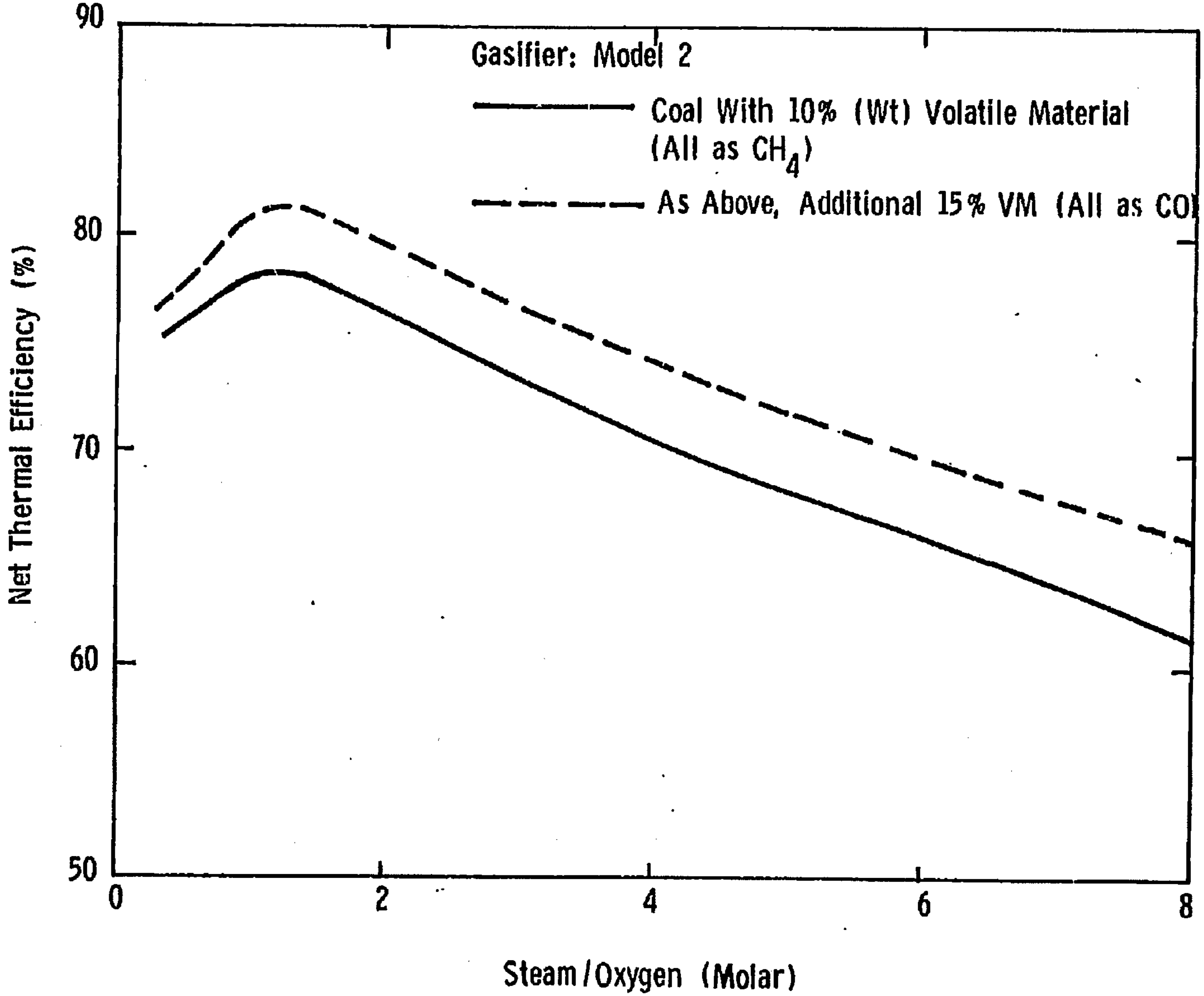


Figure 30c

OXYGEN FEED (APPROXIMATE WESTERN COAL)

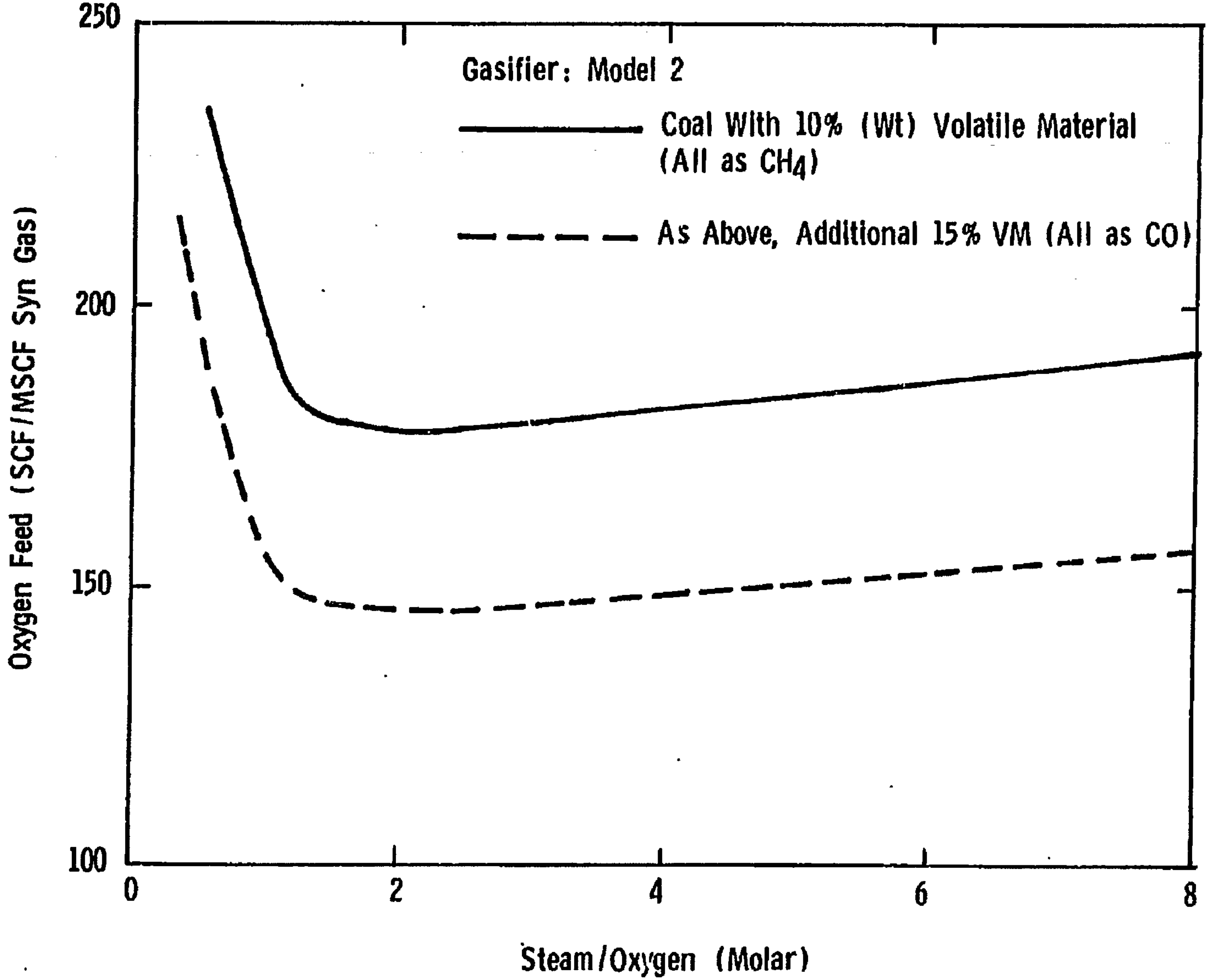
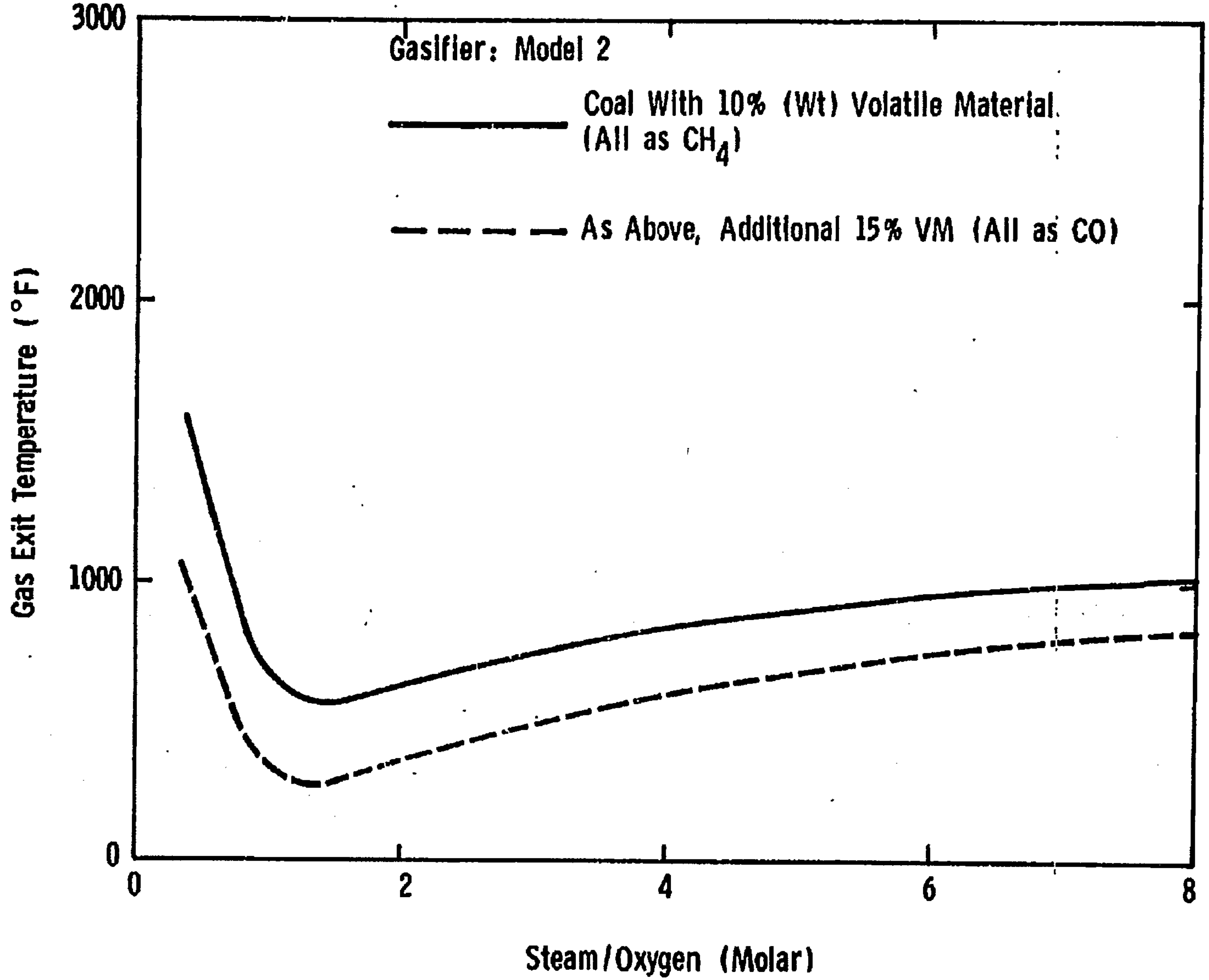


Figure 30d

GAS EXIT TEMPERATURE (APPROXIMATE WESTERN COAL)



XII. Pressure Effects

One important variable in our consideration is the gasifier pressure. A significant penalty is incurred when the pressure is close to atmospheric. Figure 31 gives the effect of gasification pressure on the net thermal efficiency for eastern coal with kinetic constraints and a steam to oxygen ratio of 1.5. The conditions are those used for the simplified gasifier of Figures 23a-h. Three curves are given showing the effects of delivering the products at 50, 200, and 400 psia. The substantial loss in net thermal efficiency for low pressure gasifiers is clearly shown. On the other hand, compression of the gas from one atmosphere to 1000 psia will require about 20% of the energy contained in the gas. The energy of compression required for a specific syngas example is given in Figure 32a and the investment required is given in Figure 32b. Both high pressures and low pressures lead to increased equipment cost. The law requires that high temperature equipment be tested at 200 psia. Thus, at low pressures, the throughput is low and large volume equipment is needed with practically no savings in steel cost per unit volume. At high pressures the volume decreases but this decrease is accompanied by increased metal cost per unit volume. For each case there is an optimum pressure. No optimization has been done for our study but the experience with similar cases leads us to the conclusion that the cost-pressure relation has a minimum between 200-500 psia and is fairly flat in this range. This is obviously only true if kinetic consideration does not require a high pressure.

For conversion of syngas to methanol, a pressure of 700 psia or higher is desirable, but for other processes under consideration, 200-300 psia is sufficient. However, compression from 300 to 700 psia gives an energy penalty of only 4%.

The gasification reaction (reactions (3), (4), and (8)) has a Langmuir-type pressure dependence. At higher pressures, longer residence times of the gas are needed to obtain the same steam conversion. Furthermore, methane formation is increased. For fuel gas production methane formation is no penalty, but for syngas production it is, especially if there are no benefits that compensate for it. Available data for SNG gasifiers (Synthane and CO_2 acceptor process) show no real advantage to operate a SNG gasifier at pressures above 300 psia.

High pressures also involve other penalties. Oxygen is expensive to compress and steam at high pressures has a higher value which has been accounted for in our calculations. It is much easier to find steam of 300-400 psia in the plant. One can, for example, get it from the methanol or Fischer-Tropsch reactor. If high pressure steam is available, one can superheat it and expand it to 300 psia. For pressures up to 600 psia presently available lock hoppers can be used to feed the coal. For pressures higher than 600 psia, the only presently proven feed system is to feed a coal slurry with all the penalties that this involves. From these considerations, the best pressure range for production of syngas and fuel gas is from 200 to 500 psia.

Figure 31

NET THERMAL EFFICIENCY WITH KINETIC CONSTRAINT
(APPROXIMATE EASTERN COAL)

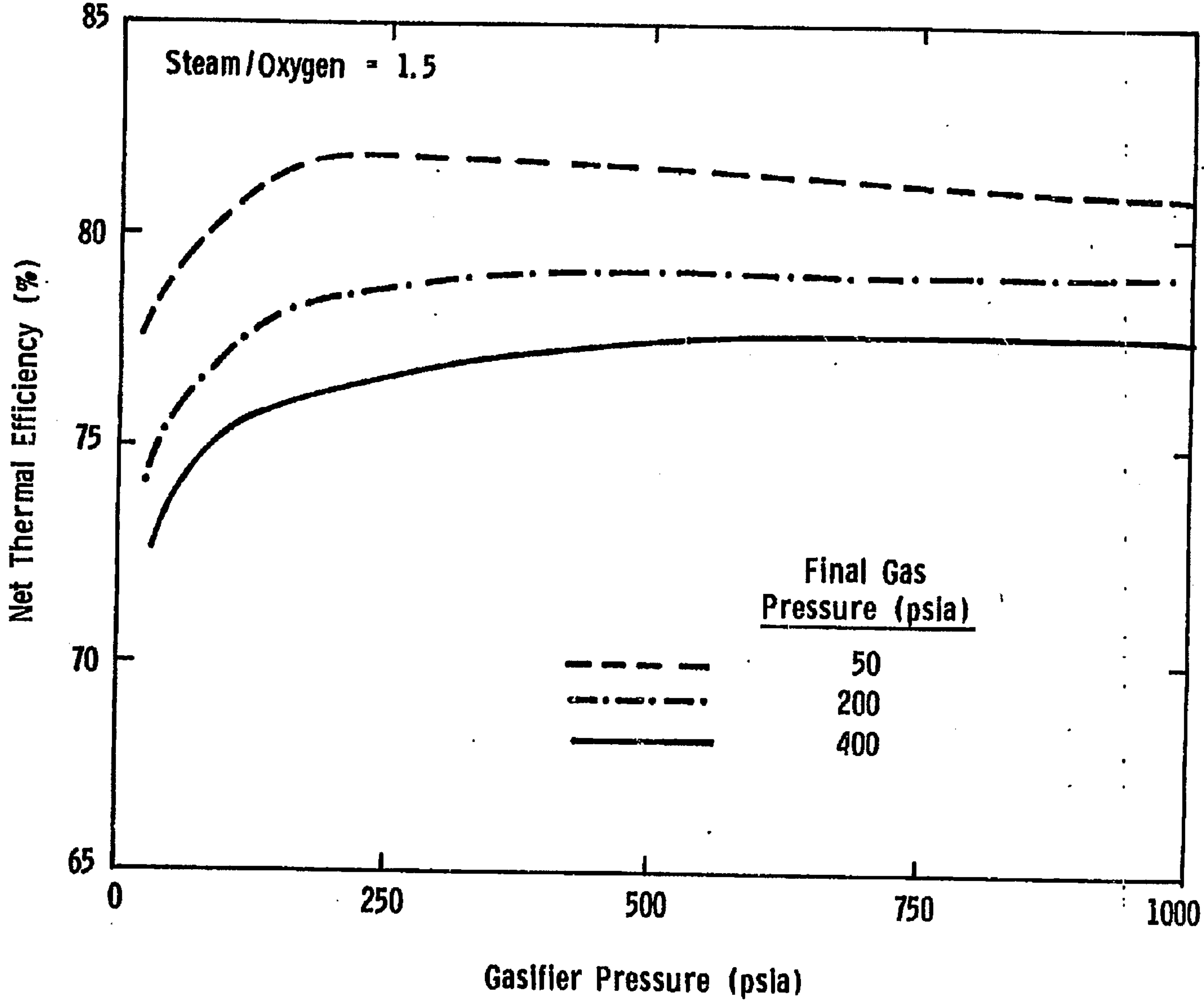


Figure 32a

FUEL REQUIREMENT FOR SYN GAS ($H_2/CO=2$) COMPRESSION

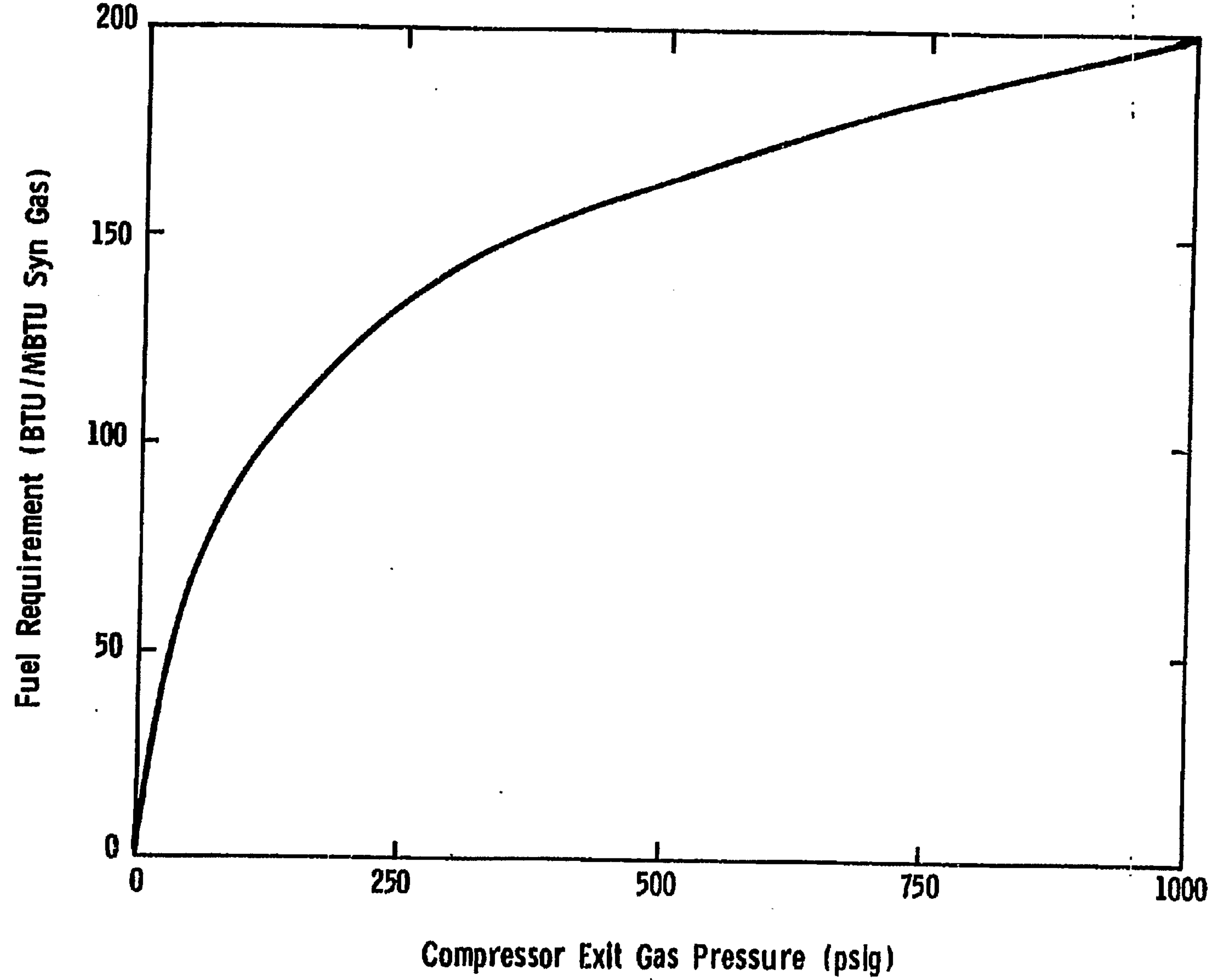
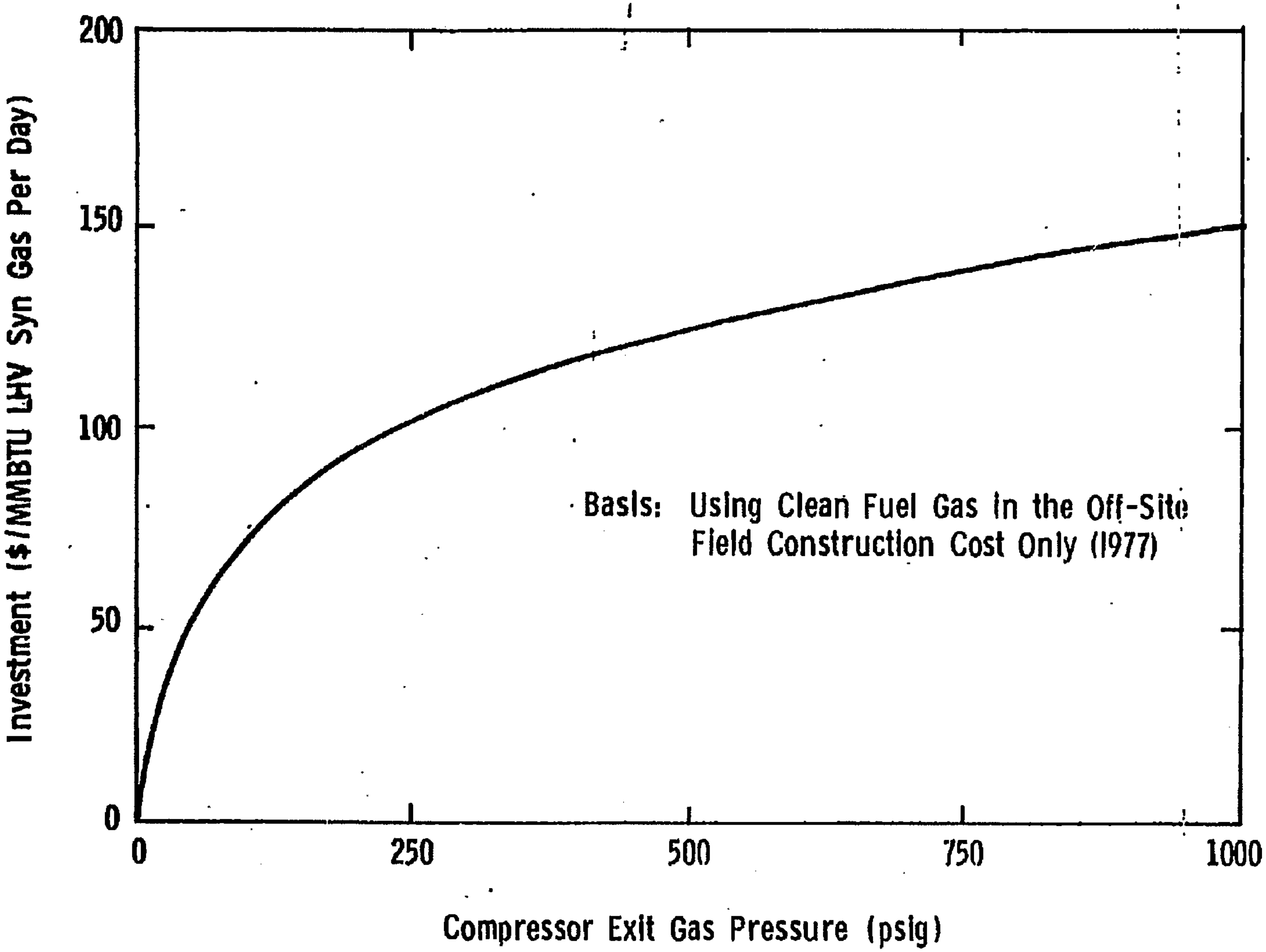


Figure 32b

INVESTMENT REQUIRED FOR SYN GAS COMPRESSION
($H_2/CO = 2$)



XIII. Temperature Constraints

There are temperature constraints for gasifiers. If the temperature is much below 1800°F, it is difficult to obtain complete coal conversion. On the other hand, if temperatures much above 1800°F but below the ash melting temperature are used, the problem of ash agglomeration comes into play.

Some gasifiers operate with low coal conversion. Hygas and Synthane gasifiers are examples of such gasifiers. Incomplete coal conversion increases the relative importance and contribution of volatilization to the thermal efficiency of the gasifier since the residual char is simply subtracted from the coal feed in the procedure used in this study. The char produced from such gasification is difficult to sell.

The high steam to oxygen parts of Figures 21, 22, and 23 are not realistic for non-catalytic gasifiers unless either the gasifier is operated at low coal conversion as discussed above or some countercurrent exchange is provided between the hot products and the cold feed. The latter operation is utilized in the dry ash Lurgi gasifier to be discussed in detail later.

Real gasifiers have other temperature constraints depending on design. For example, the top temperature in a countercurrent gasifier cannot be too low as otherwise the tar will condense on the coal which might cause problems especially in a moving bed. Second, the bottom temperature must be high enough to allow complete conversion, but must also fulfill other constraints that depend on gasifier design. For example, a BGC-Lurgi slagger requires a minimum temperature at the bottom to melt the ash whereas the dry ash Lurgi gasifier has a maximum temperature constraint. A dry ash Lurgi gasifier must, therefore, operate with high steam to oxygen ratios (5-8), whereas a BGC-Lurgi slagger operates at low ratios (1.0-1.5).

XIV. Real Gasifiers

In the preceding discussion, idealized and simplified gasifiers have been used to explore the effects of some basic variables on the efficiencies and other performance characteristics of gasifiers. Let us now see if these observations on the idealized and simplified gasifiers can be seen in real gasifiers. Reliable data in sufficient detail are hard to obtain for pilot plants for advanced gasifiers and for many of the commercial gasifiers. Table 7 gives data on eleven gasifiers that the investigators on this project felt satisfied our criteria of reliability, amount of detail and relevance to the problem of the production of syngas and fuel gas. There is one computer estimate in this set of eleven, but all others are derived from operating units.

In the preceding discussion, it was observed that the amount of unconverted steam in the exit gas of the gasifier has an important effect on the thermal efficiency of a gasifier. The gasifiers in Table 7 have been arranged in order of increasing amounts of exit steam as given in column 4. The cold gas and net thermal efficiencies are given in the last three columns. There are two values for net thermal efficiencies recorded: one for a gasifier that delivers 400 psia product gas for use as syngas and the other for one that delivers 50 psia gas for use as fuel gas. These gasifiers split into two classes. The first eight have low steam to oxygen ratios (less than 1.6) and the last three have high steam to oxygen ratios (7 to 9). The first group has H_2 to CO ratios less than 1.1 whereas the second group has ratios greater than 2. With three exceptions the first group has higher net thermal efficiencies than the second group as implied by the lower amount of exit steam. Columns 7 and 8 show that the two gasifiers with low thermal efficiency in the first group —Koppers-Totzek and Winkler— are characterized by low pressure. In addition, they have high exit gas temperatures. The data in Figure 31 show that gasifiers operating in the 30-50 psia range have substantial lower net thermal efficiency than gasifiers operating in the 300-600 psia range. This accounts for part of the difference as shown by the data in Table 8. Table 8 gives the breakdown of the components of the net thermal efficiency as % of LHV of net coal. The columns on work show much larger negative values for these two gasifiers than any other gasifier in Table 8. This is consistent with their low gasification pressure. They also have large negative values in the air separation column consistent with the higher oxygen demand needed for the high exit temperature. Although much of the heat can be recovered, some loss in thermal efficiency occurs. The other gasifier with lower thermal efficiency in the first group is the Texaco gasifier with high water (as slurry in coal feed) to oxygen ratio. The thermal efficiency in this case is penalized by the combustion due to the high water content in feed and the high exit gas temperature. The three gasifiers with large amounts of steam in the exit gas show, as expected, the large negative values in the steam column of Table 8. Certainly the gross feature of these

Table 7

Gasifier Information in the Order of Increasing Excess Steam

Gasifier	Coal	Ref.	lbs Exit Steam / 1000 scf syn Gas	Steam / Oxygen (Molar)	H_2/CO (Molar)	Exit Temp., (°F)	Pressure (psia)	Cold Gas Thermal (%) Efficiency	Net Thermal Efficiencies (%) 400 psia Product Gas	Net Thermal Efficiencies (%) 50 psia Product Gas
BGC-Lurgi Slagger	Eastern	(7)	2.1	1.15	0.52	820	300	90.1	77.6	82.3
BGC-Lurgi Slagger	Eastern	(15)	4.3	1.25	0.52	945	365	88.3	74.6	79.5
BGC-Lurgi Slagger	Frances	(9)	5.2	1.33	0.50	960	365	89.4	77.8	82.7
EPRI Agglomerating (Computer Estimates)	Western	(12)	6.9	1.48	0.85	1550	340	80.9	75.2	79.9
Koppers-Totzek	Eastern	(16)	9.0	0.783	0.64	2730	20	71.3	58.0	63.4
Texaco (Slurry Feed)	Eastern	(7)	11.9[a]	0.95[a]	0.60	2360	600	75.3	71.8	75.9
Winkler	German Brown Coal	(17)	21.4	1.52	1.09	1700	30	74.7	56.8	63.0
Texaco (Slurry Feed)	Eastern	(18)	25.2	1.59[a]	0.87	2300	815	68.2	67.8	71.1
Dry Ash Lurgi	Western	(2)	38.9	7.47	2.07	900	430	80.0	70.6	74.3
Dry Ash Lurgi	Eastern	(11,12)	56.0	8.45	2.79	1078	315	76.0	58.8	63.1
Synthane Pilot Plant	Western	(4)	72.5	8.00	2.99	1390	615	78.3	60.9	64.1

(a) As water in slurry.

Table 8

Net Thermal Efficiency Breakdown as % LHV Net Coal

Gasifier [a]	Coal	$H_2 + CO$	CH_4	C_2^+	Naph.	Heat	Steam	Air Sep.	Work 400 psia Product Gas	Work 50 psia Product Gas	Net Thermal Efficiencies (%) 400 psia Product Gas	Net Thermal Efficiencies (%) 50 psia Product Gas
BGC-Lurgi Slagger	Eastern	69.6	18.2	2.3	0	0	-1.6	-7.9	-3.0	1.8	77.6	82.4
BGC-Lurgi Slagger	Eastern	68.3	16.5	3.5	0.5	0	-1.6	-8.9	-3.7	1.2	74.6	79.5
BGC-Lurgi Slagger	Frances	71.1	16.3	2.0	3.1	0	-2.8	-8.4	-3.5	1.5	77.8	82.8
EPRI Agglomerating (Computer Estimates)	Western	63.9	17.0	0	0	13.7	-5.0	-10.3	-4.1	0.6	75.2	79.9
Koppers-Totzek	Eastern	71.3	0	0	0	12.9	-0.3	-13.7	-12.2	-6.8	58.0	63.5
Texaco (Slurry Feed)	Eastern	75.1	0.3	0	0	14.4	0	-14.1	-3.8	0.3	71.9	76.0
Winkler	German Brown Coal	69.5	5.2	0	0	8.7	1.0	-14.7	-12.9	-6.7	56.8	63.0
Texaco (Slurry Feed)	Eastern	67.2	1.0	0	0	18.7	0	-15.4	- 3.7	-0.4	67.0	71.1
Dry Ash Lurgi	Western	47.8	29.4	2.8	7.6	4.3	-13.4	-6.0	-1.9	1.8	70.4	74.1
Dry Ash Lurgi	Eastern	49.3	23.6	3.1	5.7	5.3	-16.4	-8.3	-3.5	0.8	58.8	63.2
Synthane Pilot Plant	Western	45.5	32.1	0.7	0	22.1	-26.3	-9.9	-3.3	-0.1	60.9	64.1

(a) See Table 7 for references.

gasifiers is consistent with the conclusion obtained from the study of the idealized and simplified gasifiers in the preceding sections.

Table 9 gives additional data on the eleven gasifiers of Tables 7 and 8 and includes other non-catalytic gasifiers for which less complete data were available. Representatives of two general classes of non-catalytic gasifiers are missing from Table 9. They are two-stage entrained bed gasifiers and gasifiers with separate air combustor. The reason for eliminating these gasifiers will now be discussed.

A. <u>Two-Stage Entrained Bed Gasifier (Modified Bi-Gas)</u>

This gasifier, which is discussed in detail in References 7, 11, and 12, looks reasonably attractive. However, at present it is a purely conceptual gasifier and not even bench scale data of the kind needed could be obtained to allow us to make a reasonable evaluation. In our opinion the extrapolations of available data as used in other estimates are too great for our purpose. The present Bi-Gas pilot plant is intended to work at high pressure to maximize methane formation. If it operates as intended, one could probably modify it to study fuel gas productions. Until then, any conclusions are premature.

B. <u>Indirect Combustion Gasifiers</u>

The heat required for gasification can be supplied indirectly by combusting char or coal in a separate vessel and circulating the heat to the gasifier. This is required for syngas and medium BTU fuel gas to prevent nitrogen dilution of the product. Four gasifiers of this kind were examined briefly. The old ICI gasifier was atmospheric and, therefore, does not meet our requirements. Exxon developed such a gasifier for SNG and abandoned it, as it was non-competitive, but no data from Exxon were available to us. Battelle built a pilot plant for such a gasifier but it has not operated and no data are available. The most advanced is the CO_2 acceptor process that is intended for SNG. Its present status is unclear[8]. It could probably be modified for fuel gas production.

The major advantage of the indirect combustion gasifier over the direct combustion oxygen blown gasifier is the savingsof the oxygen plant. For a syngas plant there is the added advantage that the total amount of CO_2 to be removed is smaller. As we showed in the section on stoichiometry, the total amount of CO_2 to be removed is proportional to the hydrogen content of the product, and the amount of oxygen fed to the gasifier. However, the advantage cited on page 15 for operation at point A does not apply here. The CO_2 content at the outlet of the gasifier is higher and H_2S removal is more difficult than when operating close to point A. On the other hand, the indirect combustion gasifier has to have SO_2 removed from the stack gas.

Table 9

Gasifier Information

Gasifiers (Ref.)	Pressure (psia)	Oxygen Requirement scf/mscf Syngas	Oxygen Requirement mscf/MMBTU LHV Gas	Steam Requirement lb/mscf Syngas	Steam Requirement lb/MMBTU LHV Gas	Steam/ Oxygen (Molar)	Exit Temp. (°F)	Coal Conv. (%)	H_2/CO (Molar)	4 $CH_4/(H_2+CO)$ (Molar)	Cold Gas Thermal Efficiency (%) Based on Net Coal LHV	Cold Gas Thermal Efficiency (%) Based on Coal LHV	Net Thermal Efficiency (%) 400 psia Product Gas
Eastern Coal													
Texaco (Slurry Feed) (7)	600	332	1.100	15.0[a]	49.7[a]	0.95[a]	2360	100	0.68	0.01	75.3	75.3	71.8
Dry Ash Lurgi (11,12)	315	188	0.642	75.5	258	8.45	1078	99	2.79	0.60	76.0	73.4	58.8
BGC -Lurgi Slagger (7)[c]	300	157	0.514	8.57	28.1	1.15	820	100	0.52	0.35	90.1	90.1	77.6
Koppers-Totzek (16)	20	341	1.125	12.7	41.9	0.783	2730	96	0.64	0	71.3	71.3	58.0
IGT Hygas Pilot Plant (13,14)	1035	128	0.432	87.7	296	14.4	640	61	2.87	2.06	(b)	(b)	(b)
BGC-Lurgi Slagger (15)	365	179	0.593	10.6	35.2	1.25	945	100	0.52	0.34	88.3	79.9	74.6
Texaco (Slurry Feed) (18)	815	397	1.327	29.9[a]	100[a]	1.59[a]	2300	97	0.87	0.02	68.2	67.0	67.8
Western Coal													
Dry Ash Lurgi (6)	430	130	0.441	46.2	156	7.47	900	99	2.07	0.78	80.0	76.3	70.6
Synthane PDU (5)	300	156	0.526	40.4	136	5.45	1540	76	1.55	0.56	(b)	(b)	(b)
	600	146	0.492	46.8	158	6.75	1480	66	2.05	0.98	(b)	(b)	(b)
Synthane Pilot Plant (4)	615	218	0.744	82.8	283	8.00	1390	81	2.99	0.89	78.3	62.5	60.9
IGT Hygas Pilot Plant (13,14)	1000	103	0.345	62.4	209	12.8	(b)	72	4.14	2.76	(b)	(b)	(b)
EPRI, Agglomerating Ash (Computer Estimates) (12)	340	226	0.752	15.8	52.7	1.48	1550	99	0.85	0.35	80.9	80.9	75.2
Others													
Winkler-German Brown Coal (17)	30	343	1.155	24.7	83.2	1.52	1700	86	1.09	0.10	74.7	66.1	56.8
Grand Forks Slagger-N. Dakota Lignite (10)	415	189	0.570	9.09	27.3	1.01	(b)	(b)	0.50	0.35	(b)	86.2[d]	(b)
BGC-Lurgi Slagger-Frances Coal (9)	365	168	0.555	10.6	35.0	1.33	960	100	0.50	0.32	89.4	82.5	77.8

(a) As water needed in slurry.
(b) Not available due to poor or incomplete material and energy balances.
(c) Tar, oil, naphtha and phenol are recycled.
(d) Reported number from the Reference.

Table 10 gives a comparison of the oxygen requirements for direct and indirect combustion gasifier and the air requirement for the indirect combustion gasifier. In a BGC-Lurgi slagger, direct combustion requires 160-190 scf of oxygen per 1000 scf of syngas produced. If the same heat had been supplied by air combustion, somewhat more than 760 scf of air would have been needed. However, more heat is needed when air is used in a separate vessel since about 45% of the heat generated in the combustor is taken out by the hot outlet gas. Although part of this heat can be recovered, it still requires more air to be combusted. Table 10 shows that from 1300 to 1600 scf are actually required. Note also that the cold gas thermal efficiency is lower. The high cold gas thermal efficiency of the direct combustion gasifier is very desirable for syngas conversion processes since steam is produced in these downstream processes and can be used in the coal conversion step to prepare feed. The extra steam produced by the indirect combustor is also of questionable advantage in fuel gas production even when it can be sold since its production is tied directly to the production of the fuel gas and reduces the process flexibility.

If the hot gases from the combuster could be expanded through a turbine, a good thermal efficiency would be obtained. However, at the present state of technology it must be cooled and cleaned first. Such coal-fired turbines are still far from being realized.

The improvement in thermal efficiency by not having to separate the oxygen from the air is approximately balanced by the inefficiencies of the air compression and the expansion of flue gas when indirect air combustion gasifiers are compared to low oxygen consuming gasifiers such as the BGC-Lurgi slagger. This need not hold for high oxygen consuming gasifiers such as the Texaco gasifier.

The CO_2 acceptor gasifier eliminates the scrubber problem but it introduces another problem in that a regenerator for the dolomite is required. The major problem with this gasifier is that it is difficult to have a heat balance.

The differential investments for the direct and indirect combustion gasifiers are given in Table 11. The specific investment for preparing 160-190 scf of oxygen per day is \$130-\$200 including the boiler. On the other hand, the investment to scrub the SO_2 from 1300-1600 scf of air per day is \$50-\$75, the incremental cost of the gasifier is \$50-\$100 and the investment for the compressor plus turbine power recovery is \$80-\$150. This gives a differential investment in favor of the direct combustion route of -\$20 to +\$195. In addition, only the direct combustion route has the advantage of saving the differential investment for CO_2 removal required in case of syngas production. The incremental CO_2 removal investment is \$50

Table 10

Direct Versus Indirect Combustion Gasification

	Direct[a]	Indirect[b]
SCF Oxygen/MSCF Syn Gas	160-190	250-300
SCF Air/MSCF Syn Gas	-	1300-1600
Maximum Cold Gas Thermal Efficiency, %	91	~80-82

[a] Based on BGC-Lurgi Slagger Data.

[b] Based on the investigator's own estimation.

Table 11

Investment Comparison of Direct Combustion Gasifiers to Indirect Combustion Gasifiers

Dollars Daily mscf Syngas (1977)

	Direct	Indirect	Advantage for Direct Δ
Oxygen Plant	130-200	Not Needed	-200 to -130
Air Compressor + Turbine Power Recovery	Not Needed	80-150	+80 to +150
SO_2 Scrubber	Not Needed	50-75	+50 to +75
Incremental Gasifier Cost	-	50-100	+50 to +100
Incremental CO_2 Removal (Not Applicable to Fuel Gas)	-	50-70	+50 to +70
Net for Fuel Gas			-20 to +195
Net for Syngas			+30 to +265

to \$70. No removal is required for fuel gas.

The total differential in our estimate varies from a potential advantage of \$195 per daily million BTU for the direct combustion oxygen blown gasifier to a disadvantage of -\$20. Even if the latter is true, it is too small a difference to compensate for lower cold gas thermal efficiency of the indirect combustion gasifier.

Such a gasifier might compete with a Texaco gasifier. However, no indirect combustion gasifiers are at a stage of development to allow a judgment to be made. It is felt that it does not offer as good a development potential as the direct combustion oxygen blown gasifier. Consequently, it was dropped from our study. However, it should be pointed out that the above statements apply only to non-catalytic gasifiers. In catalytic gasifiers the problem is more complex since the presence of oxygen is detrimental to methane formation reactions. Consideration of such gasifiers is outside the scope of this study.

C. Differential Evaluation of Gasifiers

Three gasifiers were chosen for the final evaluation and comparison with commercial gasifiers. These are:

a) The BGC-Lurgi moving bed slagging gasifier

b) An agglomerating fluid bed gasifier

c) The Texaco single stage entrained bed gasifier, with a water-coal slurry feed.

None of these gasifiers is completely developed. For the BGC-Lurgi slagger, a semi-commercial unit has been successfully operated. The Texaco gasifier has operated in a pilot plant though the exact data are unavailable to us and our evaluation is based on information presented in Reference 18. The Texaco gasifier really does not fit our original goal as it is a high temperature gasifier with a lower thermal efficiency, but it merits discussion since it has some specific advantages.

1. The BGC-Lurgi Moving Bed Slagging Gasifier

One way to evaluate the BGC-Lurgi slagging gasifier is to compare it to its ancestor, the dry ash Lurgi gasifier, which is the only commercially viable process in operation for over 30 years. For non-caking coals as well as lignites, the dry ash Lurgi gasifier at SASOL (South African Oil, Coal and Gas Corporation) provides a good base case. The combustion reactions (Reactions (1) and (2)) are much faster than the endothermic gasification reactions (3) and (8). Therefore, a very high local temperature is obtained where the combustion takes place. In the dry ash Lurgi gasifier, heat has to be removed if the temperature is to be kept below the melting point of the ash; thus, a large excess of steam is used as a heat transfer medium to transfer this heat out of the combustion zone to the gasification zone which reduces the temperature in the combustion zone. If the coal is more reactive, reactions (3) and (8) proceed more inside the combustion zone, and the resultant cooling of this zone reduces the steam requirements (see Table 12). Therefore, the dry ash Lurgi gasifier is better suited to reactive coals.

The BGC-Lurgi slagger offers a substantial improvement as it does not require any steam as a heat transfer medium. It needs only the steam required for the gasification itself. The concept has been demonstrated with several coals in Westfield, Scotland for extended periods with a gasifier about one-third the diameter of full size gasifiers. For fuel gas it reduces the steam requirements by a factor of five and, therefore, has a considerably improved thermal efficiency over that of the dry ash Lurgi gasifier. It offers substantial savings over the dry ash Lurgi gasifier because of:

a. Lower requirements for steam production

b. Smaller waste water treatment plant

c. Lower cooling requirements

d. Lower methane content in the offgas which is especially advantageous for syngas conversion processes

e. Higher throughput per gasifier (by a factor of two to three)

f. Higher thermal efficiency. For fuel gas there is the added advantage that the lower heating value is higher because methane and H_2 content are lower.

A quantitative evaluation of these advantages will be given later. If methanol is the desired product, the advantage will be reduced since the syngas has to be shifted to higher H_2/CO ratio gas. However, only about 60% of the gas has to be shifted and the medium grade steam from the methanol and shift reactor is available for this process. The advantages of lower gasifier capital cost, smaller waste water treatment and lower methane make can still be maintained.

For those coals with which a BGC-Lurgi slagger operates well, this is at present the best gasifier for our purpose. If the mass balances in References 7 and 11 are realistic, the BGC-Lurgi slagger is a more attractive gasifier for eastern coal than either the dry ash Lurgi or the Texaco gasifier. An estimate will be given based on the assumption that operation at the conditions given in References 7, 11, and 15 can be achieved.

Table 12

Dry Ash Lurgi Gasifier Versus BGC-Lurgi Slagger

Coal	Lurgi Dry Ash		BGC-Lurgi Slagger			Grand Forks Slagger
	Western	Eastern	Frances (Scottish Noncaking Reactive)	Eastern	Eastern	North Dakota Lignite
Reference	6	11,12	9	9	15	10
SCF Oxygen/MSCF Syngas	130	188	168	157	179	189
LB Steam/MSCF Syngas	46	76	10.6	8.6	10.6	9.1
Steam/Oxygen	7.5	8.5	1.3	1.2	1.25	1.0
H_2/CO Ratio	2.1	2.8	0.50	0.52	0.52	0.5
$4\ CH_4/CO + H_2$	0.78	0.6	0.2	0.35	0.34	0.35
Cold Gas Efficiency,% (adjusted for tars)	80	76	89	90	88	(a)
Net Thermal Efficiency, %	71	59	78	78	75	(a)

(a) Not available because of poor material and energy balances.

The data given in Table 9 are from a Westfield run with a Frances coal from Scotland. This coal is non-caking among other properties. The BGC-Lurgi slagger is reported to have operated with western coals and lignites. Specific western coal data are not available because it was obtained with private sector funds, but it is reported to us that the results are similar to Frances coal. However, data of the Grand Forks moving bed slagging gasifier operated with North Dakota lignite are reported in Table 9. The results are indeed very similar to the Westfield data with Frances coal. Table 9 also includes a recent successful run with Pittsburgh No. 8 coal at Westfield [15].

All Lurgi-type gasifiers, dry ash as well as slagger, produce tars and oils. Coal fines are formed during coal grinding that cannot be fed directly to a regular Lurgi gasifier. There are several potential uses for fines. The tar and part of the fines can be fed to the boiler supplying the power to the plant. An alternative plan is to briquette the fines with tar and feed briquettes to the gasifier. This solution has no technical drawbacks but has not been proven in practice. Another use for fines is to feed premixed tar and fines with the coal into the top of the gasifier. This has been reported as having been successfully accomplished in Westfield but no data were available to us. Other sources available to us discuss feeding the tar to both types of Lurgi gasifiers, and this task has been accomplished at Westfield. Depending on location, the fines can be sold to power plants and, if the power plant is close by, they need not be compacted for shipment. The tar could be upgraded to liquid fuel by hydrocracking similar to SRC liquids, but this is outside the scope of our study and it will be assumed that the tar is fed back to the gasifier.

The naphtha and oils are potentially useful products that can be shipped and upgraded by blending into regular petroleum feedstocks to a hydrotreating unit. For a smaller fuel gas unit, however, they might present a problem if there is no convenient refinery as it is hard to justify an upgrading plant.

Another way of usefully disposing of the tar and fines would be to take the gas liquor (water) coming from the BGC-Lurgi slagger, concentrate it, disperse the fines in it, and feed it together with the tar (or all hydrocarbons) to a Texaco gasifier. The addition of the tar allows us to use a lower water to coal ratio as compared to a regular Texaco gasifier with coal water slurry feed without losing too much in thermal efficiency. Since only about 15% of the total heating value of the coal feed is involved, the lower thermal efficiency of the Texaco gasifier is not a major problem. The oxygen would come from a common plant and the gas produced would be fed to the same gas cleaning plant. This operation would increase the CO_2 content of the gas. Again, since only 15% gas is from the Texaco gasifier, this should be within the limits that could be accepted without increasing the CO_2 separation costs. In a fuel gas syngas complex, all or part of the product from the Texaco gasifier could also be separately quenched and used for hydrogen production.

The economic advantages of the BGC-Lurgi slagger over the dry ash Lurgi gasifier are shown on Table 13 by comparing the BGC-Lurgi slagger with Frances coal to the dry ash Lurgi gasifier with western coal from Reference 6. The major economic advantages of the BGC-Lurgi slagger over the dry ash Lurgi gasifier arise as follows:

a. Lower steam requirements - Table 13 shows that the largest saving ($240 per daily million BTU) is in the steam plant. This is consistent with the discussion of idealized and simplified gasifiers and with the data of Tables 7, 8, and 9.

b. Higher throughput - The throughput for the slagger reported in Reference 9 is more than twice as much as that for the dry ash Lurgi gasifier reported in Reference 6 for western coals. However, there are claims that SASOL has achieved a 75% higher throughput than that given in Reference 6. This would almost cancel the advantage of $200 per daily million BTU given in Table 13. The reduction in gasifier costs is not directly proportional to throughput since more lock hoppers are required to feed the coal for a single gasifier.

c. Gas cooling - The next largest item is $110 per daily million BTU for gas cooling and waste heat boiler. Again the lower excess steam for the BGC-Lurgi slagger contributes greatly to the reduced investment. The exit gas from the dry ash Lurgi gasifier contains a large amount of steam as shown in Table 7 from which heat must be recovered. Waste heat boilers are expensive and the value of the heat recovered is low.

d. H_2S Removal - The combined items H_2S removal and sulfur plant account for $60 per daily million BTU. The BGC-Lurgi slagger product gas contains much less CO_2 and it is, therefore, easier and cheaper to remove the H_2S, as discussed earlier in this report. This allows a cleaner fuel gas with lower investment cost. For syngas preparation, the same amount of CO_2 would have to be removed after the syngas conversion (Reference 1) but this removal would be cheaper. Not only is there no H_2S present, but also the volume of gas to be treated is much smaller. For some snygas processes, it might even be possible to forego the removal of the CO_2, as the offgas has properties of a medium BTU fuel gas.

Table 13

Investment Comparison of BGC-Lur Slagger to Dry Ash Lurgi Gasifier

Dollars per Daily Million BTU (1977)

	Dry Ash Lurgi Gasifier Western Coal	BGC-Lurgi Slagger with Frances Coal	Advantage of Slagger
Coal Handling and Prep.	180	160	+20
Gasifier, including Coal Feed and Ash Removal	470	270	+200
Gas Cooling and Waste Heat Boiler	170	60	+110
Gas Liquor Separation	52	30	+22
Phenol Removal	40	30	+10
Ammonia Removal	53	35	+18
Waste Water Treatment	72	25	+47
H_2S Removal	80	50	+30
Sulfur Plant	80	50	+30
Oxygen Plant	300	360	-60
Steam Boiler + Superheater (including BFW preparation)	560	320	+240
General Offsites	200	200	0
Total Direct Investment	2257	1590	+667

e. Gas Liquor Separation, Phenol and Ammonia Removal, and Waste Water Treatment - These four items account for $97 per daily million BTU of investment differences. The BGC-Lurgi slagger has 5 to 10 times less liquor to process because of the low amount of steam in the exit gas. This small amount of liquor of the slagger permits evaporation, feeding back to the gasifier and combusting the organic components in the concentrated liquor. This reduces the environmental problems. The same can be done for the dry ash Lurgi gasifier but at a much higher cost.

The advantage of the BGC-Lurgi slagger for eastern coals is larger than that for western coals as the investments in Table 14 show. In the dry ash Lurgi gasifier, eastern coals are harder to gasify than western coals. On the other hand, it is claimed that eastern coals are better for the BGC-Lurgi slagger than western coals.

One problem with eastern coal is its tendency to cake and agglomerate. In the recent trials in Westfield this problem was successfully overcome for a highly caking coal (Pittsburgh No. 8).[15] Operation with this coal required the addition of about 15% (wt) blast furnace slag as a flux. The results in Table 12 indicate that the addition of a flux in such quantities does not result in a significant penalty for oxygen requirements or thermal efficiency.

Solution to this problem requires large scale trails for each specific coal. Some mechanical modifications of the unit have also been proposed, and the high attractiveness of the BGC-Lurgi slagger as a gasifier for fuel gas justifies further development in this area.

The investments in Tables 13 and 14 were computed from the design and investments (field construction costs) given in Reference 6. The values in Reference 6 were reduced by a factor of 0.8 to bring them in line with DOE guidelines (Gulf Coast, 1977). No contingency and no special expenses for a labor camp are included. The steam used in the study given in Reference 6 is generated by a boiler fired with coal fines, phenol, tar and oil with a scrubber. Alternatively, fuel gas produced in the plant could be used to fire the boiler. This slightly reduces the overall thermal efficiency of the plant, but especially in the case of the BGC-Lurgi slagger might be preferrable as it simplifies the overall plant, and eliminates the problems associated with scrubbers.

The cost of the gas produced can be estimated from the direct investment costs in Tables 13 and 14. In engineering estimates, almost all operating costs and other investment-related charges are estimated by a factor multiplying the

Table 14

Investment Comparison of BGC-Lurgi Slagger to Dry Ash Lurgi Gasifier (Eastern Coal)

Dollars per Daily Million BTU (1977)

	BGC-Lurgi Slagger With Frances Coal	Dry Ash Lurgi Gasifier With Ill. No. 6 Coal	Advantage of Slagger
Coal Handling and Prep.	160	200	+40
Gasifier, Including Coal Feed and Ash Removal	270	520	+250
Gas Cooling and Waste Heat Boiler	60	250	+190
Gas Liquor Separation	30	70	+40
Phenol Removal	35	50	+15
Ammonia Removal	30	65	+35
Waste Water Treatment	25	100	+75
H_2S Removal	50	80	+30
Sulfur Plant	50	80	+30
Oxygen Plant	360	430	+70
Steam Boiler + Superheater (including BFW preparation)	320	750	+430
General Offsites	200	250	+50
Total Direct Investment	1590	2845	+1255

investment costs. After studying a large number of such calculations, the factor of 5.6 was determined for utility financing (EPRI guidelines) to give the production cost exclusive of coal costs. Thus, the production cost including the depreciation of investment, interest, construction, profits, taxes and all operating costs is given by,

production cost per unit = 5.6 x direct investment per unit

The direct investment per unit is defined as the direct investment per unit of daily production divided by the plant life of 6600 days. The BGC-Lurgi slagger using Frances coal with a direct investment of $1590 per daily million BTU gives a direct investment per million BTU of $0.24. For the BGC-Lurgi slagger, the production cost per million BTU is then $1.35. The total coal cost is the price of coal per million BTU multiplied by the reciprocal of the net thermal efficiency. Table 12 gives the net thermal efficiency of the BGC-Lurgi slagger as 79%. Thus, the cost of gas (1977 dollars) produced by the BGC-Lurgi slagger is given by,

$$\frac{\text{Cost of gas}}{\text{per million BTU}} = 1.28 \times \frac{\text{Cost of coal}}{\text{per million BTU}} + \$1.35$$

In the same way, the cost of gas (1977 dollars) for the dry ash Lurgi gasifier for western coal is given by,

$$\frac{\text{Cost of gas}}{\text{per million BTU}} = 1.39 \times \frac{\text{Cost of coal}}{\text{per million BTU}} + \$1.92$$

For the dry ash Lurgi gasifier with eastern coal,

$$\frac{\text{Cost of gas}}{\text{per million BTU}} = 1.64 \times \frac{\text{Cost of coal}}{\text{per million BTU}} + \$2.41$$

On the same basis, SNG production from western coal using a dry ash Lurgi gasifier, which requires an investment of $3,150 per daily million BTU and has an efficiency of 63%, gives,

$$\frac{\text{Cost of SNG}}{\text{per million BTU}} = 1.59 \times \frac{\text{Cost of coal}}{\text{per million BTU}} + \$2.67$$

The BGC-Lurgi slagger, for those coals for which it operates, provides a clean industrial fuel at a cost below either the dry ash Lurgi gasifier or the conversion of coal to SNG. The problem is that we do not know for which coal it operates well. More data are needed. Although some of the required data exist, they were not accessible to us. One problem might be that, if the temperature in the top of the BGC-Lurgi slagger becomes too low, tar can condense and cause serious problems. This is aggravated by high water content coals. Such coals might have to be dried which could introduce problems with the strength of the coal. On the other hand, the data for the slagger at Grand Forks, North Dakota show excellent results (see Table 12) with North Dakota lignite. Unfortunately, the runs were of relatively short duration (4 hours at steady state).

2. <u>Texaco Gasifier</u>

The Texaco single stage entrained bed gasifier(7, 18,19) is included in the detailed study for several reasons:

a. It provides a potential alternative for eastern coals if they cannot be processed by a BGC-Lurgi slagger.

b. It can generate high purity hydrogen for snygas conversion processes.

c. It might be used in a complex to process the coal fines and the tars obtained from other coal conversion processes such as the BGC-Lurgi slagger.

d. It provides another reasonable economic comparison to the BGC-Lurgi slagger.

Table 15 gives two sets of data for the Texaco gasifier reported in References 7 and 18. This table also includes data for a commercially proven single stage entrained gasifier— Koppers-Totzek. The first case is rather optimistic since it has coal slurried in a ratio of 1 part of water to 2 parts of dry coal. This may be limited to specific coals with special grinding techniques. Both geometric arguments and experience with other systems indicate that a two peaked size distribution of the coal is probably required. However, the oxygen requirement increases and the thermal efficiency decreases as the water content rises as shown in Table 15. Figures 33a and b give data from a simplified hypothetical gasifier calculation that also illustrates these effects. In this simplified gasifier, the methane reacts in the gasification zone and is reformed to CO and H_2. The differences in thermal efficiencies between a water to dry coal ratio of 0.5 and 0.85 are consistent with the differences shown in Table 15 although the absolute values of the thermal efficiencies are higher for the hypothetical case.

The cold gas thermal efficiency of the Koppers-Totzek gasifier lies midway between the two Texaco cases in spite of the fact that the coal converted in the Koppers-Totzek is only 95%. The Texaco has an advantage over the Koppers-Totzek gasifier in that it has a high net thermal efficiency and a potentially more reliable feed system for high pressure operation.

Table 15

Texaco Gasifier versus Koppers-Totzek Gasifier

	Texaco (Ill. #6)	Texaco (Ill. #6)	Koppers-Totzek (TVA Coal)
Reference	6	18	17
Water to Dry Coal Ratio	0.5	0.85	0
Feed	Slurry	Slurry	Dry
Pressure, psia	600	800	Atmospheric
lb Steam/mscf Syngas	-	-	12.7
scf Oxygen/mscf Syngas	330	400	340
H_2/CO Ratio	0.68	0.87	0.64
Cold Gas Thermal Efficiency, %	75	68	71
Net Thermal Efficiency, %	72	68	58

Figure 33a

OXYGEN FEED VERSUS COAL SLURRY COMPOSITION
(SIMULATED TEXACO GASIFIER WITH APPROXIMATE EASTERN COAL)

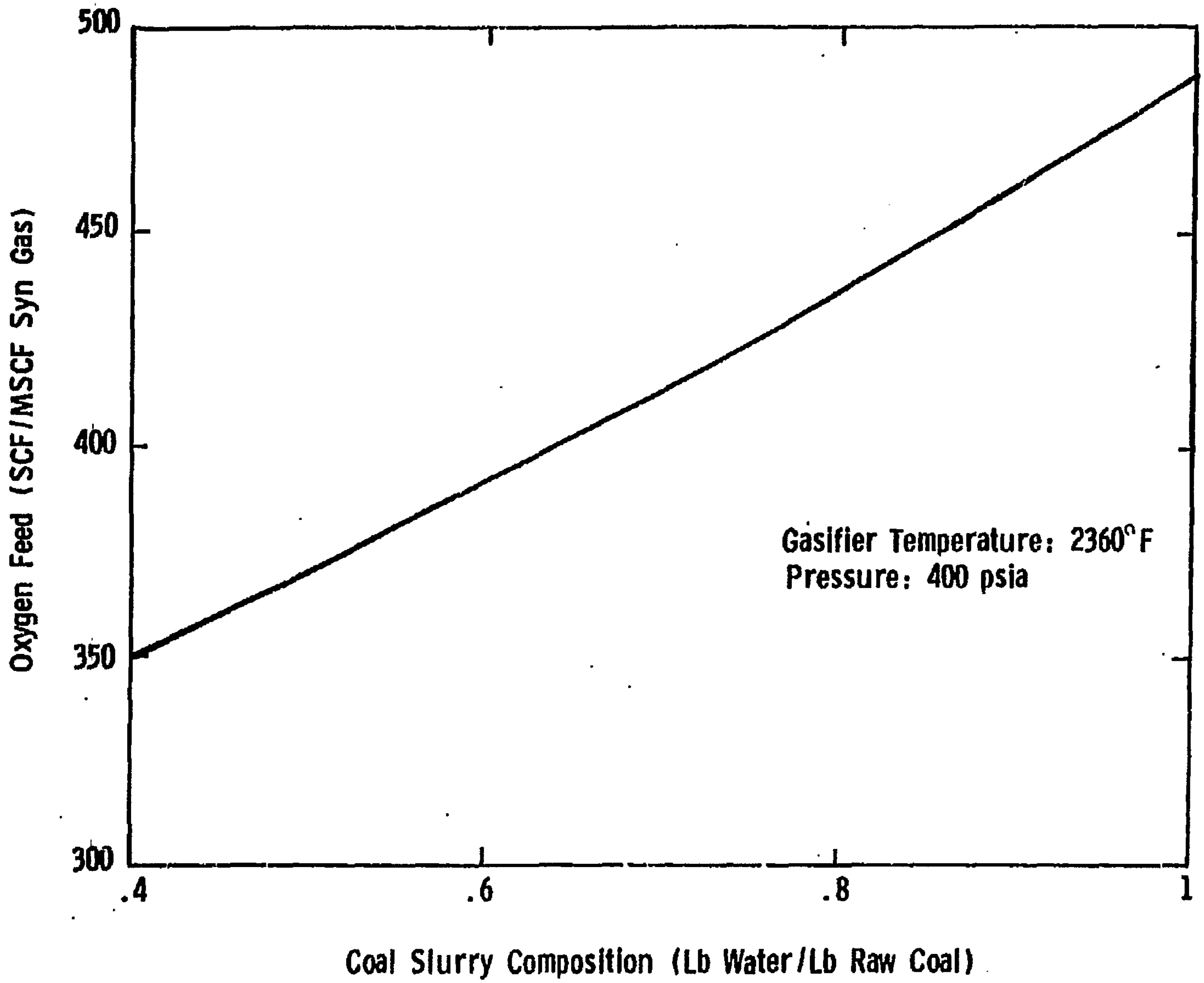


Figure 33b

COLD GAS AND NET THERMAL EFFICIENCIES VERSUS COAL SLURRY COMPOSITION
(SIMULATED TEXACO GASIFIER WITH APPROXIMATE EASTERN COAL)

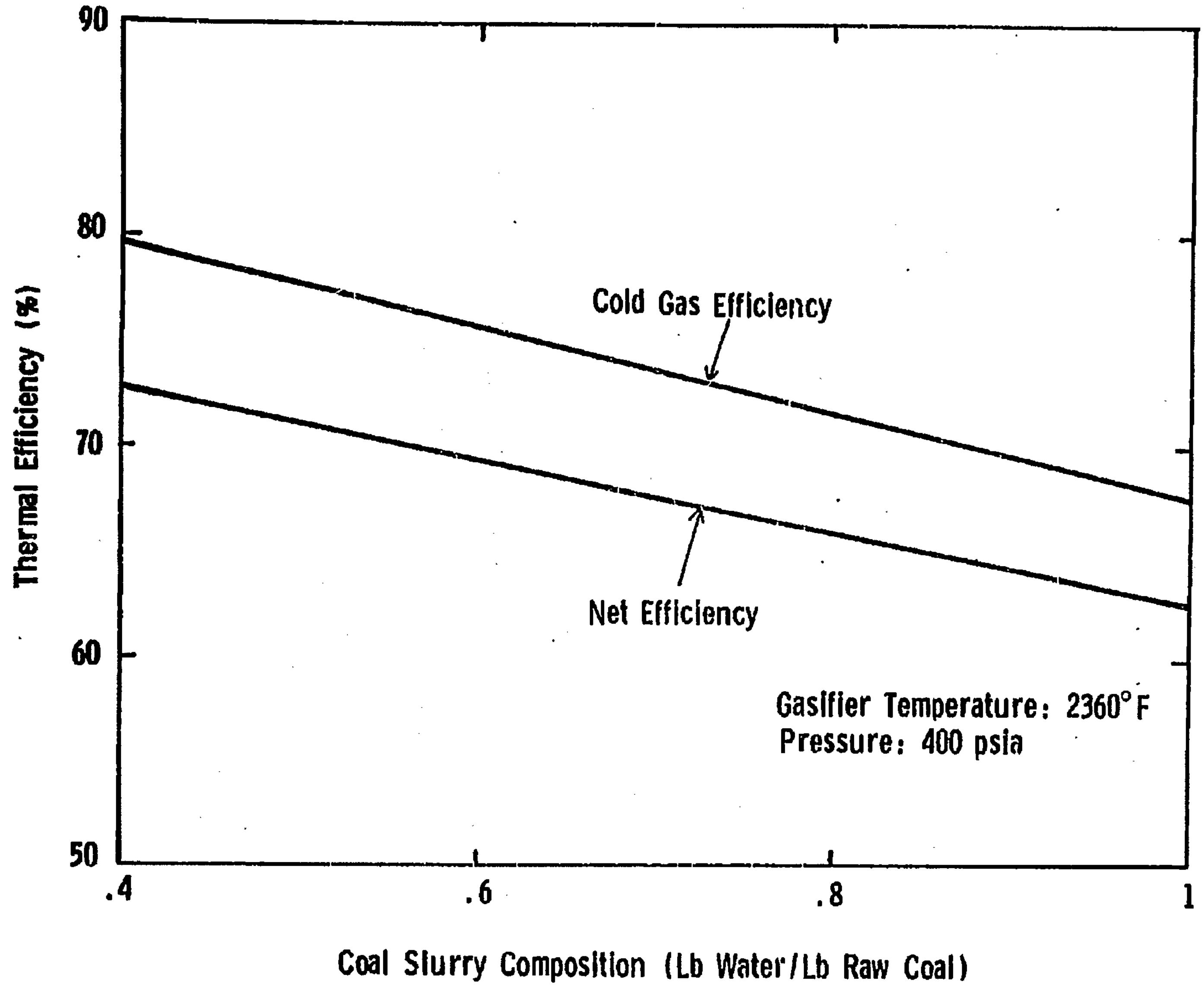


Table 16 gives a comparison of estimated investments required for the Texaco gasifier and the BGC-Lurgi slagger. Two estimates are given for the Texaco gasifier — one for an optimistic water to coal ratio (Reference 7) and the other for a higher ratio (Reference 18). The BGC-Lurgi slagger has a lower investment cost per daily million BTU (1977 dollars) than either of the Texaco cases. In the optimistic case it is $550 per daily million BTU lower and for the other $890. A comparison of the data in Tables 13 and 16 shows that the investment for the optimistic Texaco is approximately the same as that for the dry ash Lurgi gasifier for western coals. The investments and net thermal efficiencies in Tables 15 and 16 lead to a gas cost (1977 dollars) formula given by,

$$\frac{\text{Cost of gas}}{\text{per million BTU}} = 1.33 \times \frac{\text{Cost of coal}}{\text{per million BTU}} + \$1.82$$

for the optimistic case and,

$$\frac{\text{Cost of gas}}{\text{per million BTU}} = 1.49 \times \frac{\text{Cost of coal}}{\text{per million BTU}} + \$2.10$$

for the higher water content case.

Table 16 shows that the Texaco gasifier has advantages over the BGC-Lurgi slagger (a) in the low gasifier cost because it is a simpler unit, (b) in requiring no waste water treatment, and (c) in producing no phenol, oils and tars that require gas liquor separation and tar removal. A smaller steam boiler is required since the Texaco gasifier produces a large quantity of sensible heat. On the other hand, the heat recovery in the waste heat boiler is expensive. The Texaco generates more usable heat than the energy required for the preparation of the oxygen. However, this heat is not easy to recover because the hot gases from the gasifier contain molten slag, which makes the heat exchanger design difficult. The raw gas could be quenched to a temperature below the melting point of the slag, for example, 1600°F. Then the design of the heat exchanger would not present as much of a problem. However, quenching the raw gas to 1600°F shifts the quality of the steam produced so that too much low pressure steam and not enough high pressure steam is produced to supply the oxygen plant. This reduces the net thermal efficiency of the gasifier. The method of heat recovery could not be determined from the reports available to us, but Reference 19 implies that a solution without quench is being developed.

Table 16

Investment Comparison of BGC-Lurgi Slagger to Two Texaco Gasifier Schemes

Dollars per Daily Million BTU (1977)

	BGC-Lurgi Slagger	Texaco[a] Gasifier		Advantage of Slagger over Texaco	
Water to Dry Coal Ratio	0	0.5	0.85	0.5	0.85
Coal Handling and Prep.	160	200	220	+40	+60
Gasifier, including Coal Feed and Ash Removal	270	180	200	-90	-70
Gas Cooling and Waste Heat Boiler	60	500	550	+440	+490
Gas Liquor Separation	30	-	-	-30	-30
Phenol Removal	35	-	-	-35	-35
Ammonia Removal	30	50	50	+20	+20
Waste Water Treatment	25	-	-	-25	-25
H_2S Removal	50	80	90	+30	+40
Sulfur Plant	50	80	90	+30	+40
Oxygen Plant	360	730	900	+370	+540
Steam Boiler + Superheater (including BFW preparation)	320	120	160	-200	-160
General Offsites	200	200	220	0	+20
Total Investment	1590	2140	2480	+550	+890

(a) The cost estimate of two cases of the Texaco gasifier based on the balances presented in References 7 and 18. It was assumed that the heat can be recovered without quench. The design information of the Texaco gasifier given in these two references is incomplete and we had to make some approximate estimates to get a consistent estimate of the cost differential on the same basis as the BGC-Lurgi slagger. Reference 7 also contains a comparison of the BGC-Lurgi slagger with the Texaco gasifier for the purposes of a combined cycle power plant. If we look only at the part dealing with the fuel gas preparation, the cost advantage of the BGC-Lurgi slagger in Reference 7 is a similar fraction of total cost (30%) as given here.

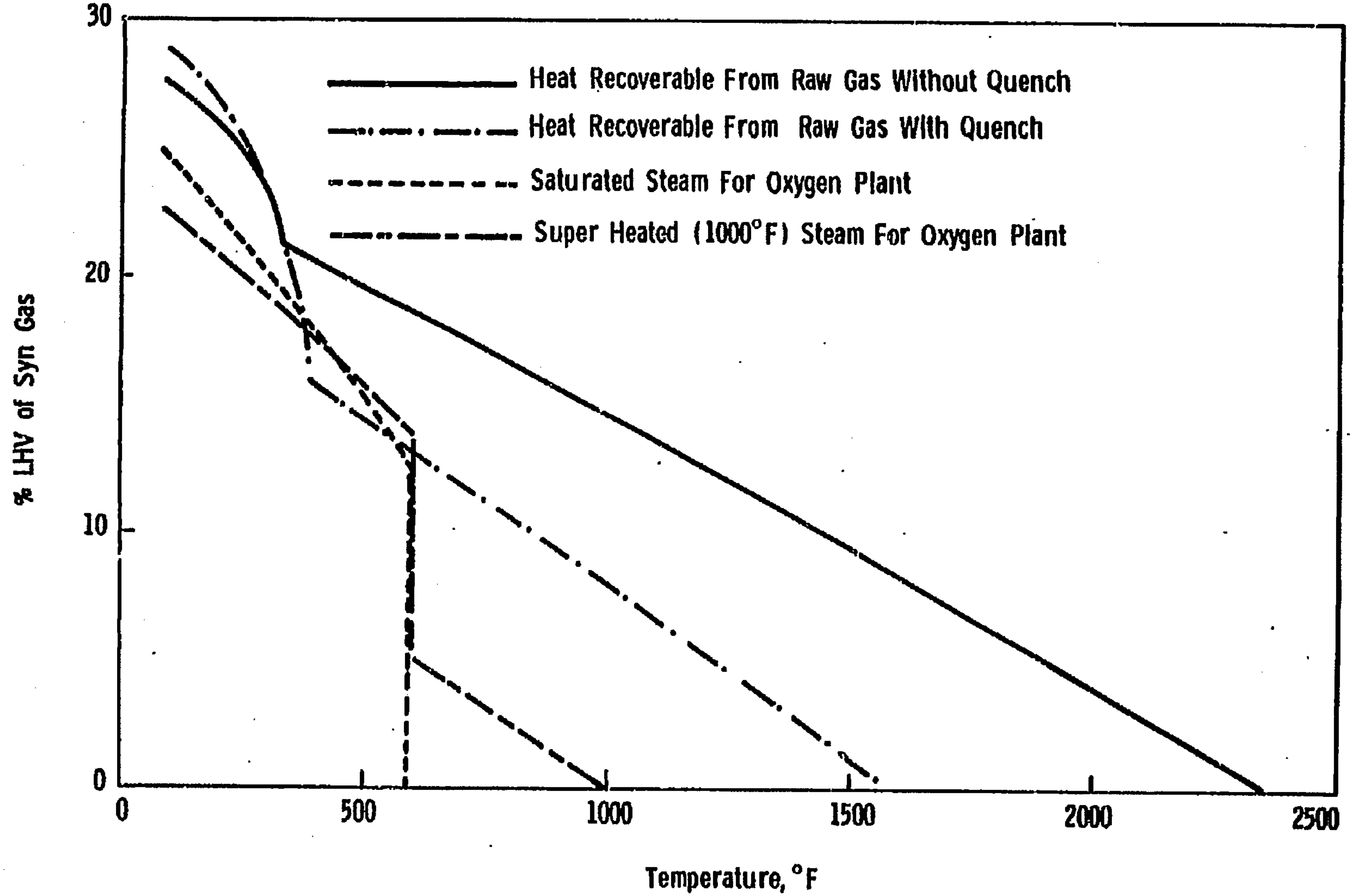
The problem arising from quenching the Texaco gasifier effluent is illustrated in Figure 34. Four curves are shown in this figure. The ordinate is the heat measured as % of the lower heating value of the clean dry gas produced and the abscissa is the temperature. Two of the curves give the heat required for the production of the amount of 1500 psia steam needed to prepare the oxygen for gasification in the Texaco gasifier. The curve marked ----------- is for saturated steam and the curve marked — - — - is for steam superheated to 1000°F. The vertical portions of these curves give the heat required to vaporize the water and left portions give the heat needed to raise the temperature of the water to the boiling point at 1500 psia. The right portion of of the superheating curve — - — - gives the heat needed to superheat the steam to 1000°F.

The other two curves give the heat that can be recovered from the hot raw gases from the gasifier as a function of temperature. Again, this heat is measured, for convenience, as % of the lower heating value of the clean gas from the gasifier. One curve, designated by the solid line, is for the raw gas with no water quench and the other given by — • — • is for the gas quenched by water to 1600°F to solidify the molten slag. The quench has the effect of transferring the heat recoverable at high temperatures to heat recoverable at a much lower temperature as shown by large amounts of heat recovered below 400°F for the curve for the quenched raw gas. In order to exchange the heat with the raw gas, a temperature differential of at least 100°F is needed to keep the heat exchanger size reasonable. The points along the unquenched raw gas curve lie to the right of the corresponding heat values of the steam production curves. Except for a very narrow region, all points are at least 100°F to the right. Thus, if the heat from the hot raw gas containing molten slag could be recovered, sufficient heat is available to prepare the oxygen for the gasification process. This is not true of the quenched case since a substantial portion of the curve —•—•—•— lies to the left of the steam production curves and, therefore, there will not be enough heat at certain temperature to prepare the amount of steam required by the oxygen plant.

The data in Table 16 clearly show that the advantages of the BGC-Lurgi slagger over the Texaco are brought about, for the most part, by the lower exit temperature of the exit gas. This lower exit temperature requires a much smaller gas cooling and waste heat recovery system in the slagger. Also, a smaller oxygen plant is needed for the BGC-Lurgi slagger. The additional oxygen that the Texaco gasifier requires goes in part to supplying the heat for the higher exit temperature and part is used to convert the slurry water to steam in the gasifier. This slurry water is the principle source of steam for the Texaco gasifier. The total investment differences for these two items provide $180 per daily million BTU advantage to the BGC-Lurgi slagger to offset the smaller steam boiler, the simpler conversion unit and the absence of tar and other undesirable products associated with the Texaco gasifier. The net advantage for the BGC-Lurgi slagger is $550 per

Figure 34

HEAT RECOVERABLE FROM A TEXACO GASIFIER EFFLUENT GAS AND THE HEAT FOR MAKING STEAM REQUIRED IN THE OXYGENT PLANT
(STEAM PRESSURE: 1500 psia)



daily million BTU over the Texaco of 0.5 water to dry coal ratio.

As we noted in Table 15 and Figures 33a and 33b, the oxygen requirements and thermal efficiency of the Texaco gasifier strongly depend on the water content of the coal slurry fed to the gasifier. For a coal of low BTU content per unit volume, the gasifier would suffer a similar effect as it increases the amount of water that has to be evaporated and treated per unit of syngas. We have no data for the dependence of thermal efficiency on coal properties but our results allow an approximate estimate of the thermal penalties that might be involved.

On the other hand, Texaco gasifier could be useful for gasification of tars, oils, and fines obtained from other gasification processes as discussed earlier. The addition of one gasifier to a complex containing many other gasifiers should cause no problems.

3. Fluid Bed Gasifiers

An important deficiency of the BGC-Lurgi slagger is that it cannot use coal fines as feed. The mining operation may yield as much as 15 to 40% of the coal in a size range too small to be used in either the BGC-Lurgi slagger or the dry ash Lurgi gasifier. In addition, it is not known how suited the BGC-Lurgi slagger is for western coals or lignites. Both of these problems could be solved by using an appropriate fluid bed gasifier. It should be possible to develop a fluid bed gasifier that meets the basic requirements established for the most thermally efficient medium BTU fuel gas and syngas gasifiers:

a) The gasifier should operate in the neighborhood of point A established in P. 23 of this report.

b) The gasifier should possess a gasification zone with a temperature sufficiently high for essentially complete conversion of the coal.

c) The gasifier should provide a second zone with a lower temperature appropriate for devolatilization.

d) The gasifier should provide heat exchange between regions of combustion, gasification and devolatilization to improve the thermal efficiency.

In their present state of development, the fluid bed gasifiers, Hygas, Snythane and Winkler, do not satisfy our criterion and are not competitive with the dry ash Lurgi gasifier. However, the experience that has been gained with these gasifiers leads us to the conclusion that a thermally efficient fluid bed gasifier could be developed. Such a gasifier would have the advantages of a BGC-Lurgi slagger, provide a method of disposing of tars and fines, and, as experience with the Synthane gasifier has shown, could gasify western coal without presenting caking problems. In addition, such a fluid bed gasifier should be easier to operate and require less highly skilled personnel. Although the fluid bed gasifier probably would not lead to less investment than the BGC-Lurgi slagger, it would probably cost approximately the same if properly designed. Such a gasifier is really a third generation gasifier since no pilot unit satisfying the required conditions is close to operation.

A gasifier, which is under development, that might satisfy the conditions is the Westinghouse gasifier[20] shown in Figure 35. However, no data were available to us on its performance.

An appropriate fluid bed gasifier might be designed having the various zones contained in a single vessel. Such a potential design is shown in Figure 36. Steam and oxygen are introduced into a narrow bottom zone at high velocity to create a well-mixed region with high velocity recirculation. This provides an agglomerating zone. Above this is the gasification zone at lower velocity in which essentially complete gasification of the coal takes place. This is followed by a dilute devolatilization zone into which coal is introduced at the top (into the freeboard). This top zone provides heat exchange between the incoming coal and the exit gas. There will be some mixing between the devolatilization zone and the gasification zone but it should not be enough to destroy too much of the product from devolatilization. Such a separation into zones exits in presently operating fluid bed gasifiers.

If the coal is introduced directly into the gasification zone, the disadvantage of high exit temperature is obtained and methane is reformed as discussed previously. The data given in Figure 28 provides a basis for assessing the effect of methane reforming. The higher oxygen required and lower thermal efficiency caused by the methane reforming are clearly shown. Additional oxygen is used to provide for the higher exit gas temperature, which further lowers the thermal efficiency, since not all the energy used in preparing the additional oxygen can be recovered as useful heat. Nevertheless, such reforming of methane has advantages when clean hydrogen is to be produced, when low methane containing syngas is needed, and when methane cannot be sold as fuel gas. In this case it is cheaper to reform the methane in the gasifier than in a separate step since no cooling and heating of the methane between units is required.

A single staged agglomerating bed operating between 1900 and 2000°F with coal fed into the bottom of the reactor gives a gas almost free of methane. Such a gasifier would have a higher thermal efficiency than the Texaco gasifier and could operate with coals not suited for the Texaco gasifier. Furthermore, it would yield a gas having a lower CO_2 content. If a cold gas thermal efficiency of 80% could be achieved, the agglomerating single stage fluid bed gasifier would be very attractive for syngas processes requiring a gas containing less than 3% methane.

Figure 35
SCHEMATIC WESTINGHOUSE GASIFIER

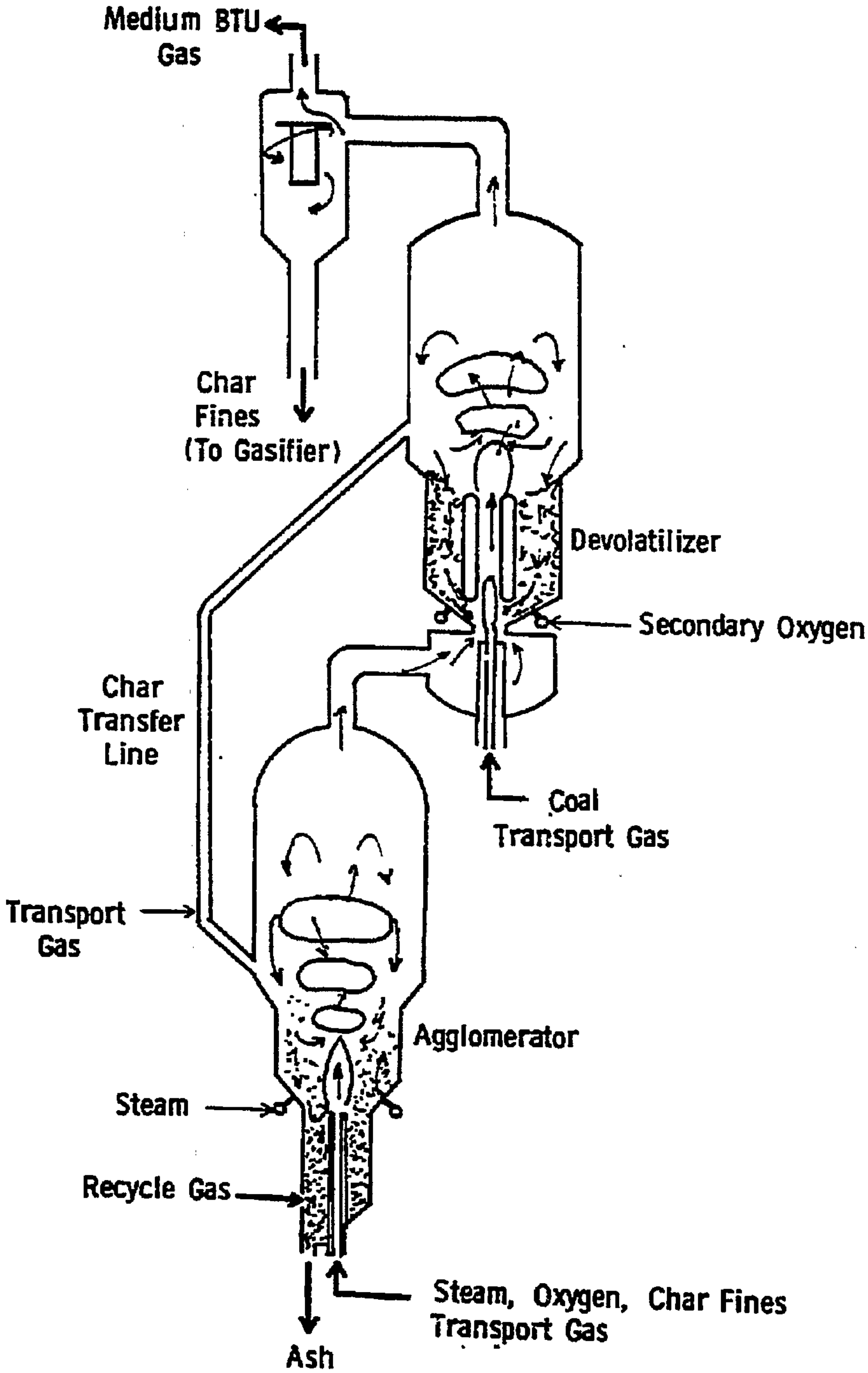
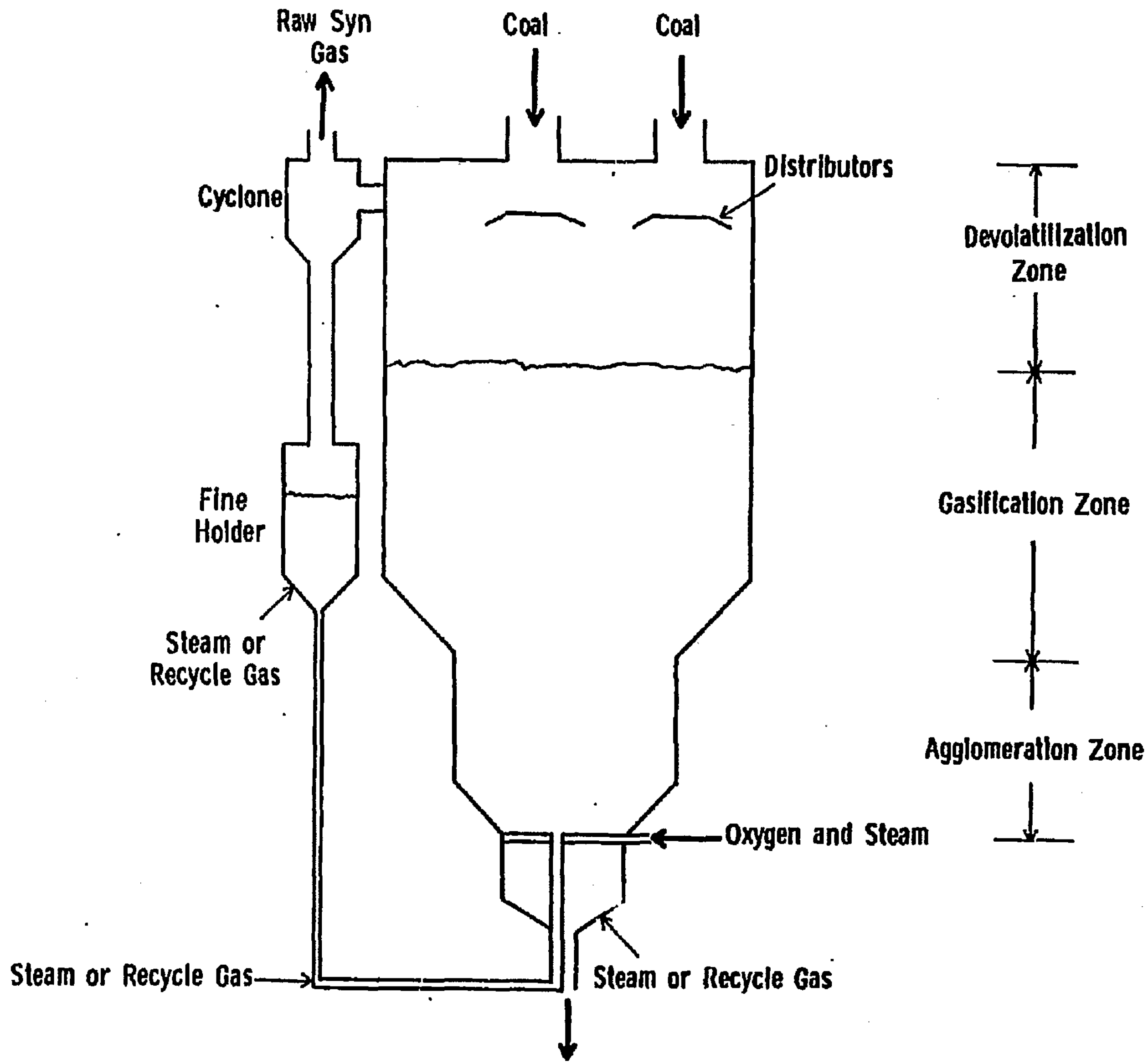


Figure 36

SCHEMATIC FLUID-BED GASIFIER



Let us look at some of the properties of current fluid bed gasifiers and examine some of the reasons that they fall short of our goals. The IGT and Synthane gasifiers(4, 13, 14) fail to utilize one of the major advantages of a fluidized bed — namely, good heat dispersal by rapid solid mixing. Like the dry ash Lurgi gasifier, these gasifiers have to rely on excess steam to control the maximum temperature. The amount of steam used is in excess of what originally planned and is even in excess of that required by the dry ash Lurgi gasifier. The Westinghouse gasifier prevents this problem by using high velocities in the inlet region to promote mixing.

The Synthane pilot unit has demonstrated that tars and phenols hydrocrack to hydrogen, methane, and char when coal is introduced into the bottom of the top bed when its temperature is about 1400°F. However, the synthane pilot unit operates at a high pressure (600 psia) and uses a large excess of steam. It is not known how much cracking of tars and phenols would occur in the absence of this high pressure steam. The elimination of the tars would improve the quality of the recoverable heat since no quench is needed to prevent tar condensation in the heat exchanger.

The problem of feeding almost any kind of coal to a fluid bed at 400 psia has been solved in the Synthane pilot plant. Eastern coals require pretreatment, unless they are fed to a high velocity zone. However, the fluid bed gasifier probably will show the most advantage over the BGC-Lurgi slagger for processing coals with high reactivity and high moisture content such as western coals, lignites and peat.

The fluid bed gasifier can have a problem with fines different from that of feeding coal fines to the reactor. Very non-reactive char fines can be formed in the gasification process. These fines are recirculated to the unit by the cyclones and their concentration can build up to such an extent that a substantial lowering of the effective density of the bed occurs. This problem is well known in the history of the fluid bed gasifier. For example, the Synthane and Winkler gasifiers have this problem. It was solved in the Winkler by introducing some oxygen near the top of the bed to combust these fines. This decreases the thermal efficiency of the Winkler gasifier.

The problem might be solved by reintroducing the fines into the hot combustion zone of the gasifier. Also, if the gasification zone was operated sufficiently hot, the problem might be avoided. This would require an agglomerating bed. The original design of the Hygas gasifier contained such an agglomerating bed and was based on and is similar to the design used in the U-Gas gasifier. The agglomerates are removed through the feed nozzle for the steam-oxygen mixture[14]. It is well known that solid removal from a reactor must be independently adjustable and must not be strongly dependent on particle size distribution; otherwise, it will cause difficult control problems. Such is not true for the above design and we, therefore, have strong reservations about it. However, such adjustability and size independence has been achieved in the Westinghouse pilot plant.

If we solely look at the performance that could be achieved by various gasifiers, a well designed fluid bed gasifier has the best potential to become the versatile workhorse of the industry. It would have to overcome first several development problems, the most critical of which is an agglomerating bottom zone that permits high conversion. However, in terms of its present status the fluid bed gasifier is less advanced than either the BGC-Lurgi slagger or the Texaco gasifier, and the chance that such a gasifier will actually be developed is uncertain. Its high potential would justify a strong effort in that direction.

The fluid bed gasifier has the best development potential of all the gasifiers considered except the BGC-Lurgi slagger. It still awaits good engineering development, however. The most important item to develop is a good agglomerating bottom zone that gives high conversion.

The data in Table 17 summarize the costs for fuel gas and syngas production from the various gasifiers that have been considered in detail and are at a stage where sufficient information for cost estimates exist. The BGC-Lurgi slagger clearly offers the lowest cost and most efficient operation if low H_2 to CO ratio gases can be used and if the desired coal can be gasified in it.

Table 17

Cost of Fuel Gas From Various Gasifiers
Utility Financing DOE Guidelines (1977)

	D.A. Lurgi SNG Western Coal	D.A. Lurgi Fuel Gas Western Coal	Two Stage Fluid Bed or BGC-Lurgi Slagger[c] Western Coal	BGC-Lurgi Slagger Eastern Coal	D.A. Lurgi Eastern Coal	Texaco H_2O/Coal=0.5	Texaco H_2O/Coal=0.85
Investment per[a] Daily MMBTU	3150	2260	1600	1590	2850	2140	2480
Cost of Coal[b] per MMBTU Product	0.80	0.70	0.63	1.28	1.64	1.33	1.49
Total Operating[c] Cost per MMBTU Product including all Capital Charges	2.67	1.92	1.36	1.35	2.41	1.82	2.10
Total Cost	3.47	2.62	1.99	2.63	4.05	3.15	3.59

a) Not including contingencies, working capital or interest during construction (the last two are included in the total operating cost).

b) Western coal at $.5 per million BTU and Eastern coal at $1.00 per million BTU.

c) A western coal with gasification properties similar to Frances coal.

XV. The Use of External Shift with Low H_2 to CO Ratio Gasifiers

Some of the syngas conversion processes for the production of liquid fuels such as the SASOL Fischer-Tropsch and methanol synthesis require syngas with H_2 to CO ratio greater than two. On the other hand, some syngas conversion processes can produce high quality transportation fuels from synthesis gas with much lower ratios of H_2 to CO. The slurry Fischer-Tropsch conversion process requires a ratio of only 0.6 and dimethylether production with subsequent conversion to high octane gasoline by the Mobil process requires a ratio in the neighborhood of unity.

The slurry Fischer-Tropsch process requires little or no additional external shift of the CO with steam to produce additional H_2 when coupled with gasifiers that operate near point A (H_2 to CO ratio near 0.5). While the dimethylether process requires some additional shift, the production of methanol and the SASOL-type Fischer-Tropsch require considerable additional shift. The question then arises as to whether or not the advantages obtained by the use of low H_2 to CO ratio gasifiers disappear when an external shift reactor is employed with the low H_2 to CO ratio gasifier to obtain high H_2 to CO ratio syngases. The answer is that for many cases considerable advantages can remain while others are approximately a break-even proposition.

The differences in steam requirement will serve to illustrate the situation. The two dry ash Lurgi gasifiers of Table 9 will be compared to a BGC-Lurgi slagging gasifier coupled with an external shift reactor. One of the dry ash Lurgi gasifier examples in Table 18 uses eastern coal and produces a syngas with a H_2 to CO ratio of 2.6. The other uses western coal and produces a ratio of 2.1. Because of the difference in reactivity of the coals, the eastern coal requires considerably more excess steam than the western coal. The eastern coal requires 2.6 lbs steam per lb of dry ash-free coal while the western coal requires only 1.6 lbs of steam per lb of dry ash-free coal.

On the other hand, the BGC-Lurgi slagger produces a syngas with a H_2 to CO ratio of 0.5 and requires only 0.35 lbs of steam per lb of coal. The amount of steam required to shift the syngas to a ratio of 2.1 is 0.85 lbs steam per lb of dry ash-free coal; and to a ratio of 2.6, 1.0 lbs of steam per lb of coal is required. This requirement already allows for an amount of excess steam in the shift reactor equal to that converted to suppress the Boudouard reaction (Reaction (8)). Thus to produce a H_2/CO ratio of 2.6 requires 1.35 lbs of steam per lb of coal for a BGC-Lurgi slagger plus external shift reaction while the dry ash Lurgi gasifier requires 2.6 lbs of steam per lb of coal for eastern coal. This represents considerable savings. On the other hand, for the more reactive western coal, the situation is more nearly even: 1.2 lbs steam vs. 1.6 lbs of steam per lb of coal. Thus, the situation needs to be determined on an individual basis.

Table 18

Effect of Use of External Shift with Low H_2/CO Ratio Gasifiers

Gasifier	H_2/CO	Steam Required lb/lb DAF Coal	External Shift Steam Required lb/lb DAF Coal	Total Steam Required to Produce H_2/CO of Dry Ash Lurgi
Eastern Coal				
Dry Ash Lurgi	2.6	2.6	0	2.6
BGC-Lurgi Slagger	0.5	0.35	1.0	1.35
Western Coal				
Dry Ash Lurgi	2.1	1.6	0	1.6
BGC-Lurgi Slagger	0.5	0.35	0.85	1.2

XVI. Summary, Conclusions and Recommendations

The operability of a gasifier is the prime consideration since a gasifier that operates poorly or not at all is of no use no matter how thermally efficient or cost effective it is in principle. Nevertheless, in the search for practical gasifiers, consideration of basic scientific and engineering principles can serve as a guide as to the operating conditions that give the best thermal efficiency and lowest potential cost if gasifiers can be made to operate practically in the manner required. The examination of the basic stoichiometric, thermal, equilibrium and kinetic constraints that apply to all gasifiers (Section III to VII) shows that the most thermally efficient operations are obtained with gasifiers that operate at low steam to oxygen ratios and give low H_2 to CO ratios (point A of Figure 20a). Such gasifiers give the best utilization of the steam. However, the thermal efficiency will be decreased if the low steam to oxygen ratios are obtained by feeding excess amounts of oxygen to the gasifier so that the temperature of the exit gas becomes high and the gasifier becomes, in part, an oxygen-fired coal combustor that supplies heat for steam generation. It is more thermally efficient to obtain steam by a high-efficiency boiler using air.

For syngas production, operating at too low a pressure has both a thermal efficiency and a cost penalty. The major part of the poorer thermal efficiency (Table 7) for the low pressure Koppers-Totzek and Winkler gasifiers is caused by the compression losses required to compress the gases to 400 psia (Figure 32). Too high a pressure has an increased cost penalty also. The best range of pressure appears to be from 200 to 500 psia. This pressure range for the coal gasification is also the best for fuel gas production delivered at 50 psia, if a power recovery turbine can be used to expand the gas from 200-500 psia to 50 psia (Figure 31).

There is no gain in thermal efficiency over the operation in the neighborhood of point A in promoting direct methane formation from the carbon of the coal (Section VIII and Figures 25c-e) even though simple stoichiometric considerations indicate a higher efficiency (Section III). The high excess steam requirements imposed by equilibrium constraints are responsible for this decrease in thermal efficiency. On the other hand, the methane formed during devolatilization is obtained with little thermal penalty. When it can be used or sold, it should not be reformed to CO and H_2 since reforming the methane leads to a decrease in thermal efficiency (Section X and Figure 28). Thus gasifiers with a good devolatilization zone are more thermally efficient.

The tars and phenols present problems when devolatilization zones exist. To take full advantage of the presence of devolatilization zones, better methods of handling these materials are needed.

The reactivity and other properties of the coal can restrict the operating conditions so that the best thermal efficiency may not be obtainable for certain designs. An example of this is the need for large amounts of excess steam for eastern coals with the dry ash Lurgi gasifier.

Examination of the characteristics of gasifiers for which sufficient reliable data exist shows that real gasifiers follow the pattern of behavior required by the basic stoichiometric, thermal, equilibrium and kinetic constraints developed in this study (Section XIV). Furthermore, the patterns of the sources of decreased thermal efficiency are reflected in correspondent patterns of increases in cost (Tables 7, 8, 12, 13, 14, and 16).

There are syngas conversion processes that can use the low H_2 to CO syngas with little or no further shift. The slurry Fischer-Tropsch typically can use a ratio of 0.6 and the production of dimethylether for conversion to gasoline by the Mobil process can use as low a ratio as unity. For processes, such as conversion to methanol and SASOL Fischer-Tropsch, considerably higher ratios are required. In many cases this can be most efficiently and cost effectively supplied by the thermally efficient low H_2/CO ratio gasifier with an external shift reactor.

The BGC-Lurgi slagging gasifier, which is presently close to commercialization, conforms closely to the basic requirments and potentially could produce syngas at a low production cost relative to other gasifiers such as the dry ash Lurgi gasifier. The H_2 to CO and steam to oxygen ratios of this gasifier are about 0.5 and 1.3, respectively. These values are very close to the best theoretical values of 0.45 and 1.6 respectively.

There is considerable evidence from the data obtained from fluid bed gasifier pilot plants that a multi-staged fluid bed gasifier could be developed which operates at low H_2 to CO and steam to oxygen ratios required by the basic concepts. However, a better conceptual design with considerably further development is required before the multi-staged fluid bed gasifier is ready for large-scale pilot plant tests and commercialization. Potentially such a gasifier could operate on coals not well suited for the BGC-Lurgi slagging gasifier.

For syngas containing no methane, the Texaco gasifier closely approaches the basic requirements but has a high oxygen demand that reduces its thermal efficiency and increases the cost over that of gasifiers such as the BGC-Lurgi slagger.

The results of this study show the need for an aggressive effort to complete the development of the BGC-Lurgi slagging gasifier. The BGC-Lurgi slagger must be shown to be scalable to commercial size and to be operable for long periods of time. The range of coals for which it can be used needs to be established. The development of other gasifiers such as multi-staged fluid bed gasifiers that operate in the high thermal efficiency region should be very actively pursued. Such gasifiers would not only be the thermally most efficient but would also provide the most cost effective route to the production of high quality clean transportation fuels as well as clean industrial fuel gas.

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APPENDIX A

GUIDELINES FOR GASIFIER THERMAL EFFICIENCY CALCULATION

I. Objective

To calculate gasifier thermal efficiency in a realistic and consistent way.

II. Process Scheme Cuideline

To compare different gasifiers, it is essential to have a consistent process scheme that eliminates the variation of processing steps attached to each gasifier and that reflects a realistic picture on the energy inputs and outputs of the system. Eypothetical process schemes were thus set up for this purpose. Depending on the raw gas pressure after its cooling, two schemes were constructed:

Case 1A - Low and Medium Pressure Gasifier, (Raw Gas Pressure After Cooling <460 psia).

Case 1B - High Pressure Gasifier (Raw Gas Pressure After Cooling $\geq$460 psia).

These schemes are given as Figures A-1 and A-2. The important points incorporated into these schemes are:

1. Oxygen is available at 1 ATM and 100°F.

2. Clean gas at 400 psia and 100°F is the final product.

3. It takes 60 psia pressure drop to drive the raw gas through the purification stage.

4. For clean gas at a pressure higher than 400 psia, energy is recovered by heating the gas to 320°F and then expanding it to 400 psia in a single stage expander.

III. Thermal Energy Calculation Guidelines

To reflect a realistic thermal energy content for all input-output streams, the following guidelines were given for their calculations.

1. Chemical energy from coal, syn gas, tar, and other chemicals —— use low heating value (LHV) at 77°F, 1 ATM.

2. Eigh sensible heat — heat above 700°F for output streams and 600°F for input streams.

3. Low sensible heat in output streams — all heat between 350 to 700°F, including heat of condensation. Multiply this heat by 0.45, i.e., discounting it by 55% to make this low potential heat equivalent to the high potential heat. For input streams, the low sensible heat covers 250 to 600°F.

4. If there is tar in the gasifier effluent, skip Items 2 and 3 since the stream must now be quenched to remove the tar. The water condensed from the effluent is used for this quenching. For the present purpose, assume that the recycled water temperature is 50°F below the dew point. If this dew point is above 350°F, go to Item 3 to calculate the amount of low sensible heat.

5. Thermal energy used in producing oxygen at 1 ATM — 170 BTU (LHV fuel) per SCF oxygen.

6. State of oxygen for gasifier (if not given) — same pressure as the pressure at which the steam is delivered to the gasifier, and the temperature as the discharge temperature from the oxygen compressor.

7. Compression work —

   - Calculate the theoretical work required by limiting compression ratio at each stage at $\leq$ 3.

   - Calculate the fuel BTU equivalence of the actual work by assuming 25% efficiency.

8. State of steam for gasifier (if not given) —

   Pressure = 25+ gasifier pressure if gasifier pressure >50 psia.

   Pressure = 20+ gasifier pressure if gasifier pressure $\leq$50 psia.

9. Thermal Energy for steam —

   - Calculate the maximum work extractable from isentropic expansion of the steam to 2 psia.

   - Calculate the fuel BTU from the useful work using 37% efficiency.

10. Thermal energy from expansion work obtained from clean syn gas —

   - Take dry clean gas at 320°F and at a pressure 60 psia less than the pressure of the raw syn gas after cooling.

   - Calculate the theoretical work recoverable from this gas by expanding it to 400 psia in a single state expander.

   - Calculate the thermal energy in fuel equivalent basis from this work using 37% efficiency.

11. Normalize the heat loss to 0.5% of the net coal LHV feed.

12. No utility consumption for raw gas purification is accounted for.

The treatment in Item 9 provides one way of differentiating the steams of different quality. Item 10 gives the minimum work recoverable from the gas expansion.

IV. Definition of Three Thermal Efficiencies

Many types of thermal efficiencies can be used for gasifier calculation. Three types of efficiencies most often used are defined as follows:

$$\text{Clean Gas Eff.} \equiv \frac{\text{LHV Clean Gas}}{\text{Net Coal LHV}}$$

$$\text{Clean Fuel Eff.} \equiv \frac{\text{LHV (Clean Gas + Naphtha + Oil)}}{\text{Net Coal LHV}}$$

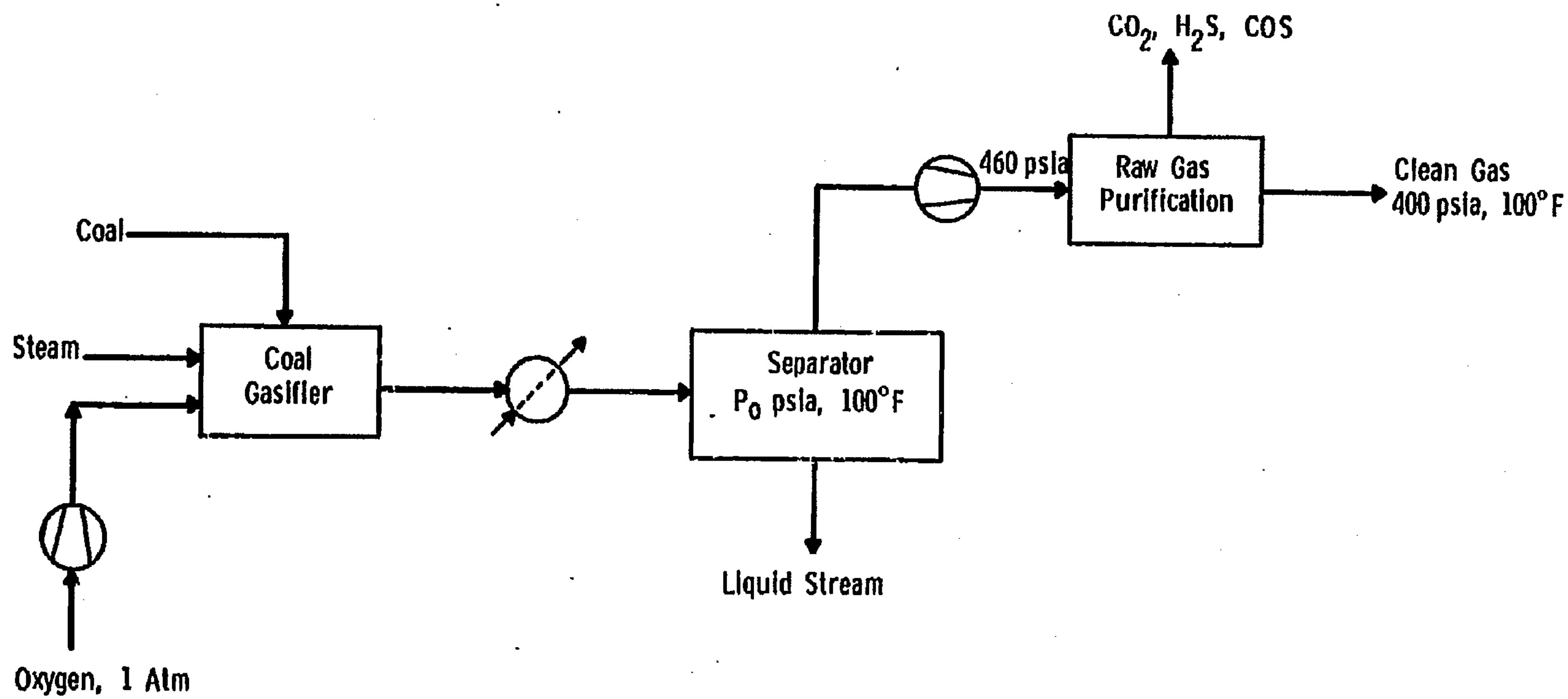
$$\text{Net Eff.} \equiv \frac{\text{All energy Output}^{(1)} - \text{impurity LHV} - \text{All energy input}^{(1)}}{\text{Net Coal LHV}}$$

where net coal LHV ≡ LHV coal - LHV (Char + Tar + Phenol).

(1) Excluding recycle streams at exit state.

Figure A-1

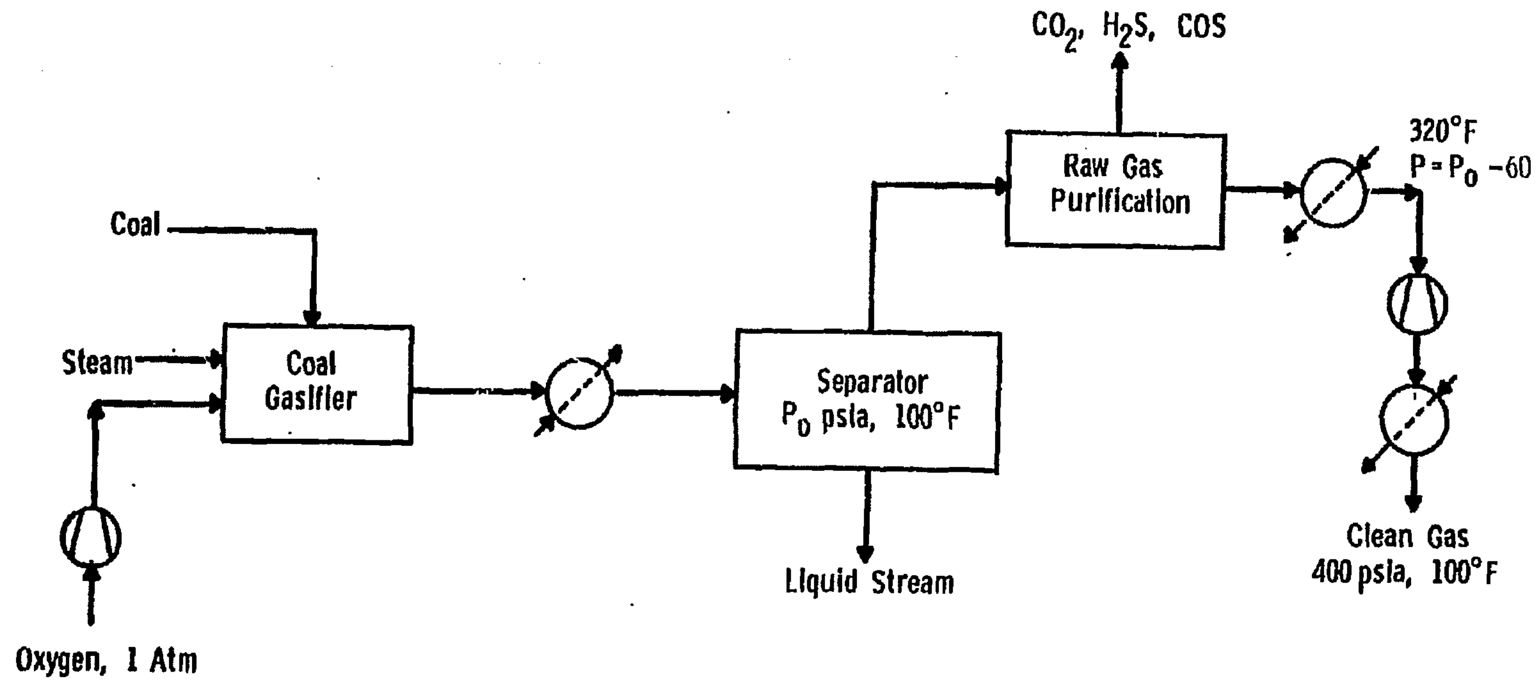
HYPOTHETICAL PROCESS SCHEME AS A CONSISTENT BASIS FOR CALCULATING GASIFIER THERMAL EFFICIENCY - CASE 1A, LOW AND MEDIUM PRESSURE GASIFIER (P_0 < 460 psia)*



* P_0 = Raw Gas Pressure After Cooling

Figure A-2

HYPOTHETICAL PROCESS SCHEME AS A CONSISTENT BASIS FOR CALCULATING GASIFIER THERMAL EFFICIENCY - CASE 1B, HIGH PRESSURE GASIFIER ($P_0 \geq 460$ psia)*



*P_0 = Raw Gas Pressure After Cooling

APPENDIX B

SUMMARY OF DATA FROM

GASIFICATION CALCULATIONS

Unless specified otherwise, the following basis and conditions were used in all calculations,

100 lbs of raw coal or char
steam and oxygen at 700°F
gasifier pressure at 400 psia
no methane made other than from devolatilization

Also, only two types of coal were used. An approximate eastern coal is represented as 65% (wt) fixed carbon, 15% volatile material, 10% moisture and 10% ash; while an approximate western coal is represented as 50% fixed carbon, 10% volatile material, and 30% moisture. For both coals, the volatile material is approximated by methane and accounts for 25% of the coal LHV.

The cases with kinetic constraints were calculated with all the gasification reactions at equilibrium except for the carbon-steam reaction which is given by pseudo-equilibrium conditions in which the pseudo-equilibrium constant is a fraction of the actual equilibrium constant. This fraction is 0.1 unless it is specified otherwise.

I. Model 1 Gasifier (Single Staged, Adiabatic and Completely Mixed) with the Approximate Eastern and Western Coals

Tables B-I-1 and -2 show the cases with the equilibrium constraint for the approximate eastern and western coals, respectively. The similar cases with the kinetic constraint were given in Tables B-I-3 and 4. Table B-I-4 also includes a case at 0.63 steam/oxygen ratio in which the methane formed during devolatilization is reformed further during gasification.

II. Isothermal Indirectly Heated Gasifier with Equilibrium Constaint

Cases with and without methane formation were calculated. They are given in Tables B-II-1 and -2, respectively.

Table B-I-1

Model 1 Gasifier, Equilibrium Constraint,
Approximate Eastern Coal

GASIFIER TEMP	OXYGEN FEED	STEAM FEED	YIELD (LB-MOLE)					THERMAL EFFICIENCY (%)	
(°F)	(LB-MOLE)	(LB-MOLE)	H2	CO	CO2	CH4	H2O	COLD GAS	NET
2995	2.459	0.000	0.553	5.463	0.004	0.894	0.002	83.56	80.37
2439	2.254	0.451	0.983	5.438	0.029	0.894	0.018	86.98	81.24
2106	2.138	0.855	1.328	5.331	0.136	0.894	0.082	88.79	81.47
1946	2.084	1.250	1.610	5.157	0.311	0.894	0.196	89.44	81.23
1853	2.054	1.643	1.861	4.966	0.501	0.894	0.337	89.68	80.83
1789	2.033	2.033	2.091	4.776	0.691	0.894	0.498	89.75	80.36
1730	2.015	2.518	2.356	4.549	0.918	0.894	0.718	99.74	79.73
1684	2.000	3.000	2.600	4.334	1.133	0.894	0.955	89.68	79.07
1612	1.978	3.956	3.038	3.940	1.527	0.894	1.473	89.48	77.73
1557	1.961	4.904	3.423	3.588	1.879	0.894	2.036	89.26	76.36
1511	1.948	5.844	3.764	3.274	2.193	0.894	2.634	89.04	74.99
1438	1.927	7.706	4.345	2.737	2.731	0.894	3.917	88.62	72.25
1379	1.910	9.550	4.816	2.298	3.169	0.894	5.289	88.28	69.51
1329	1.896	11.377	5.202	1.939	3.528	0.894	6.730	87.99	67.21
1286	1.885	13.192	5.521	1.644	3.823	0.894	8.226	87.76	65.26
1248	1.874	14.995	5.784	1.401	4.065	0.894	9.766	87.59	63.36

Table B-I-2

Model 1 Gasifier, Equilibrium Constraint,
Approximate Western Coal

GASIFIER TEMP (°F)	OXYGEN FEED (LB-MOLE)	STEAM FEED (LB-MOLE)	YIELD (LB-MOLE) H2	CO	CO2	CH4	H2O	THERMAL EFFICIENCY (%) COLD GAS	NET
1786	1.660	0.000	1.329	3.553	0.548	0.670	0.336	86.67	79.11
1731	1.647	0.329	1.516	3.392	0.708	0.670	0.479	86.66	78.55
1688	1.636	0.655	1.687	3.242	0.859	0.670	0.633	86.60	77.97
1652	1.628	0.977	1.846	3.100	1.001	0.670	0.796	86.52	77.37
1621	1.621	1.296	1.994	2.966	1.134	0.670	0.968	86.43	76.78
1593	1.614	1.614	2.133	2.840	1.260	0.670	1.147	86.33	76.17
1563	1.607	2.009	2.295	2.692	1.409	0.670	1.379	86.20	75.42
1535	1.601	2.401	2.446	2.553	1.548	0.670	1.620	86.08	74.66
1488	1.590	3.181	2.720	2.301	1.800	0.670	2.125	85.83	73.14
1447	1.581	3.953	2.962	2.077	2.024	0.670	2.657	85.61	71.63
1411	1.574	4.721	3.176	1.879	2.222	0.670	3.210	85.40	70.12
1350	1.561	6.242	3.537	1.544	2.557	0.670	4.371	85.05	67.16
1298	1.550	7.749	3.825	1.277	2.824	0.670	5.590	84.78	64.97
1254	1.541	9.215	4.057	1.063	3.038	0.670	6.853	84.58	62.87
1215	1.533	10.730	4.235	0.891	3.210	0.670	8.150	84.42	60.82
1180	1.526	12.206	4.398	0.752	3.349	0.670	9.473	84.32	58.82

Table B-I-3

Model 1 Gasifier, Kinetic Constraint, Approximate Eastern Coal

GASIFIER TEMP. (oF)	GASIFIER PRESS. (PSIA)	OXYGEN FEED (LB-MOLE)	STEAM FEED (LB-MOLE)	YIELD (LB-MOLE) H2	CO	CO2	CH4	H2O	THERMAL EFFICIENCY (o/o) COLD GAS	NET
3046	400	2.481	0.000	0.538	5.435	0.033	0.894	0.018	83.15	80.20
2627	400	2.331	0.466	0.933	5.339	0.128	0.894	0.080	85.54	80.64
2404	400	2.260	0.904	1.235	5.180	0.287	0.894	0.224	86.52	80.49
2273	400	2.223	1.334	1.488	5.001	0.467	0.894	0.401	86.88	80.10
2184	400	2.201	1.761	1.711	4.821	0.646	0.894	0.605	86.99	79.60
2116	400	2.187	2.187	1.913	4.648	0.819	0.894	0.829	86.99	79.04
2049	400	2.175	2.718	2.143	4.442	1.025	0.894	1.130	86.89	78.31
1995	400	2.167	3.250	2.354	4.247	1.220	0.894	1.451	86.74	77.54
1908	400	2.158	4.315	2.731	3.888	1.579	0.894	2.139	86.38	75.96
1840	400	2.154	5.384	3.061	3.566	1.901	0.894	2.879	85.99	74.34
1783	400	2.153	6.458	3.353	3.276	2.191	0.894	3.659	85.59	72.69
1691	400	2.155	8.618	3.852	2.773	2.694	0.894	5.321	84.83	69.37
1617	400	2.160	10.798	4.259	2.355	3.112	0.894	7.094	84.15	66.88
1554	400	2.166	12.995	4.595	2.007	3.460	0.894	8.955	83.54	64.44
1500	400	2.172	15.206	4.873	1.716	3.751	0.894	10.888	83.02	62.03
1452	400	2.179	17.431	5.104	1.473	3.994	0.894	12.802	92.56	59.65
1650	30	1.983	2.975	2.625	4.343	1.124	0.894	0.905	89.97	72.81
1711	50	2.014	3.021	2.580	4.327	1.140	0.894	0.996	89.44	74.11
1798	100	2.059	3.089	2.512	4.303	1.164	0.894	1.132	88.64	75.38
1853	150	2.088	3.132	2.469	4.288	1.179	0.894	1.218	88.13	76.22
1893	200	2.110	3.165	2.437	4.277	1.190	0.894	1.283	87.74	76.51
1925	250	2.128	3.191	2.412	4.268	1.199	0.894	1.335	87.43	76.88
1951	300	2.142	3.214	2.390	4.260	1.207	0.894	1.379	87.17	77.13
1974	350	2.155	3.233	2.371	4.253	1.214	0.894	1.417	86.95	77.32
1995	400	2.167	3.250	2.354	4.247	1.220	0.894	1.451	86.74	77.54
2013	450	2.177	3.265	2.339	4.241	1.226	0.894	1.481	86.56	77.65
2029	500	2.186	3.280	2.326	4.236	1.231	0.894	1.509	86.40	77.70
2058	600	2.203	3.305	2.302	4.227	1.240	C.894	1.558	86.10	77.76
2104	800	2.230	3.345	2.262	4.211	1.256	0.894	1.638	85.62	77.76
2142	1000	2.252	3.379	2.231	4.199	1.269	0.894	1.703	85.23	77.70

Table B-I-4

Model 1 Gasifier, Kinetic Constraint,
Approximate Western Coal

GASIFIER TEMP	OXYGEN FEED	STEAM FEED	YIELD (LB-MOLE)					THERMAL EFFICIENCY (%)	
(oF)	(LB-MOLE)	(LB-MOLE)	H2	CO	CO2	CH4	H2O	COLD GAS	NET
2131	1.774	0.000	1.161	3.493	0.607	0.670	0.504	84.00	78.01
2067	1.764	0.353	1.325	3.348	0.752	0.670	0.693	83.93	77.37
2014	1.758	0.703	1.474	3.212	0.889	0.670	0.895	83.81	76.72
1970	1.753	1.052	1.611	3.084	1.017	0.670	1.106	83.67	76.04
1931	1.750	1.400	1.739	2.963	1.138	0.670	1.327	83.51	75.36
1897	1.748	1.748	1.858	2.848	1.253	0.670	1.555	83.34	74.67
1859	1.746	2.182	1.998	2.712	1.388	0.670	1.850	83.12	73.80
1824	1.745	2.617	2.127	2.585	1.516	0.670	2.155	82.91	72.92
1764	1.744	3.487	2.363	2.351	1.749	0.670	2.790	82.48	71.14
1713	1.744	4.360	2.571	2.143	1.958	0.670	3.455	82.07	69.35
1668	1.745	5.236	2.756	1.955	2.146	0.670	4.146	81.68	67.73
1591	1.749	6.997	3.070	1.633	2.468	0.670	5.593	80.98	65.08
1526	1.754	8.770	3.323	1.370	2.731	0.670	7.112	80.37	62.47
1470	1.759	10.553	3.529	1.155	2.946	0.670	8.690	79.85	59.92
1422	1.764	12.345	3.696	0.978	3.123	0.670	10.314	79.40	57.40
1378	1.768	14.144	3.833	0.832	3.268	0.670	11.976	79.03	54.91
2000 *	2.050	1.281	2.619	3.711	0.947	0.113	1.442	82.47	72.68

* WITH METHANE IN V.M. REFORMED

Table B-II-1

Isothermal Indirectly Heated Char Gasifier, Equilibrium Constraint with Methane Formation

GASIFIER TEMP (oF)	GASIFIER PRESS. (PSIA)	STEAM FEED (LB-MOLE)	YIELD (LB-MOLE) H2	CO	CO2	CH4	H2O
1200	100	15.451	5.189	1.934	4.010	2.382	5.497
1300	100	13.931	6.211	3.262	3.270	1.795	4.130
1400	100	12.190	6.873	4.745	2.323	1.259	2.807
1500	100	10.659	7.220	6.049	1.432	0.846	1.747
1600	100	9.579	7.417	6.971	0.789	0.566	1.029
1700	100	8.947	7.579	7.529	0.411	0.386	0.596
1800	100	8.618	7.729	7.845	0.212	0.270	0.349
1900	100	8.455	7.860	8.022	0.112	0.193	0.209
2000	100	8.376	7.965	8.125	0.061	0.141	0.129
1200	200	15.698	3.934	1.363	4.125	2.839	6.086
1300	200	14.542	4.942	2.392	3.605	2.330	4.941
1400	200	13.078	5.776	3.690	2.840	1.797	3.709
1500	200	11.557	6.360	5.027	1.983	1.316	2.564
1600	200	10.278	6.744	6.154	1.234	0.939	1.657
1700	200	9.392	7.026	6.952	0.706	0.669	1.028
1800	200	8.862	7.267	7.457	0.387	0.482	0.630
1900	200	8.575	7.478	7.761	0.212	0.353	0.390
2000	200	8.427	7.655	7.945	0.118	0.263	0.246
1200	400	15.801	2.903	0.953	4.174	3.199	6.499
1300	400	14.921	3.768	1.714	3.819	2.793	5.568
1400	400	13.753	4.595	2.758	3.243	2.325	4.508
1500	400	12.402	5.293	3.972	2.507	1.847	3.415
1600	400	11.004	5.830	5.160	1.751	1.416	2.423
1700	400	10.007	6.241	6.145	1.115	1.067	1.632
1800	400	9.255	6.582	6.857	0.666	0.803	1.067
1900	400	8.792	6.882	7.331	0.385	0.610	0.690
2000	400	8.531	7.147	7.637	0.222	0.467	0.449
1200	1000	15.831	1.896	0.593	4.193	3.541	6.853
1300	1000	15.188	2.526	1.086	3.980	3.260	6.142
1400	1000	14.341	3.197	1.808	3.606	2.912	5.320
1500	1000	13.291	3.855	2.743	3.069	2.514	4.409
1600	1000	12.122	4.447	3.801	2.424	2.101	3.473
1700	1000	10.985	4.953	4.845	1.768	1.714	2.605
1800	1000	10.021	5.387	5.749	1.198	1.379	1.875
1900	1000	9.302	5.775	6.451	0.769	1.107	1.314
2000	1000	8.823	6.131	6.957	0.478	0.891	0.909
1200	1500	15.827	1.562	0.481	4.193	3.653	6.960
1300	1500	15.254	2.076	0.885	4.024	3.418	6.322
1400	1500	14.516	2.683	1.488	3.718	3.120	5.592
1500	1500	13.591	3.282	2.295	3.263	2.769	4.771
1600	1500	12.525	3.850	3.251	2.687	2.388	3.899
1700	1500	11.427	4.359	4.253	2.063	2.010	3.048
1800	1500	10.426	4.808	5.184	1.477	1.665	2.288
1900	1500	9.619	5.211	5.958	0.998	1.371	1.666
2000	1500	9.035	5.586	6.550	0.647	1.129	1.191

Table B-II-2

Isothermal Indirectly Heated Char Gasifier, Equilibrium Constraint Without Methane Formation

GASIFIER TEMP, (oF)	GASIFIER PRESS, (PSIA)	STEAM FEED (LB-MOLE)	YIELD (LB-MOLE) H2	CO	CO2	H2O
1200	100	26.054	13.654	2.999	5.328	12.400
1300	100	19.599	12.271	4.382	3.945	7.328
1400	100	15.072	10.928	5.725	2.602	4.144
1500	100	12.122	9.855	6.797	1.529	2.266
1600	100	10.373	9.147	7.505	0.821	1.226
1700	100	9.419	8.748	7.905	0.421	0.672
1800	100	8.921	8.542	8.111	0.216	0.379
1900	100	8.661	8.440	8.213	0.113	0.222
2000	100	8.523	8.388	8.265	0.062	0.135
1200	200	32.008	14.234	2.418	5.908	17.773
1300	200	24.124	13.012	3.640	4.686	11.112
1400	200	18.419	11.714	4.939	3.380	6.705
1500	200	14.455	10.544	6.109	2.217	3.911
1600	200	11.882	9.649	7.004	1.323	2.233
1700	200	10.335	9.064	7.588	0.738	1.271
1800	200	9.460	8.726	7.927	0.399	0.734
1900	200	8.978	8.543	8.110	0.217	0.435
2000	200	8.713	8.446	8.206	0.120	0.267
1200	400	39.525	14.716	1.937	6.389	24.810
1300	400	29.908	13.669	2.984	5.343	16.239
1400	400	22.840	12.479	4.174	4.153	10.360
1500	400	17.725	11.302	5.351	2.975	6.423
1600	400	14.175	10.289	6.364	1.962	3.886
1700	400	11.850	9.530	7.123	1.203	2.320
1800	400	10.415	9.028	7.625	0.701	1.387
1900	400	9.567	8.726	7.926	0.400	0.841
2000	400	9.077	8.555	8.098	0.229	0.522
1200	1000	52.564	15.218	1.435	6.891	37.346
1300	1000	40.010	14.392	2.261	6.066	25.618
1400	1000	30.728	13.390	3.263	5.064	17.338
1500	1000	23.817	12.305	4.347	3.979	11.512
1600	1000	18.747	11.257	5.396	2.931	7.490
1700	1000	15.142	10.350	6.302	2.024	4.792
1800	1000	12.683	9.647	7.006	1.321	3.036
1900	1000	11.076	9.152	7.500	0.826	1.924
2000	1000	10.062	8.831	7.821	0.505	1.231
1200	1500	59.740	15.398	1.255	7.071	44.342
1300	1500	45.587	14.661	1.992	6.334	30.926
1400	1500	35.125	13.746	2.907	5.419	21.380
1500	1500	27.287	12.725	3.928	4.398	14.562
1600	1500	21.447	11.698	4.955	3.371	9.749
1700	1500	17.188	10.764	5.888	2.438	6.424
1800	1500	14.181	9.996	6.657	1.669	4.185
1900	1500	12.135	9.419	7.234	1.093	2.716
2000	1500	10.790	9.020	7.632	0.694	1.770

III. Model 1 Char Gasifier with Equilibrium Constraint

Cases with and without methane formation were calculated. They are given in Tables B-III-1 and -2, respectively.

IV. Model 2 Gasifier

Tables B-IV-1 and -2 include the cases using the approximate eastern and western coals, respectively. In the gasification zone, the kinetic constraints on the steam-carbon reaction was assumed. The calculations to simulate the equilibrium methane made in the gasification zone at the zone temperature of 1600, 1800 , and 2000 °F are summarized in Table B-IV-3. Table B-IV-4 gives the cases using the approximate eastern coal with equilibrium constraint in the gasification zone.

V. Model 3 Gasifier

Tables B-V-1 and -2 include the cases using the approximate eastern and western coals, respectively, with the kinetic constraint assumed in the gasification zone.

VI. Model 1 and Model 2 Gasifier Using Approximate Western Coal with an Additional 15% (wt) Volatile Material as Carbon Monoxide

Tables B-VI-1 and -2 show, respectively, the results from Model 1 and Model 2 gasifier calculations.

VII. Simulated Texaco Gasifier with Various Coal Slurry Compositions

Calculations were done only for the approximate eastern coal. The gasifier temperature is at 2360°F by varying the steam-carbon reaction conversion. The result is summarized in Table B-VII-1.

Table B-III-1

Model 1 Char Gasifier, Equilibrium Constraint With Methane Formation

GASIFIER TEMP. (°F)	GASIFIER PRESS. (PSIA)	OXYGEN FEED (LB-MOLE)	STEAM FEED (LB-MOLE)	YIELD (LB-MOLE) H2	YIELD (LB-MOLE) CO	YIELD (LB-MOLE) CO2	YIELD (LB-MOLE) CH4	YIELD (LB-MOLE) H2O	THERMAL EFFICIENCY (%) COLD GAS	THERMAL EFFICIENCY (%) NET
3385	400	3.470	1.388	1.383	8.323	0.002	0.002	0.002	82.13	76.45
2769	400	3.208	1.925	1.898	8.307	0.011	0.008	0.012	85.95	77.19
2307	400	3.011	2.409	2.289	8.223	0.074	0.029	0.061	88.64	77.65
2059	400	2.896	2.896	2.568	8.006	0.248	0.072	0.184	89.87	77.65
1908	400	2.807	3.509	2.841	7.643	0.545	0.139	0.391	90.37	77.33
1821	400	2.739	4.109	3.062	7.259	0.854	0.213	0.620	90.52	76.91
1713	400	2.619	5.239	3.390	6.518	1.432	0.377	1.095	90.54	76.09
1642	400	2.507	6.268	3.601	5.841	1.936	0.549	1.548	90.48	75.37
1587	400	2.399	7.197	3.728	5.233	2.370	0.723	2.023	90.43	74.76
1505	400	2.196	8.785	3.808	4.207	3.058	1.061	2.855	90.43	73.86
1441	400	2.013	10.065	3.747	3.396	3.560	1.371	3.575	90.56	73.26
1389	400	1.849	11.096	3.609	2.754	3.924	1.648	4.191	90.77	72.89
1344	400	1.704	11.931	3.431	2.246	4.188	1.892	4.717	91.05	72.67
1305	400	1.577	12.612	3.236	1.842	4.379	2.105	5.166	91.35	72.55
3390	1000	3.471	1.389	1.375	8.318	0.004	0.004	0.005	82.10	77.63
2794	1000	3.216	1.929	1.867	8.282	0.026	0.018	0.026	85.76	78.50
2389	1000	3.037	2.429	2.209	8.146	0.124	0.056	0.108	88.04	78.92
2176	1000	2.929	2.929	2.440	7.893	0.318	0.115	0.250	89.01	78.86
2033	1000	2.838	3.548	2.658	7.512	0.613	0.202	0.486	89.44	78.50
1945	1000	2.764	4.146	2.826	7.120	0.912	0.295	0.731	89.57	78.06
1830	1000	2.632	5.265	3.060	6.371	1.466	0.489	1.227	89.59	77.19
1753	1000	2.510	6.276	3.193	5.690	1.950	0.687	1.708	89.54	76.41
1693	1000	2.395	7.184	3.259	5.081	2.365	0.881	2.164	89.51	75.75
1602	1000	2.182	8.726	3.256	4.059	3.024	1.244	2.983	89.54	75.12
1532	1000	1.992	9.962	3.150	3.256	3.505	1.566	3.681	89.71	74.76
1474	1000	1.826	10.955	2.992	2.624	3.856	1.846	4.271	89.96	74.58
1424	1000	1.680	11.760	2.812	2.127	4.111	2.088	4.771	90.27	74.52
1380	1000	1.552	12.416	2.625	1.733	4.296	2.298	5.196	90.61	74.54

Table B-III-2

Model 1 Char Gasifier, Equilibrium Constraint Without Methane Formation

GASIFIER TEMP. (°F)	GASIFIER PRESS. (PSIA)	OXYGEN FEED (LB-MOLE)	STEAM FEED (LB-MOLE)	YIELD (LB-MOLE) H2	CO	CO2	H2O	THERMAL EFFICIENCY (o/o) COLD GAS	NET
3385	400	3.471	1.388	1.386	8.325	0.002	0.002	82.13	76.44
2767	400	3.211	1.927	1.915	8.315	0.012	0.012	85.95	77.17
2302	400	3.024	2.419	2.355	8.251	0.076	0.064	88.64	77.56
2052	400	2.928	2.928	2.728	8.069	0.258	0.200	89.83	77.39
1902	400	2.873	3.591	3.149	7.757	0.569	0.442	90.25	76.79
1814	400	2.842	4.263	3.537	7.431	0.895	0.726	90.29	76.03
1706	400	2.805	5.609	4.235	6.808	1.518	1.374	90.06	74.38
1634	400	2.780	6.951	4.850	6.242	2.084	2.101	89.71	72.67
1579	400	2.762	8.286	5.398	5.730	2.596	2.887	89.34	70.94
1494	400	2.735	10.939	6.337	4.846	3.480	4.602	88.63	67.47
1430	400	2.714	13.572	7.109	4.115	4.211	6.463	88.01	64.00
1376	400	2.698	16.188	7.750	3.506	4.820	8.438	87.49	60.61
1330	400	2.684	18.791	8.286	2.998	5.328	10.505	87.05	58.07
1290	400	2.673	21.381	8.734	2.573	5.753	12.647	86.69	55.59
3389	1000	3.473	1.389	1.384	8.322	0.004	0.005	82.10	77.62
2791	1000	3.223	[illegible].934	1.[illegible]	8.300	0.026	0.027	85.76	78.45
2381	1000	3.061	2.449	2.332	8.198	0.128	0.117	88.02	78.75
2168	1000	2.983	2.983	2.692	7.994	0.332	0.291	88.92	78.47
2026	1000	2.936	3.670	3.096	7.685	0.642	0.575	89.22	77.76
1937	1000	2.910	4.365	3.464	7.369	0.958	0.901	89.21	76.90
1823	1000	2.880	5.761	4.124	6.768	1.558	1.637	88.89	75.03
1746	1000	2.864	7.159	4.703	6.222	2.104	2.456	88.46	73.11
1686	1000	2.853	8.558	5.219	5.728	2.598	3.339	88.00	71.51
1594	1000	2.839	11.357	6.104	4.870	3.456	5.253	87.12	68.35
1523	1000	2.832	14.159	6.834	4.156	4.171	7.325	86.33	65.26
1465	1000	2.827	16.963	7.441	3.557	4.769	9.521	85.65	62.22
1415	1000	2.824	19.769	7.950	3.054	5.272	11.819	85.06	59.24
1371	1000	2.822	22.577	8.378	2.631	5.695	14.199	84.56	56.30

Table B-IV-1

Model 2 Gasifier Using Approximate Eastern Coal

TEMPERATURE (oF)		OXYGEN FEED	STEAM FEED	YIELD (LB-MOLE)					THERMAL EFFICIENCY (o/o)	
GASIFIER	EXIT	(LB-MOLE)	(LB-MOLE)	H2	CO	CO2	CH4	H2O	COLD GAS	NET
4055	2251	2.280	0.912	0.908	5.465	0.002	0.894	0.559	86.58	79.34
3618	1830	2.109	1.265	1.255	5.461	0.006	0.894	0.566	89.46	80.27
3182	1483	1.970	1.576	1.545	5.449	0.018	0.894	0.586	91.78	80.98
2795	1243	1.873	1.873	1.782	5.406	0.062	0.894	0.647	93.35	81.35
2487	1114	1.816	2.271	2.021	5.281	0.186	0.894	0.805	94.13	81.22
2318	1084	1.796	2.694	2.229	5.113	0.354	0.894	1.020	94.24	80.75
2133	1097	1.786	3.573	2.603	4.759	0.708	0.894	1.525	93.89	79.50
2022	1126	1.789	4.473	2.935	4.421	1.046	0.894	2.093	93.36	78.10
1941	1152	1.796	5.387	3.235	4.108	1.359	0.894	2.707	92.80	76.65
1822	1189	1.813	7.252	3.758	3.550	1.917	0.894	4.049	91.71	73.67
1733	1209	1.832	9.160	4.199	3.072	2.395	0.894	5.516	90.70	70.65
1660	1216	1.851	11.103	4.572	2.661	2.806	0.894	7.086	89.80	68.33
1599	1216	1.868	13.079	4.890	2.307	3.160	0.894	8.744	88.99	66.03
1545	1210	1.885	15.083	5.160	2.003	3.464	0.894	10.478	88.27	63.76

Table B-IV-2

Model 2 Gasifier Using Approximate Western Coal

TEMPERATURE (oF)		OXYGEN FEED	STEAM FEED	YIELD (LB-MOLE)					THERMAL EFFICIENCY (o/o)	
GASIFIER	EXIT	(LB-MOLE)	(LB-MOLE)	H2	CO	CO2	CH4	H2O	COLD GAS	NET
4055	1502	1.710	0.684	0.681	4.099	0.002	0.670	1.668	86.57	75.74
3618	1148	1.582	0.949	0.941	4.096	0.004	0.670	1.673	89.46	76.80
3182	860	1.478	1.182	1.159	4.087	0.014	0.670	1.689	91.78	77.60
2795	666	1.405	1.405	1.337	4.055	0.046	0.670	1.734	93.35	78.17
2487	578	1.362	1.703	1.516	3.961	0.140	0.670	1.853	94.13	78.44
2318	578	1.347	2.021	1.672	3.835	0.265	0.670	2.014	94.24	78.01
2133	637	1.340	2.680	1.952	3.570	0.531	0.670	2.393	93.89	76.56
2022	703	1.342	3.355	2.201	3.316	0.785	0.670	2.819	93.36	74.89
1941	761	1.347	4.041	2.427	3.081	1.020	0.670	3.279	92.80	73.45
1822	849	1.360	5.440	2.819	2.663	1.438	0.670	4.286	91.70	70.76
1733	909	1.374	6.870	3.149	2.304	1.797	0.670	5.386	90.70	68.41
1660	949	1.388	8.328	3.430	1.996	2.105	0.670	6.564	89.80	66.09
1599	975	1.401	9.810	3.668	1.731	2.370	0.670	7.808	88.99	63.80
1545	991	1.414	11.313	3.871	1.503	2.598	0.670	9.108	88.26	61.53

Table B-IV-3

Model 2 Gasifier Using Approximate Eastern Coal
With and Without Methane Formation in the Gasification Zone

TEMPERATURE (oF) GASIFIER	EXIT	OXYGEN FEED (LB-MOLE)	STEAM FEED (LB-MOLE)	YIELD (LB-MOLE) H2	CO	CO2	CH4	H2O	THERMAL EFFICIENCY (o/o) COLD GAS	NET
WITH METHANE FORMATION IN GASIFICATION ZONE :										
1600	1205	1.597	12.774	3.719	1.790	3.119	1.452	8.494	89.62	68.70
1800	1443	2.088	16.707	4.019	1.857	3.390	1.114	12.802	83.37	61.71
2000	1673	2.590	20.723	3.710	1.736	3.654	0.971	17.414	75.55	54.96
WITHOUT METHANE FORMATION IN GASIFICATTON ZONE :										
1600	1263	1.954	15.635	5.020	2.006	3.462	0.894	11.170	[illegible].11	62.90
1800	1464	2.252	18.013	4.484	1.947	3.529	0.894	14.084	82.[illegible]1	59.15
2000	1679	2.657	21.256	3.853	1.767	3.700	0.[illegible]94	17.958	74.92	53.95

Table B-IV-4

Model 2 Gasifier Using Approximate Eastern Coal
With Equilibrium Constraint in the Gasification Zone

TEMPERATURE (°F)		OXYGEN FEED	STEAM FEED	YIELD (LB-MOLE)					THERMAL EFFICIENCY (%)	
GASIFIER	EXIT	(LB-MOLE)	(LB-MOLE)	H2	CO	CO2	CH4	H2O	COLD GAS	NET
4051	2246	2.278	0.911	0.911	5.467	0.000	0.894	0.555	86.62	79.36
3607	1816	2.103	1.262	1.261	5.467	0.001	0.894	0.556	89.56	80.32
3140	1444	1.955	1.564	1.560	5.465	0.002	0.894	0.559	92.07	81.13
2650	1129	1.831	1.831	1.816	5.456	0.011	0.894	0.570	94.14	81.76
2166	903	1.738	2.172	2.082	5.377	0.090	0.894	0.645	95.60	82.00
1958	856	1.707	2.560	2.318	5.203	0.264	0.894	0.797	95.87	81.65
1785	871	1.685	3.371	2.754	4.809	0.658	0.894	1.171	95.67	80.55
1693	901	1.675	4.188	3.147	4.437	1.030	0.894	1.597	95.30	79.35
1628	928	1.669	5.006	3.500	4.096	1.371	0.894	2.060	94.93	78.12
1535	966	1.660	6.639	4.114	3.500	1.967	0.894	3.080	94.22	75.64
1466	987	1.654	8.268	4.627	3.000	2.467	0.894	4.196	93.60	73.16
1410	998	1.649	9.895	5.059	2.577	2.890	0.894	5.391	93.07	70.67
1362	1001	1.645	11.518	5.424	2.219	3.248	0.894	6.649	92.62	68.52
1321	1001	1.642	13.139	5.733	1.916	3.551	0.894	7.961	92.24	66.70
1284	997	1.640	14.758	5.995	1.660	3.808	0.894	9.318	91.91	64.93

Table B-V-1

Model 3 Gasifier Using Approximate Eastern Coal

TEMPERATURE (oF)		OXYGEN FEED	STEAM FEED	YIELD (LB-MOLE)					THERMAL EFFICIENCY (o/o)	
GASIFIER	EXIT	(LB-MOLE)	(LB-MOLE)	H2	CO	CO2	CH4	H2O	COLD GAS	NET
3929	2270	2.281	0.912	0.908	5.464	0.003	0.894	0.560	86.56	79.45
3297	1882	2.116	1.270	1.249	5.453	0.014	0.894	0.576	89.33	80.45
2807	1604	2.003	1.602	1.523	5.406	0.061	0.894	0.634	91.17	80.93
2523	1462	1.945	1.945	1.745	5.299	0.168	0.894	0.755	91.99	80.89
2334	1386	1.912	2.390	1.985	5.125	0.342	0.894	0.960	92.30	80.47
2220	1351	1.896	2.844	2.202	4.941	0.526	0.894	1.197	92.31	79.90
2077	1323	1.881	3.762	2.588	4.584	0.883	0.894	1.729	92.05	78.60
1981	1313	1.876	4.690	2.928	4.254	1.213	0.894	2.317	91.66	77.21
1909	1308	1.875	5.626	3.232	3.951	1.516	0.894	2.949	91.23	75.77
1798	1299	1.881	7.522	3.758	3.415	2.053	0.894	4.319	90.37	72.82
1714	1289	1.890	9.450	4.198	2.956	2.511	0.894	5.807	89.56	69.99
1645	1277	1.901	11.407	4.569	2.563	2.904	0.894	7.393	88.80	67.71
1586	1262	1.913	13.391	4.884	2.224	3.243	0.894	9.063	88.12	65.45
1534	1246	1.925	15.400	5.151	1.934	3.533	0.894	10.804	87.50	63.21

Table B-V-2

Model 3 Gasifier Using Approximate Western Coal

TEMPERATURE (oF)		OXYGEN FEED	STEAM FEED	YIELD (LB-MOLE)					THERMAL EFFICIENCY (o/o)	
GASIFIER	EXIT	(LB-MOLE)	(LB-MOLE)	H2	CO	CO2	CH4	H2O	COLD GAS	NET
3658	1571	1.713	0.685	0.679	4.097	0.004	0.670	1.671	86.52	76.18
3056	1263	1.596	0.957	0.931	4.080	0.021	0.670	1.692	89.12	77.15
2654	1077	1.525	1.220	1.130	4.021	0.080	0.670	1.755	90.59	77.47
2437	998	1.491	1.491	1.295	3.924	0.177	0.670	1.861	91.17	77.32
2283	963	1.470	1.[illegible]	1.477	3.784	0.316	0.670	2.025	91.38	76.85
2183	954	1.458	2.187	1.642	3.643	0.458	0.670	2.210	91.38	76.26
2052	961	1.446	2.891	1.936	3.374	0.726	0.670	2.620	91.14	74.95
1962	977	1.440	3.600	2.193	3.128	0.973	0.670	3.072	90.79	73.55
1892	993	1.438	4.314	2.423	2.902	1.198	0.670	3.556	90.41	72.11
1786	1020	1.439	5.756	2.818	2.505	1.596	0.670	4.603	89.63	69.67
1703	1038	1.443	7.217	3.148	2.167	1.934	0.670	5.735	88.88	67.37
1635	1048	1.450	8.697	3.425	1.878	2.223	0.670	6.937	88.19	65.10
1577	1053	1.456	10.195	3.659	1.629	2.471	0.670	8.201	87.56	62.85
1526	1054	1.464	11.709	3.858	1.416	2.684	0.670	9.516	86.99	60.62

Table B-VI-1

Model 1 Gasifier Using Approximate Western Coal
With an Additional 15% (wt) Volatile Material as Carbon Monoxide

GASIFIER TEMP	OXYGEN FEED	STEAM FEED	YIELD (LB-MOLE)					THERMAL EFFICIENCY (%)	
(°F)	(LB-MOLE)	(LB-MOLE)	H_2	CO	CO_2	CH_4	H_2O	COLD GAS	NET
1957	1.332	0.000	0.952	2.624	0.763	0.670	0.713	83.38	78.76
1916	1.328	0.266	1.052	2.532	0.856	0.670	0.879	83.28	78.16
1860	1.325	0.530	1.145	2.445	0.942	0.670	1.050	83.17	77.55
1847	1.322	0.793	1.231	2.364	1.024	0.670	1.227	83.05	76.94
1817	1.320	1.056	1.312	2.287	1.100	0.670	1.410	82.92	76.32
1790	1.319	1.319	1.388	2.214	1.173	0.670	1.596	82.80	75.70
1759	1.317	1.646	1.476	2.129	1.259	0.670	1.836	82.64	74.93
1730	1.316	1.974	1.559	2.048	1.339	0.670	2.081	82.49	74.14
1678	1.315	2.629	1.708	1.902	1.485	0.670	2.587	82.20	72.57
1633	1.314	3.285	1.839	1.772	1.615	0.670	3.111	81.92	71.03
1592	1.314	3.942	1.955	1.656	1.731	0.670	3.652	81.66	69.88
1522	1.314	5.258	2.150	1.460	1.927	0.670	4.773	81.20	67.61
1462	1.316	6.578	2.305	1.303	2.084	0.670	5.939	80.81	65.39
1410	1.317	7.902	2.428	1.177	2.211	0.670	7.139	80.48	63.21
1364	1.318	9.229	2.528	1.075	2.313	0.670	8.366	80.22	61.06
1324	1.320	10.557	2.608	0.992	2.395	0.670	9.614	79.95	58.93

Table B-VI-2

Model 2 Gasifier Using Approximate Western Coal
With an Additional 15% (wt) Volatile Material as Carbon Monoxide

TEMPERATURE (oF)		OXYGEN FEED	STEAM FEED	YIELD (LB-MOLE)					THERMAL EFFICIENCY (o/o)	
GASIFIER	EXIT	(LB-MOLE)	(LB-MOLE)	H2	CO	CO2	CH4	H2O	COLD GAS	NET
4055	951	1.190	0.476	0.474	3.386	0.001	0.670	1.667	88.91	77.31
3618	684	1.100	0.660	0.654	3.384	0.003	0.670	1.671	91.29	78.43
3182	467	1.028	0.822	0.806	3.378	0.010	0.670	1.682	93.21	80.05
2795	325	0.977	0.977	0.929	3.355	0.032	0.670	1.713	94.51	81.13
2487	270	0.947	1.184	1.054	3.290	0.097	0.670	1.796	95.15	81.55
2318	286	0.937	1.405	1.163	3.203	0.185	0.670	1.908	95.24	81.17
2133	362	0.932	1.864	1.358	3.018	0.369	0.670	2.171	94.95	79.78
2022	441	0.933	2.333	1.531	2.842	0.546	0.670	2.467	94.52	78.40
1941	511	0.937	2.810	1.688	2.670	0.709	0.670	2.788	94.05	76.95
1822	621	0.946	3.783	1.960	2.387	1.000	0.670	3.488	93.15	74.35
1733	700	0.956	4.778	2.190	2.138	1.250	0.670	4.253	92.32	71.97
1660	757	0.965	5.792	2.385	1.923	1.464	0.670	5.072	91.57	70.07
1599	798	0.975	6.822	2.551	1.739	1.648	0.670	5.937	90.90	68.18
1545	828	0.983	7.868	2.692	1.580	1.807	0.670	6.841	90.31	66.31

Table B-VII-1

Simulated Texaco Gasifier with Various Coal Slurry Compositions(Approximate Eastern Coal)

WATER FEED (LB)	OXYGEN FEED (LB-MOLE)	YIELD (LB-MOLE)					THERMAL EFFICIENCY (o/o)	
		H2	CO	CO2	CH4	H2O	COLD GAS	NET
40	2.982	2.952	5.424	0.895	0.042	1.526	79.58	72.67
50	3.079	2.995	5.191	1.129	0.041	2.041	77.62	71.00
60	3.178	3.022	4.972	1.349	0.040	2.571	75.65	69.31
70	3.277	3.037	4.764	1.558	0.039	3.114	73.69	67.63
80	3.377	3.041	4.566	1.757	0.037	3.668	71.73	65.94
90	3.478	3.035	4.378	1.948	0.036	4.232	69.77	64.25
100	3.579	3.020	4.196	2.131	0.034	4.805	67.81	62.53

APPENDIX C

SUMMARY ON GASIFIER INFORMATION

Gasifiers	Reference Indexes[a]	
	Analysis Done[b]	Insufficient Data for Analysis
Battelle/Carbide	-	9,11,12,19,24, 33
BGC-Lurgi Slagger	2(45),51,53	12,13,16,18,49
Dry Ash Lurgi	1(2),4,17,26,42	9,10,12,14,19, 20,21,27,28,29
Texaco	45,58	-
Fluidized Bed Synthane	5,5,21,44(32),56	6,9,10,12,13 15,19,31,36,37,43
IGT-U-Gas	1	9,17,19,34,35,36
IGT-Hygas	38,54,55	9,11,13,14,19, 36,37,39,40,43
Winkler	52	9,15,19
Bi-Gas, High Pres.	3,3,21,23	9,11,14,19,20, 37,43
Med. Pres.	17,25,45(2)	8
Others	-	13,15
Combustion Engrg.	1	9,11,14,19,22, 36,37,39,40
Koppers-Totzek	26	41,50

(a) In quoting the references given in this Appendix, the reference indexes will be preceded by a capital letter "C".
(b) The reference within the parenthesis contains design data that are so similar to those of the reference preceding the parenthesis that no preliminary analysis was done.

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APPENDIX D

MATERIAL AND ENERGY BALANCES, AND THERMAL EFFICIENCIES FOR GASIFIERS

Table D-1

Gasifier Material Balance and Thermal Efficiencies
(BGC-Lurgi Slagger, Grand Forks Slagger, Dry Ash Lurgi, and Texaco Gasifiers)

Gasifier	BGC-Lurgi Slagger	BGC-Lurgi Slagger	BGC - Lurgi Slagger	Grand Forks Slagger	Dry Ash Lurgi	Dry Ash Lurgi	Texaco Slurry Feed	Texaco Slurry Feed
Reference	C45	C51	C57	C53	C42	C1	C45	C58
Coal Type	Ill, #6	Frances	Pitts. #8	Lignite	Wyoming	Ill. #6	Ill. #6	Ill. #6
Gasifier Exit								
Temp., °F	820	960	945	(f)	900	1078	2360	2300
Pressure, psia	300	365	365	415	430	315	600	815
Raw Gas Comp. (Mole %)								
H_2	28.56	25.51	26.03	27.7[d]	21.88	20.92	28.83	25.54
CO	54.51	50.57	49.96	55.4	10.59	7.49	42.45	29.46
CO_2	1.82	2.42	3.07	8.5	16.65	15.29	8.71	13.59
CH_4	7.29	6.04	6.47	7.2	6.35	4.28	0.08	0.27
H_2O	4.67	10.62	8.95	-	43.67	50.50	17.88	9.56
N_2	0.42	4.10	3.97	-	0.15	0.18	0.78	0.45
H_2S	1.29	0.09	0.48	0.6	0.08	0.59	1.01	1.07
COS	0.06	0.01	0.02	-	(<0.01)	0.03	0.06	0.06
NH_3	0.85	0.21	0.29	-	0.28	0.40	0.20	-
C_2H_6	0.53	0.43	0.76	0.4	0.35	0.32	-	-
	100.00	100.00	100.00	99.8	100.00	100.00	100.00	100.00
scf[a]/lb coal[b]	34.3	33.9	33.7	26.8	23.8	26.5	34.3	30.5
LHV/scf[a]	359	337	345	355	400	378	299	300
H_2/CO	0.524	0.504	0.521	0.500	2.07	2.79	0.679	0.867
Steam/ Oxygen (Mole)	1.15	1.33	1.25	1.01	7.47	8.45	-	-
Steam/Coal[b] (wt)	0.346	0.400	0.407	0.260	1.49	2.58	-	-
Oxygen/Coal[b] (wt)	0.534	0.535	0.579	0.458	0.354	0.541	0.953	1.03
$4\ CH_4/(H_2 + CO)$	0.351	0.318	0.340	0.347	0.782	0.603	0.0045	0.020
Oxygen/Syngas[c] (scf)	0.157	0.168	0.179	0.189	0.130	0.188	0.332	0.397
lb Steam/mscf Syngas	8.57	10.6	10.6	9.09	46.2	75.5	-	-
Thermal Efficiencies								
Cold Gas	90.1	89.4	88.3	(e)	80.0	76.0	75.3	68.2
Clean Fuel	90.1	92.5	88.8	(e)	87.6	81.7	75.3	68.2
Net	77.6	77.8	74.6	(e)	70.6	58.8	71.8	67.8

(a) Clean, dry gas.
(b) Dry and ash-free coal.
(c) Syngas = $H_2 + CO + 3\ CH_4 + 5.2\ C_2H_6$.
(d) Dry basis.
(e) Not calculated due to poor material and energy balances.
(f) Not given.

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TABLE D-2

Gasifier Material Balance and Thermal Efficiencies
(Synthane, Winkler, Koppers-Totzek, EPRI Agglomerating, and Hygas Gasifiers)

Gasifier	Synthane Pilot Plant	Synthane PDU	Synthane PDU	Winkler	Koppers-Totzek	EPRI Agglomerating	IGT Hygas	IGT Hygas
Reference	C44	C56	C56	C52	C26	C1	C54	C55
Coal Type	Rosebud	Rosebud	Rosebud	German Brown	TVA	Ill. #6	Ill. #6	Rosebud
Gasifier Exit								
Temp., °F	1390	1540	1480	1300	2730	1550	640	(f)
Pressure, psia	615	300	600	30	20	340	1035	1000
Raw Gas Comp. (Mole %)								
H_2	16.17	38.74[d]	35.76[d]	25.61	29.35	30.79	11.15	26.81[d]
CO	5.41	24.96	17.43	23.56	45.66	36.27	3.88	6.47
CO_2	17.75	26.15	32.12	14.61	8.55	13.11	13.02	35.61
CH_4	4.78	8.89	13.09	1.20	-	5.88	7.74	22.98
H_2O	55.34	-	-	33.28	14.16	12.06	59.85	-
N_2	0.12	-	-	0.73	1.01	0.71	3.22	6.47
H_2S	0.15	0.26	0.31	0.91	1.18	1.11	0.47	0.04
COS	-	-	-	0.10	0.09	0.04	-	-
NH_3	0.22	-	-	-	-	0.03	0.43	-
C_2H_6	0.06	1.00	1.29	-	-	-	0.24	1.62
	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00
scf[a]/lb Coal[b]	19.4	19.9	15.2	25.3	33.2	32.2	15.0	11.1
LHV/scf[a]	400	387	440	307	299	345	447	512
H_2/CO	2.99	1.55	2.05	1.09	0.638	0.849	2.87	4.14
Steam/Oxygen (Mole)	8.00	5.45	6.75	1.52	0.783	1.48	14.4	12.8
Steam/Coal[b] (wt)	2.20	1.05	1.06	0.645	0.415	0.586	1.98	1.194
Oxygen/Coal[b] (wt)	0.488	0.341	0.278	0.755	0.942	0.705	0.244	0.166
4 $CH_4/(H_2 + CO)$	0.886	0.558	0.98	0.098	-	0.351	2.06	2.76
Oxygen/Syngas[c] (scf)	0.218	0.156	0.146	0.343	0.341	0.226	0.128	0.103
lb Steam/mscf Syngas[c]	82.0	40.4	46.8	24.7	12.7	15.8	87.7	62.4
Thermal Efficiencies								
Cold Gas	78.3	(e)	(e)	74.7	71.3	80.9	(e)	(e)
Clean Fuel	78.3	(e)	(e)	74.7	71.3	80.9	(e)	(e)
Net	60.9	(e)	(e)	56.8	58.0	75.2	(e)	(e)

(a) Clean, dry gas.
(b) Dry and ash-free coal.
(c) Syngas = $H_2 + CO + 3\ CH_4 + 5.2\ C_2H_6$.
(d) Dry basis.
(e) Not calculated due to poor material and energy balances.
(f) Not given.

TABLE D-3
Gasifier Material Balance and Thermal Efficiencies
(Dry Ash Lurgi, Synthane, IGT Hygas, and Combustion Engineering Gasifiers)

Gasifier	Dry Ash Lurgi	Dry Ash Lurgi	Synthane	Synthane	Synthane	IGT Hygas	Combustion Engineering
Reference	C17	C26	C5	C5	C21	C38	C1
Coal Type	Ill.#6	TVA	Wyodak	Pittsburgh	Western	Montana	Ill.#6
Gasifier Exit							
Temp, °F	1000	1100	750	750	1697	·600	1700
Pressure, psia	315	315	1000	1000	1013	1170	~15
Raw Gas Comp (Mole%)							
H_2	23.16	23.58	14.20	17.37	16.02	15.23	29.99
CO	11.53	12.24	14.80	10.42	8.86	12.95	63.29
CO_2	18.34	16.98	22.84	18.07	28.39	12.00	2.44
CH_4	8.64	6.77	13.50	15.29	9.91	8.28	0.05
H_2O	35.85	39.10	32.17	36.73	34.83	50.53	1.76
N_2	0.25	0.52	0.50	0.50	0.14	0.04	1.12
H_2S	0.85	0.81	0.30	0.40	0.21	0.12	1.26
COS	0.06	-	-	-	(<0.01)	-	0.09
NH_3	0.57	-	0.70	0.72	0.42	0.22	-
C_2H_6	0.75	-	0.99	0.50	1.22	0.63	-
	100.00	100.00	100.00	100.00	100.00	100.00	100.00
scf[a]/lb coal[b]	23.4	27.2	13.5	17.0	12.3	23.6	34.8
LHV/scf[a]	431	384	512	518	504	455	303
H_2/CO	2.01	1.93	0.960	1.67	1.81	1.18	0.474
Steam/Oxygen (Mole)	6.94	6.10	3.67	6.96	3.79	7.34	-
Steam/Coal[b] (wt)	1.51	1.77	0.699	1.09	0.750	1.05	-
Oxygen/Coal[b] (Wt)	0.386	0.515	0.339	0.279	0.351	0.253	0.932
$4CH_4/(H_2+CO)$	0.996	0.756	1.86	2.20	1.59	1.17	0.002
Oxygen/Syngas[c] (scf)	0.134	0.172	0.175	0.112	0.201	0.084	0.313
lb steam/mscf Syngas[c]	44.2	49.8	30.5	37.1	36.3	29.3	-
Thermal Efficiencies							
Cold Gas	78.4	77.1	84.5	86.6	79.5	88.3	78.1
Clean Fuel	87.0	82.1	84.5	86.6	79.5	92.6	78.1
Net	67.5	61.4	73.4	76.4	64.6	77.8	72.5

(a) Clean, dry gas.
(b) Dry and ash-free coal.
(c) Syngas = $H_2 + CO + 3\ CH_4 + 5.2\ C_2H_6$.

TABLE D-4

Gasifier Material Balance and Thermal Efficiencies
(Medium and High Pressure Bi-Gas Gasifiers)

Gasifier	M.P. Bi-Gas	M.P. Bi-Gas	M.P. Bi-Gas	M.P. Bi-Gas	H.P. Bi-Gas	H.P. Bi-Gas	H.P. Bi-Gas	H.P. Bi-Gas
Reference	C2	C17	C25	C45	C3	C3	C21	C23
Coal Type	Ill.#6	Ill.#6	Eastern	Ill.#6	Montana	W.Ky.	Western	Eastern
Gasifier Exit								
Temp, °F	1700	1700	1700	1700	1500	1600	1600	1700
Pressure, psia	360	440	485	373	1215	1175	1230	1000
Raw Gas Comp (Mole%)								
H_2	30.10	29.78	34.85	29.50	20.51	19.75	21.99	25.27
CO	35.04	43.16	40.86	35.33	26.42	35.65	16.23	28.61
CO_2	12.32	8.31	9.57	12.51	16.06	11.35	19.32	14.74
CH_4	6.01	8.23	1.78	6.04	10.95	12.60	9.86	10.85
H_2O	14.39	8.42	11.29	14.40	25.52	18.90	31.96	18.70
N_2	0.44	0.73	0.70	0.45	0.35	0.60	0.13	0.50
H_2S	0.97	1.17	0.92	1.03	0.19	1.15	0.16	1.11
COS	0.10	0.07	0.02	0.10	-	-	0.01	-
NH_3	0.63	0.13	0.01	0.64	-	-	0.34	0.22
C_2H_6	-	-	-	-	-	-	-	-
	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00
scf[(a)]/lb coal[(b)]	32.2	33.4	36.7	32.0	25.3	29.1	24.1	29.7
LHV/scf[(a)]	349	360	311	350	413	413	419	398
H_2/CO	0.859	0.690	0.853	0.835	0.776	0.554	1.35	0.883
Steam/Oxygen (Mole)	1.85	1.58	1.59	1.82	2.52	1.46	4.42	3.21
Steam/Coal[(b)] (Wt)	0.693	0.507	0.635	0.693	0.697	0.473	1.14	0.933
Oxygen/Coal[(b)] (Wt)	0.665	0.569	0.708	0.677	0.492	0.576	0.458	0.516
$4CH_4/(H_2+CO)$	0.369	0.451	0.094	0.373	0.933	0.910	1.03	0.806
Oxygen/Syngas[(c)] (scf)	0.199	0.163	0.221	0.206	0.165	0.163	0.160	0.156
lb steam/mscf Syngas[(c)]	17.5	12.2	16.7	17.7	19.7	11.3	33.6	23.7
Thermal Efficiencies								
Cold Gas	82.8	85.6	84.3	81.6	89.3	86.0	87.3	85.8
Clean Fuel	82.8	85.6	84.3	81.6	89.3	86.0	87.3	85.8
Net	73.4	79.7	75.6	73.7	78.0	79.6	69.2	77.0

(a) Clean, dry gas.
(b) Dry and ash-free coal.
(c) Syngas = $H_2 + CO + 3\ CH_4 + 5.2\ C_2H_6$.

Table D-5

Gasifier Energy Balance
(BGC-Lurgi Slagger, Grand Forks Slagger, Dry Ash Lurgi, and Texaco Gasifiers)

Gasifier	BGC-Lurgi Slagger	BGC-Lurgi Slagger	BGC-Lurgi Slagger	Grand Forks Slagger	Dry Ash Lurgi	Dry Ash Lurgi	Texaco Slurry Feed	Texaco Slurry Feed
Reference	C45	C51	C57	C53	C42	C1	C45	C58
Coal Type	Ill. #6	Frances	Pitts.#8	Lignite	Wyoming	Ill. #6	Ill. #6	Ill. #6
Product Heating Value (% LHV Net Coal)								
H_2 + CO	69.6	71.1	68.3	(b)	47.8	49.3	75.0	67.2
CH_4	18.2	16.3	16.5		29.4	23.6	0.3	1.0
C_2	2.3	2.0	3.5		2.8	3.1	-	-
Naphtha/Oil	-	3.1	0.5		7.6	5.7	-	-
Recoverable Heat	-	-	-		4.3	5.3	14.4	18.7
Steam	-1.6	-2.8	-1.6		-13.4	-16.4	-	-
Impurities	3.0	0.4	1.2		0.7	3.1	2.4	2.7
Waste Heat	7.8	9.4	11.1		19.4	25.3	7.4	9.9
Slag/Ash/Dust	0.2	-	-		0.9	0.5	-	-
Heat Loss[(a)]	0.5	0.5	0.5		0.5	0.5	0.5	0.5
	100.0	100.0	100.0		100.0	100.0	100.0	100.0
Other Energy Consmption (% LHV Net Coal)								
Air Separation	7.9	8.4	8.9	(b)	6.0	8.3	14.1	15.4
Work	3.0	3.5	3.7		1.9	3.5	3.8	3.7

(a) Heat loss is prmalized to 0.5% for all cases.
(b) Not calculate due to poor or incomplete material and energy balances.

TABLE D-6

Gasifier Energy Balance
(Synthane, Winkler, Koppers-Totzek, EPRI Agglomerating, and Hygas Gasifiers)

Gasifier	Synthane Pilot Plant	Synthane PDU	Synthane PDU	Winkler	Koppers-Totzek	EPRI Agglomerating	IGT Hygas	IGT Hygas
Reference	C44	C56	C56	C52	C26	C1	C54	C55
Coal Type	Rosebud	Rosebud	Rosebud	German Brown	TVA	Ill. #6	Ill. #6	Ill. #6
Product Heating Value (% LHV Net Coal)								
H_2 + CO	45.5	(b)	(b)	69.5	71.3	63.9	(b)	(b)
CH_4	32.1			5.2	-	17.0		
C_2	0.7			-	-	-		
Naphtha/Oil	-			-	-	-		
Recoverable Heat	22.1			8.7	12.9	13.7		
Steam	-26.3			1.0	-0.3	-5.0		
Impurities	1.2			2.8	2.4	2.2		
Waste Heat	24.2			12.3	9.1	6.9		
Slag/Ash/Dust	-			-	4.1	0.8		
Heat Loss[(a)]	0.5			0.5	0.5	0.5		
	100.0			100.0	100.0	100.0		
Other Energy Consumption (% LHV Net Coal)								
Air Separation	9.9	(b)	(b)	14.7	13.7	10.3	(b)	(b)
Work	3.3			12.9	12.2	4.1		

(a) Heat loss is normalized to 0.5% for all cases.
(b) Not calculated due to poor or incomplete material and energy balances.

TABLE D-7
Gasifier Energy Balance
(Dry Ash Lurgi, IGT Hygas, and Combustion Engineering Gasifiers)

Gasifier	Dry Ash Lurgi	Dry Ash Lurgi	Synthane	Synthane	Synthane	IGT Hygas	Combustion Engineering
Reference	C17	C26	C5	C5	C21	C38	C1
Coal Type	Ill.#6	TVA	Wyodak	Pittsburgh	Western	Montana	Ill.#6
Product Heating Value, (% LHV Net Coal)							
H_2+CO	41.2	48.4	32.4	30.8	31.9	43.6	78.0
CH_4	32.2	28.7	46.1	52.8	39.4	39.4	0.1
C_2	5.0	-	6.0	3.0	8.2	5.3	-
Naphtha/Oil	8.6	5.0	-	-	-	4.3	-
Recoverable Heat	-	-	4.8	5.1	11.3	2.2	3.6
Steam	-10.7	-9.7	-6.0	-9.3	-13.8	-13.3	17.3
Impurities	3.1	2.2	1.6	1.9	1.2	0.8	2.2
Waste Heat	19.4	23.1	14.6	15.2	21.3	16.8	-2.0
Slag/Ash/Dust	0.7	1.8	-	-	-	0.4	0.3
Heat Loss(a)	0.5	0.5	0.5	0.5	0.5	0.5	0.5
	100.0	100.0	100.0	100.0	100.0	100.0	100.0
Other Energy Consumption (% LHV Net Coal)							
Air Separation	6.0	7.7	8.4	5.5	9.2	4.2	13.9
Work	2.8	3.3	1.5	0.5	3.2	-0.5	12.6

(a) Heat loss is normalized to 0.5% for all cases.

TABLE D-8
Gasifier Energy Balance
(Medium and High Pressure Bi-Gas Gasifiers)

GASIFIER	M. P. Bi-Gas	M. P. Bi-Gas	M. P. Bi-Gas	M. P. Bi-Gas	H. P. Bi-Gas	H. P. Bi-Gas	H. P. Bi-Gas	H. P. Bi-Gas
Reference	C2	C17	C25	C45	C3	C3	C21	C23
Coal Type	Ill.#6	Ill.#6	Eastern	Ill.#6	Montana	W.Ky.	Western	Eastern
Product Heating Value, (% LHV Net Coal)								
H_2+CO	64.7	63.9	78.7	63.6	52.3	51.2	48.5	53.2
CH_4	18.1	21.7	5.6	18.0	37.0	34.8	38.8	32.6
C_2	-	-	-	-	-	-	-	-
Naphtha/Oil	-	-	-	-	-	-	-	-
Recoverable Heat	10.6	10.1	10.9	10.9	8.2	7.9	6.9	9.7
Steam	-5.8	-4.9	-5.9	-4.8	-9.5	-3.6	-16.3	-9.4
Impurities	3.0	2.3	1.9	3.0	0.4	2.1	1.0	2.4
Waste Heat	8.3	6.4	7.5	8.2	11.1	7.1	20.6	10.8
Slag/Ash/Dust	0.6	-	0.8	0.6	-	-	--	0.2
Heat Loss(a)	0.5	0.5	0.5	0.5	0.5	0.5	0.5	0.5
	100.0	100.0	100.0	100.0	100.0	100.0	100.0	100.0
Other Energy Consumption (% LHV Net Coal)								
Air Separation	9.9	8.1	10.6	10.0	8.5	8.3	8.0	7.6
Work	4.3	3.0	3.1	4.0	1.5	2.4	0.7	1.5

(a) Heat loss is normalized to 0.5% for all cases.

APPENDIX E

INTERVIEWS WITH PROCESS LICENSERS, CONTRACTORS
AND PILOT PLANT OPERATORS TO OBTAIN GASIFIER INFORMATION

I. Visit to Electric Power Research Institute

Dr. C. D. Prater and Professor R. Shinnar visited EPRI (Electric Power Research Institute) at Palo Alto, California on January 23 and 24, 1978. Their discussion with EPRI personnel, Drs. A. Gluckman and Nolt and others, contains the following highlights:

1. They supplied us a recent report comparing a Modified Texaco Gasifier to a BGC-Lurgi Slagger and a Combustion Engineering Gasifier for the purpose of combined cycle power generation.

2. The heat recovery from the slag containing effluent gas of the Modified Texaco Gasifier has been demonstrated.

II. Visit to Synthane PDU and Pilot Plant at Pittsburgh Energy Research Center

Professor R. Shinnar and Dr. J. C. W. Kuo visited PERC (Pittsburgh Energy Research Center) at Bruceton, Pennsylvania on February 2, 1978. Their discussion with PERC personnel, W. P. Haynes and J. P. Starkey, on the subject of the Synthane Gasifier covered the following highlights:

1. They supplied us two recent reports on Synthane PDU data, one report on Synthane Process assessment based on a set of data from the Synthane Pilot Plant, a draft report on the computer simulation of Synthane PDU data, and some additional PDU test data.

2. The Synthane Gasifier has the advantages of stable operation, no tar and low phenol formation. It, however, has the disadvantages of low conversion, low gas linear velocity, high steam and oxygen consumption, and the requirement for pretreating the caking coals.

3. The Synthane Pilot Plant is designed for 600-1000 psia operation; while 300-400 psia is probably more optimal for Mobil's processes. PDU data at 300 psia show higher conversion, higher H_2 + CO selectivity, but lower throughput and higher tar yield. To run the pilot plant at 300-400 psia would probably require modifications.

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4. One of the major problems in the Synthane Pilot Plant operation is clinker formation. To prevent its formation, more steam and oxygen than those used in the PDU must be used. The clinker formation may be due to poor gas feed distributor design, low gas linear velocity, and local overheating.

They also toured the Synthane PDU and Pilot Plant, and discussed the pilot plant operation with plant manager, R. Lewis.

APPENDIX F

ESTIMATION OF PRODUCT COST BASED ON DIRECT INVESTMENT COST

In the text we assumed that we can estimate the product cost based on utility financing by using a single factor that multiplies the direct investment per unit produced. This gives an estimate for all the costs associated with production, such as depreciation of investment, interest during construction, maintenance and insurance, operating cost of plant (exclusive of raw materials), repayment of loans, return on investment and taxes. This factor changes with the method of financing and the return of investment required. It should also change with the type of process, as different processes have different requirements for maintenance and operating personnel. While the estimates made by large engineering companies claim to have done that the end results are remarkably similar for different processes. We give in Table F-1 the results of different EPRI and DOE studies using utility financing.

In the following table the cost of raw materials per unit is defined by taking the total raw material cost subtracting from it the value of by-products and dividing it by the total number of units produced.

The direct investment cost used here is obtained by taking the total direct plant cost, including engineering and contractor's fee but exclusive of interest during construction and working capital and dividing it by the total number of units produced over 20 years of plant life.

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Table F-1

Estimates of Total Cost
Based on Direct Investment Cost

Case	Unit Used	Direct Investment Per Unit	Total Cost Per Unit Excluding Coal*	Col.4 / Col.3
Fuel Gas from BGC-Lurgi Slagger (Reference 21)	1 MMBTU	0.234	1.31	5.6
Combined Cycle Power Plant Based on Texaco Gasifier (Reference 19)	KWH	0.052	0.28	5.4
Fuel Gas from Dry Ash Lurgi Gasifier (Reference 12)	–	0.54	3.17	5.8
Coal Fired Boiler with Stack Gas Scrubber (Reference 7)	KWH	0.51	0.28	5.5

* By-product value subtracted from coal cost.

*U.S. GOVERNMENT PRINTING OFFICE : 1979 0-620-097/2202

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