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A STUDY OF THE CONVERSION OF COAL TO HYDROGEN, METHANE, AND LIQUID FUELS FOR AIRCRAFT

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ABSTRACT

A STUDY OF THE CONVERSION OF COAL TO HYDROGEN, METHANE, AND LIQUID FUELS FOR AIRCRAFT

Depletion of domestic crude oil resources has led to the consideration of alternate fuels derived from more plentiful coal resources. This study addresses the conversion of coal to hydrogen, methane, and synthetic aviation kerosene, for use as alternate aviation fuels. Thermal efficiencies are derived for producing a) gaseous hydrogen via the Koppers-Totzek, U-GAS®, and the continuous Steam-Iron Processes; b) gaseous methane via the HYGAS® and CO_2 Acceptor Processes; c) synthetic aviation kerosene via hydrocracking and aromatic hydrogenation of the heavy fuel oil produced by Consol Synthetic Fuel Process; d) gasoline via hydrocracking the heavy fuel oil produced by the Consol Synthetic Fuel Process. (The gasoline product would supplement the gasoline produced by a conventional crude oil refinery, thus reducing the amount of light distillate normally catalytically reformed, and thereby freeing a greater amount of light distillate for the manufacture of aviation kerosene.) Process economics are analyzed for all processes except the Koppers-Totzek and CO_2 Acceptor Processes.

EXECUTIVE SUMMARY

The shortage of U.S. supplies of petroleum and natural gas has greatly stimulated research and development work on the conversion of coal to fluid fuels. Reduced supplies of petroleum will limit the supply of jet fuel from this source. The possibility of using liquefied hydrogen or methane as alternatives to standard petroleum-derived jet fuel has led to the study of the conversion of coal to these fuels and to other liquid fuels.

IGT's task was to study the conversion of coal to gaseous hydrogen or methane at 6996 kN/m^2 (1000 psig) and to kerosene or to wide-cut jet fuel.

Product Quality

Product quality is governed partly by the requirements for liquefaction.

1. Hydrogen — A minimum purity of 92% is necessary. The remaining 8% (or less) can be N_2, Argon, CO, CH_4, and CO_2. It is preferrable to reduce the CO_2 level to 0.1% or less because, while the cost of removing the first four before liquefaction mainly involves extracting the last few ppm, the cost of removing CO_2 is more dependent on the total amount removed.

2. Methane — A typical pipeline gas from coal would be satisfactory, i.e., 92% to 94% CH_4, and the rest hydrogen, N_2, Ar, CO, and CO_2, with the two carbon oxides at the 0.1% level.

3. Kerosene or Wide-Cut Jet Fuel — Although jet fuel would be the final desired product, the proposal for this study did not address the inclusion of facilities to produce a specification product and no specifications were proposed. The project plan calls for an estimate of the energy requirements (overall efficiency) for converting coal to syncrude and upgrading the crude to a kerosene-type product. We have considered petroleum refinery light distillate and wide-cut kerosene as suitable products for this part of the study.

Plant Capacity

Studies of liquid hydrogen supply for airports have indicated fuel requirements (HHV) at the 316 or 369 GJ/d (300 or 350 billion Btu/d) level. These quantities are at the general level of output of the "standard" pipeline

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gas from coal plant — 264 GJ/d (250 billion Btu/d). Outputs for the two liquid products are at the 211 and 316 GJ/d (200 and 300 billion Btu/d) levels. These capacities are beyond the point where economies of scale are significant.

Type of Coal

The scope of this work did not include the effect of different coals on a given process; also, neither a single plant location nor the coal supply was specified. For purposes of comparison, it is desirable to keep coal variations to a minimum. However, the project plan allows the selection of coals that show a process to the best advantage. For the manufacture of hydrogen and methane, Montana subbituminous coal was selected because of 1) its greater reactivity and nonagglomerating characteristics during gasification, 2) its abundant low-sulfur rate, and 3) its lower cost relative to Eastern coals. For coal liquefaction, as discussed in another section of this report, Eastern coal was selected because of its apparent process advantages over Western coals. Table ES-1 gives the properties of these two coals. Two major areas are covered in this study: analyses of process efficiencies and process economics.

Processes Evaluated

Hydrogen

1. Koppers-Totzek gasifier
2. U-GAS® gasifier
3. Continuous Steam-Iron Process

Methane

1. HYGAS® Process
2. CO_2 Acceptor Process

Kerosene or Jet-Fuel Component

The Consol Synthetic Fuel (CSF) Process was used as the basic liquefaction process. The products from this process were upgraded by two alternatives:

1. Increased petroleum refinery output by importing coal-derived gasoline
2. Hydrocracking and saturation of aromatics.

Table ES-1. ANALYSES OF COALS

	Montana Subbituminous	Pittsburgh Seam Bituminous
	wt %	wt %
Proximate Analysis		
Moisture	22.0	14.4
Volatile Matter	29.4	33.4
Fixed Carbon	42.6	40.9
Ash	6.0	11.3
Total	100.0	100.0
Ultimate Analysis (Dry)		
Carbon	67.70	68.97
Hydrogen	4.61	4.90
Nitrogen	0.85	1.28
Oxygen	18.46	7.34
Sulfur	0.66	4.29
Ash	7.72	13.22
Total	100.00	100.00
Dry HHV, kJ/kg (Btu/lb)	26,251 (11,290)	29,390 (12,640)

Overall efficiencies were determined for all seven of these processes from process designs based on coal as the primary energy source. In order to have a reasonable measure of the overall efficiency, it was necessary to go into considerable detail to develop material and energy balances and to determine utility requirements.

The original project plan specified that process economics be determined for one process in each of the product categories. The U-GAS and HYGAS Processes were selected for hydrogen and methane, respectively. Later the sponsor requested that a cost estimate for hydrogen by the Steam-Iron Process be included in the economics. For the jet-fuel component, we decided to estimate the economics for increased petroleum refinery output by importing coal-derived gasoline. For the second liquid process, the process information on hydrocracking and aromatics hydrogenation provided by Universal Oil Products Co. also included some costs, so the economics for this case are also presented.

Process Descriptions

For this summary, we have presented simple block flow diagrams for each of the processes. More detailed flow diagrams are presented in the main body of this report.

HYDROGEN FROM COAL

Hydrogen by Koppers-Totzek Gasification (Figure ES-1)

Information on raw material requirements and operating conditions for the Koppers-Totzek gasifier (Figure ES-1) was obtained from the Koppers Co. This is a commercially available process using suspension gasification of pulverized coal at essentially atmospheric pressure under slagging conditions.

Coal is first dried to 2% moisture and then pulverized to 70% through 200 mesh. Ground coal is discharged into a mixing nozzle where it is entrained in a stream of oxygen and steam and carried into the gasifier. About one-half of the coal ash drops out as slag into a quench tank below the gasifier. The rest is carried out as fly ash and is recovered by water scrubbing and electrostatic precipitation after waste-heat recovery from the hot gas.

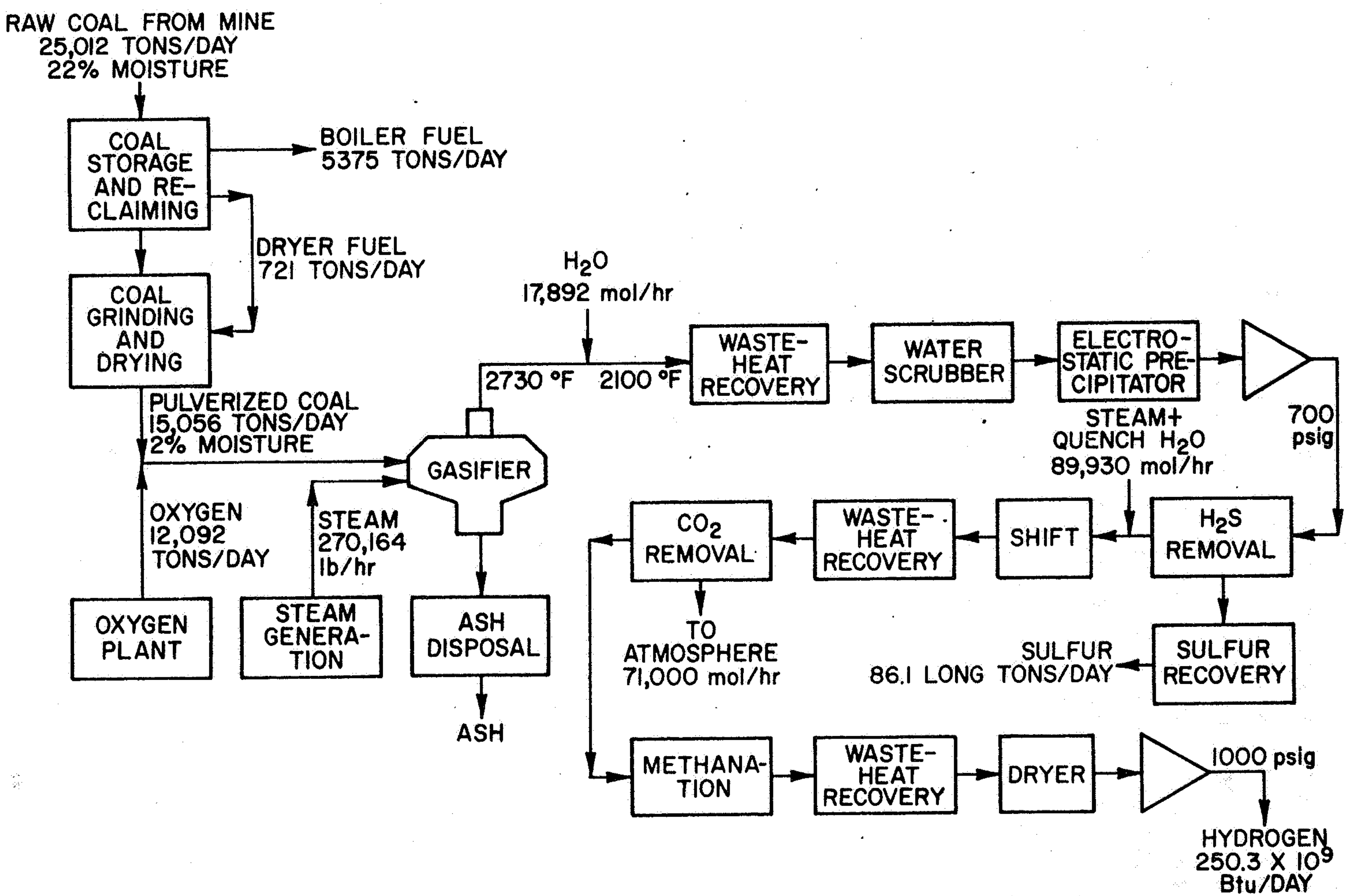

Figure ES-1. HYDROGEN BY KOPPERS-TOTZEK GASIFICATION

In order to upgrade the gas to process the hydrogen product, the synthesis gas is compressed to 4928 kN/m^2 (700 psig) and is carried through H_2S removal, CO shift, CO_2 removal, methanation, and final compression to 6996 kN/m^2 (1000 psig). The raw gas composition, on a dry basis, is 58.3% CO, 10.0% CO_2, 30.4% H_2, 1.0% N_2 and Ar, and 0.3% H_2S + COS. The CO must be converted by the well-known shift reaction, $CO + H_2O = CO_2 + H_2$. Because of the large amount of CO present and the need to convert nearly all of it, the process design is based on a low-temperature shift catalyst. This requires much less steam than a high-temperature shift catalyst because of the more favorable equilibrium, and results in a higher plant efficiency. The low-temperature shift catalyst cannot be used with sulfur-containing gas. However, since acid gases (CO_2 and H_2S) must be removed at some stage of the process, it is desirable to remove H_2S before and CO_2 after CO conversion.

A two-stage Rectisol system similar to units that have been successfully designed for commercial operation is used. Because this is a physical absorption system, the synthesis gas is pressurized to 4928 kN/m^2 (700 psig) before entering the H_2S scrubbing tower. After CO conversion, the CO_2 is removed by the second stage of the Rectisol unit.

Since the product specifications call for at least 92% H_2 (page v), a few percent CO can be left in the gas after the shift. The residual CO is reduced to 0.1% by cleanup methanation. It is more economical to methanate than to shift the CO, and the heat of methanation is recovered. A single-stage recycle quench methanation system is used.

Hydrogen by U-GAS Gasification (Figure ES-2)

The U-GAS Process (Figure ES-2) uses a single-stage fluidized-bed gasifier to generate synthesis gas. Operating conditions are based on information developed at IGT. Montana subbituminous coal, in contrast to Eastern caking coals, requires no pretreatment for this reactor.

The coal is first dried to 10% moisture and reduced to a size range of 100% below 8 mesh and a maximum of 15% below 100 mesh. Gasification occurs at 2411 kN/m^2 (335 psig), so some method must be used to feed the coal into the reactor. A lockhopper system is used in this design, because

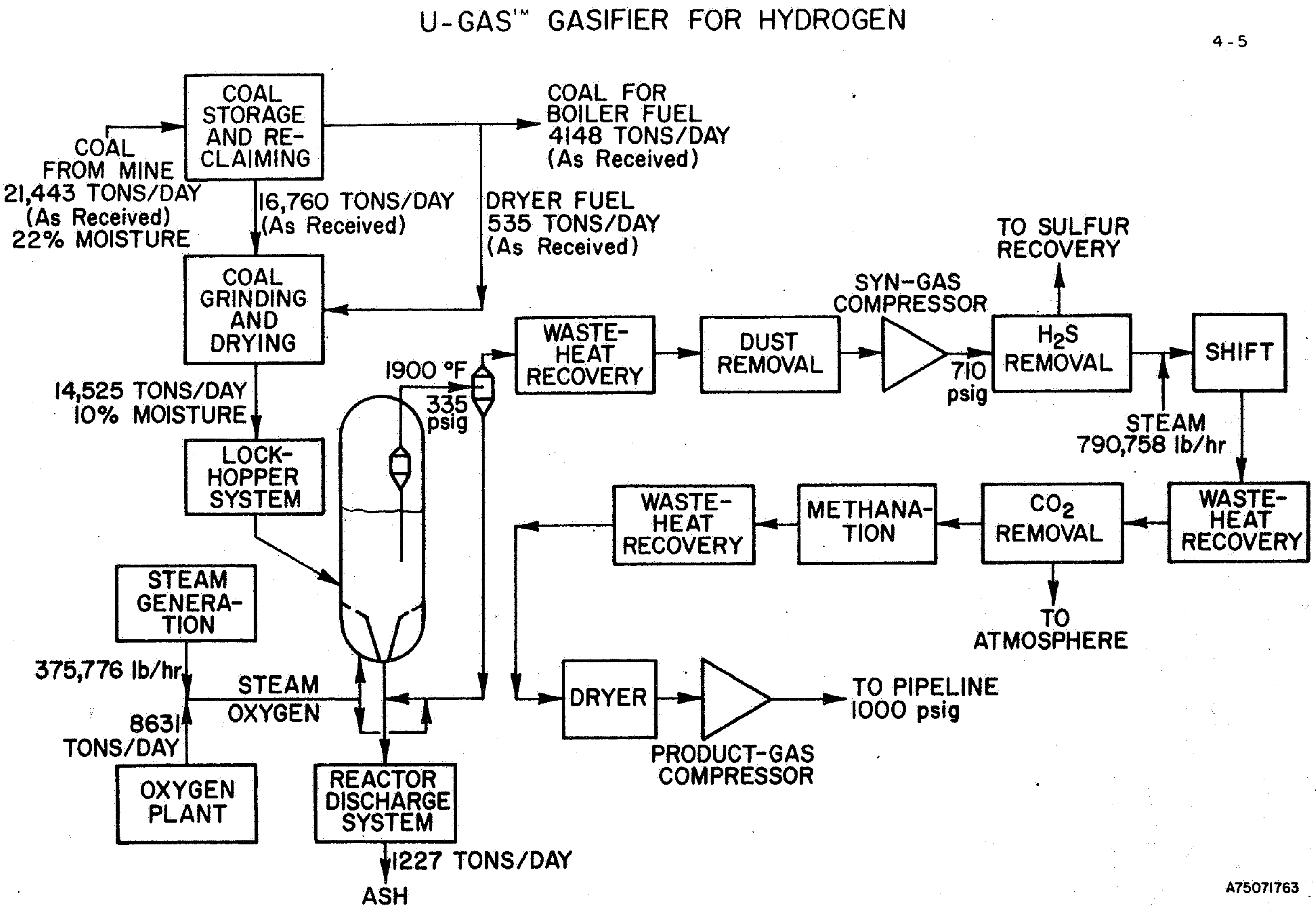

Figure ES-2. HYDROGEN BY U-GAS™ GASIFICATION this feed method has been used commercially at this pressure. The temperature in the fluidized-bed reactor is 1311 K (1900°F), so the operation is non-slagging. The removal of ash is accomplished by a method of agglomerating low-carbon-content ash particles for selective separation from the fluidized bed of coal. Fines elutriated from the bed are recovered in internal cyclones and also in an external set of cyclones through which the fines are collected and returned to the bed.

The raw gas is cooled to 422 K (300°F) in a waste-heat boiler and then passed through a venturi scrubber to remove fine particles carried over from the gasifier cyclones. The composition,on a dry basis, is 50.1% CO, 11.5% CO_2, 35.3% H_2, 0.7% N_2 + Ar, 2.1% CH_4, and 0.3% H_2S.

Following the gasifier, all steps in the U-GAS Process are analogous to those described above for the Koppers-Totzek Process. Two major advantages result from the U-GAS Process, which operates at lower temperature and elevated pressure:

1. The oxygen requirements are lowered from 10,970 to 7,830 t/d (12,092 to 8,631 short tons/d) which reduces the power requirement from 151,400 kW to 105,000 kW (203,000 to 141,000 hp).
2. The compression of synthesis gas from 2377 to 4997 kN/m^2 (330 to 710 psig), instead of from 142.7 to 4928 kN/m^2 (6 to 700 psig), reduces the power requirement from 202,800 to 32,800 kW (272,000 to 44,000 hp).

Hydrogen by the Steam-Iron Process (Figure ES-3)

Hydrogen by the Steam-Iron Process (Figure ES-3) is derived by the decomposition of steam by reaction with iron oxide, rather than synthesis-gas generated from coal. Coal is gasified to provide a producer gas for the regeneration of iron oxide. Because hydrogen is not derived from the producer gas, air can be used in the gasifier; nitrogen cannot contaminate the hydrogen because of the iron oxide barrier.

The iron oxide circulates between zones of oxidation and reduction. The following reactions are typical of those occurring in the steam-iron reactor section:

Reductor: $Fe_3O_4 + CO \rightarrow 3FeO + CO_2$

$Fe_3O_4 + H_2 \rightarrow 3FeO + H_2O$

Oxidizer: $3FeO + H_2O \rightarrow Fe_3O_4 + H_2$

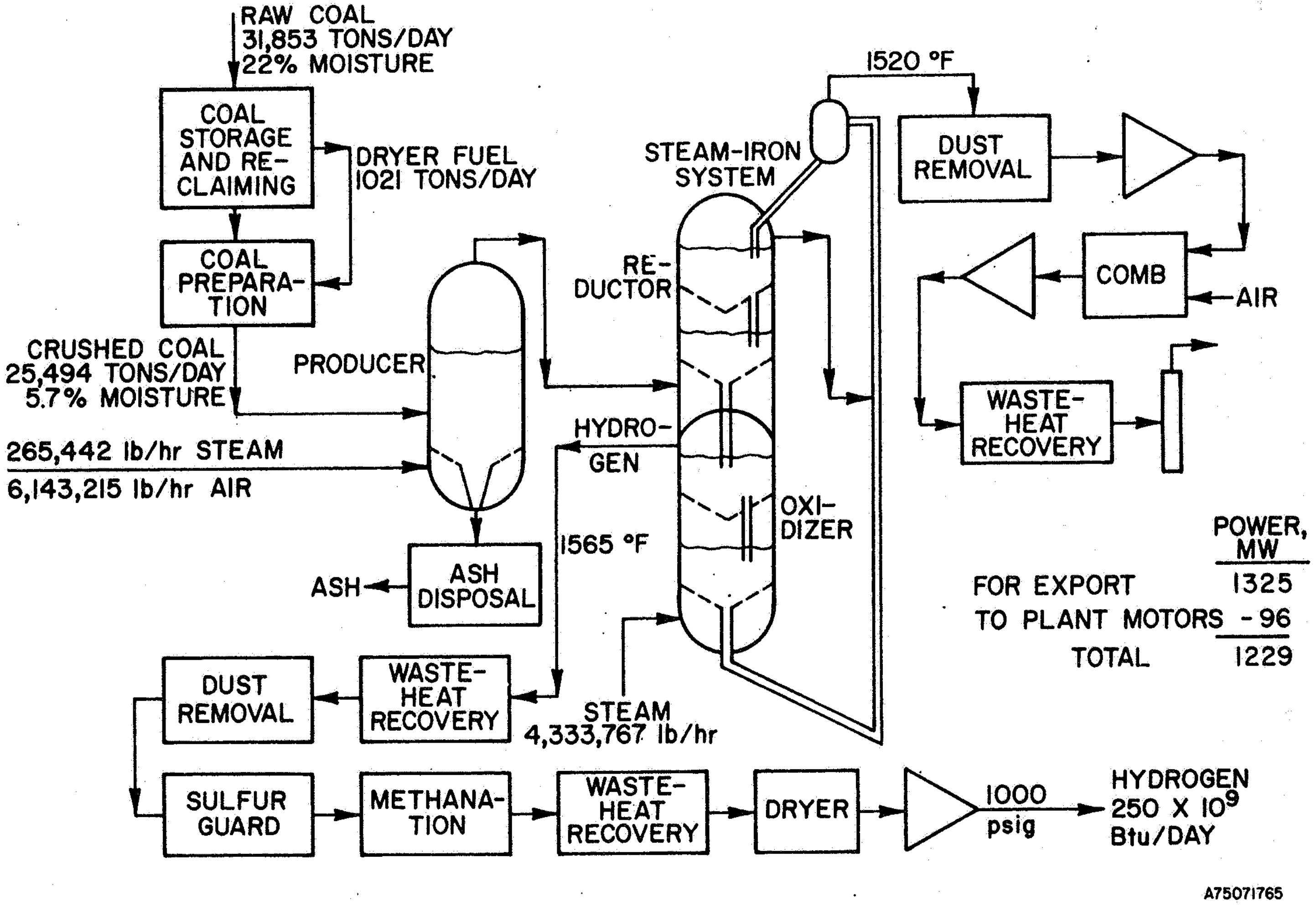

Figure ES-3. HYDROGEN BY THE FLUIDIZED STEAM-IRON PROCESS

The oxidizer effluent contains 37% hydrogen and 61% steam plus small amounts of nitrogen and carbon oxides. Condensation of the steam leaves a gas containing 95.9% H_2, 1.6% carbon oxides, and 2.5% N_2. No CO "shift" or acid-gas scrubbing is needed. A cleanup methanation step reduces carbon oxides to 0.2% followed by drying and compressing to 6996 kN/m^2 (1000 psig) to give product gas.

Power Generation From Reductor Off-Gas

The carbon monoxide and hydrogen in the producer gas are not completely converted in the reduction of iron oxide. Heating value plus sensible heat at 1100 K (1520°F) in the reductor off-gas represent 54% of the input coal fuel value. Part of this energy, 15% of the coal fuel value, is utilized in the plant to compress air and generate steam. A larger amount is used to generate electric power. After dust removal with cyclone separators and electrostatic precipitators, the effluent gas at 1100 K and 2517 kN/m^2 (1520°F and 365 psia) is used in a combined-cycle gas turbine-steam turbine system. We obtained the services of the United Technologies Research Center (UTRC) of United Technologies Corp. for estimating the amount and cost of this power recovery.

The gas is burned in a combustor with excess air at 2000 kN/m^2 (290 psia); 517 kN/m^2 (75 psi) control valve loss is assumed. The effluent from the combustor, at 1589 K (2400°F), is expanded to 114 kN/m^2 (16.5 psia) and 921 K (1198°F) in the gas turbine. A portion of the expansion power is used to drive the combustor and producer air compressors. The remaining shaft power is converted to 1083.8 MW of electric power in a generator. At present, the maximum temperature range is 1255 to 1366 K (1800° to 2000°F), but gas turbines with an inlet temperature of 1589 K (2400°F) are expected to be available by the time this process becomes commercial.

The expanded gas is used in a steam-power cycle to generate 241.6 MW of electric power from steam at 16,649 kN/m^2 and 811 K (2400 psig, 1000°F) generated in a waste-heat boiler, and also to generate a portion of the required process steam for the oxidizer. The cooled combustor gas leaves the steam cycle waste-heat boiler at 450 K (350°F). A total of 1325.4 MW of power is generated in addition to the shaft power used for air compression. Within the plant, 95.9 MW of this power is used for motor drives, etc., leaving 1229.5 MW of power as a by-product.

Product compositions for the three designs are given in Table ES-2.

Table ES-2. PRODUCT COMPOSITIONS FOR THREE HYDROGEN PROCESSES

	Process		
	Koppers-Totzek	U-GAS	Steam-Iron
Component	mol % (dry basis)		
CO	0.1	0.1	0.1
CO_2	<50 ppm	<50 ppm	0.1
H_2	93.1	94.3	95.7
CH_4	5.5	4.8	1.5
N_2 + Ar	1.3	0.8	2.6
Total	100.0	100.0	100.0

METHANE (PIPELINE GAS) FROM COAL

Pipeline Gas (Methane) by the HYGAS Process (Figure ES-4)

The gasifier design is based on operation at 8030 kN/m^2 (1150 psig). A water slurry feed system is used to feed coal into the hydrogasification reactor, based on successful pilot plant operation of the slurry feed. This vessel, at the top of the reactor, has a fluidized-bed dryer where the slurry water is evaporated by hot, gaseous effluent. The reactor coal feed passes through three zones of conversion:

1. A low-temperature (811 K, 1000 °F) transport reactor, where the coal is devolatilized and rapid-rate conversion to methane enriches the product gas.
2. The main fluidized bed at 1200 K (1700°F), where most of the methane is formed.
3. The steam-oxygen gasification zone at 1283 K (1850°F) , where synthesis gas is generated from the hydrogasifier char. Oxygen requirements are 2722 t/d (3000 short tons/d), much less than for hydrogen from coal.

The effluent gas is scrubbed to remove dust carry-over, then sent to a CO-conversion reactor where the H_2/CO ratio is raised from 1.18 to 3.3 in preparation for methanation. The catalyst is an oil- and sulfur-resistant,

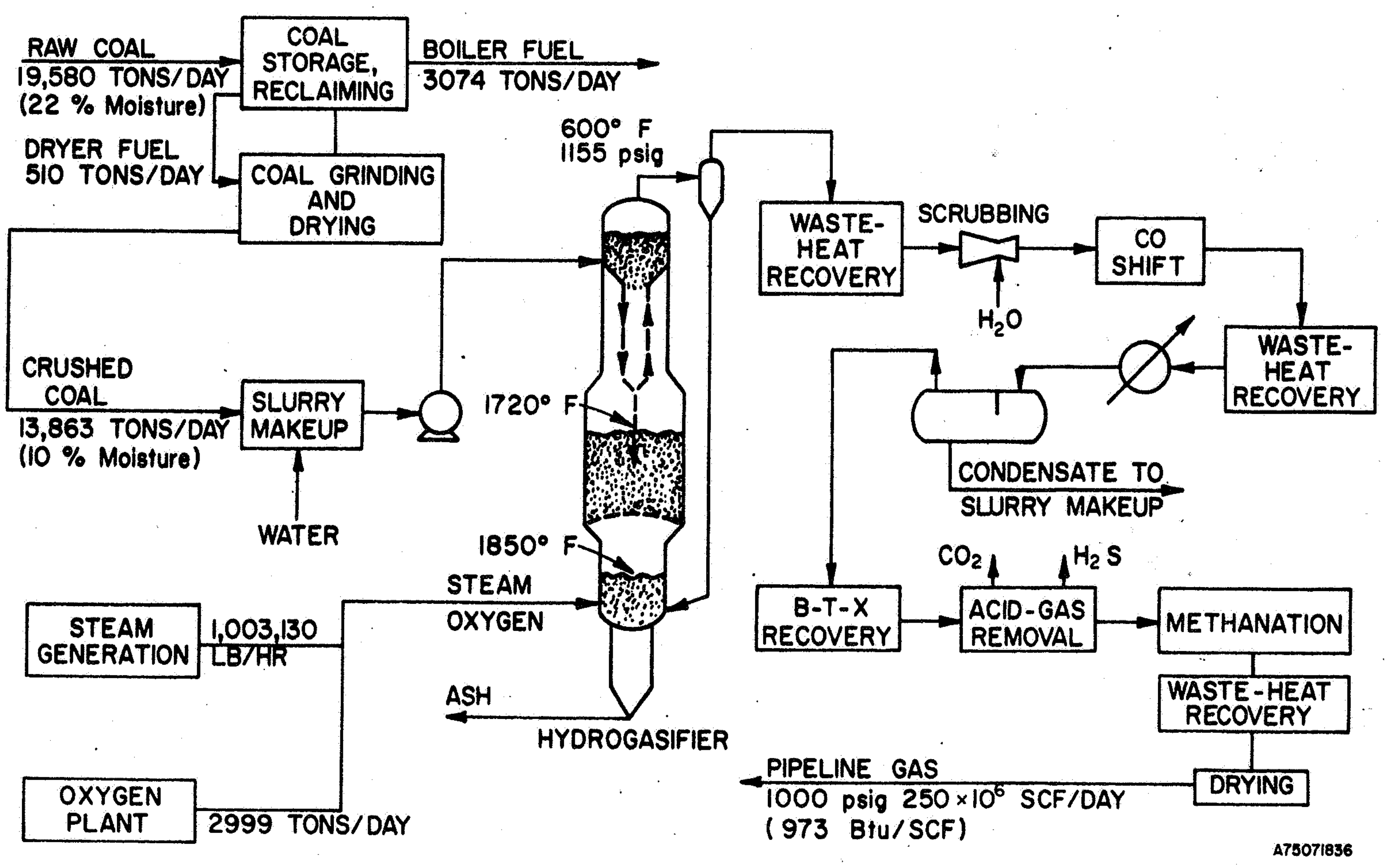

Figure ES-4. PIPELINE GAS (METHANE) BY HYGAS® PROCESS high-temperature shift catalyst. Because the dust scrubbing is carried out at a temperature high enough to prevent condensation of the vaporized slurry water, it is unnecessary to generate steam for the CO conversion reaction. This helps raise the overall efficiency.

The CO shift effluent is cooled to remove condensible oils and water. The remaining B-T-X* is also recovered for sale as a by-product. Most of the CO_2 and H_2S is removed by scrubbing with a hot carbonate solution. Final traces of H_2S are removed by dry-bed scrubbing prior to methanation.

The methanation step reacts CO and CO_2 with H_2 to make 49% of the total methane in the product gas. Because the reaction is highly exothermic, a four-stage recycle quench methanation system that limits the maximum catalyst-bed temperature to 755 K (900°F) is used. Product gas is available at 6996 kN/m^2 (1000 psig).

Pipeline Gas (Methane) by the CO_2 Acceptor Process (Figure ES-5)

In this process, coal is gasified in a fluidized bed with steam to generate methane and synthesis gas (Figure ES-5). Heat is supplied by the reaction of CO_2 with the acceptor (MgO-CaO), which is showered through the bed of char and circulates between the gasifier and the regenerator vessels. In the regenerator vessels, the acceptor is regenerated by burning residual gasifier char with air to decompose the $CaCO_3$. No oxygen plant is needed, and the gasification system operates at 1115 kN/m^2 (147 psig).

Gasifier effluent is scrubbed to remove dust and H_2S prior to methanation. No CO shift is necessary because the H_2/CO ratio is more than sufficient for methanation. CO_2 is removed after methanation because of the necessity of reacting it with the excess hydrogen, which would otherwise result in a reduced heating value. Final CO_2 removal occurs after methanation.

The gasifier and regenerator operate at a much lower pressure, 1115 kN/m^2 (147 psig) than in the HYGAS Process. The pressure is raised to an intermediate level, 3100 kN/m^2 (435 psig), prior to H_2S removal and methanation, and then to the final pipeline pressure after the final CO_2 removal.

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* B-T-X = Benzene-Toluene-Xylene.

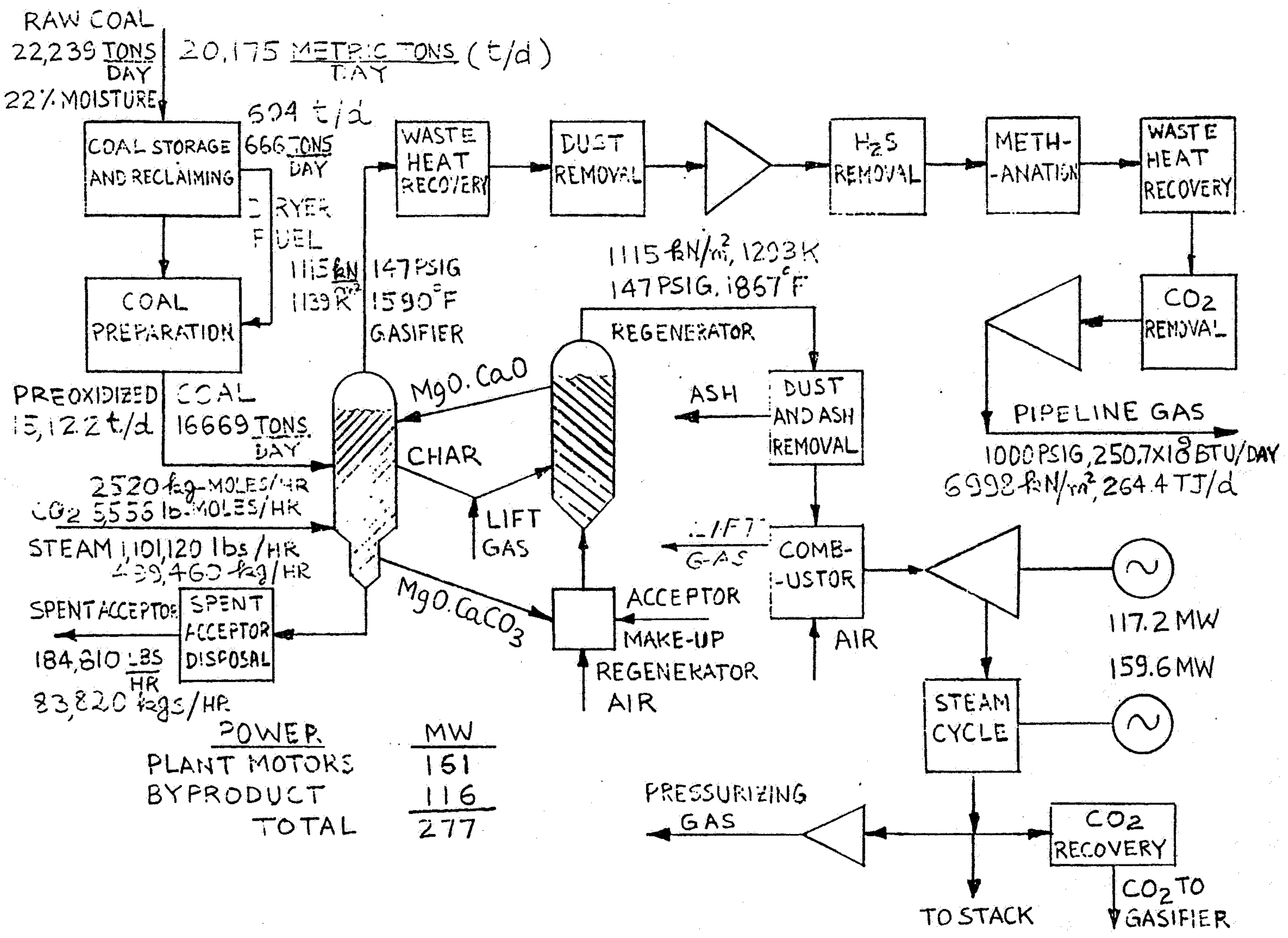

Figure ES-5. PIPELINE GAS BY CO_2 ACCEPTOR PROCESS

The methanation step is carried out in the same type of system used for HYGAS. Because less methane is made in the gasifier than in HYGAS, 63% of the total methane is formed in the methanation reactor. Raw gasifier effluents and product gases for the two processes are compared below (Table ES-3).

Table ES-3. COMPARISON OF RAW AND PRODUCT GASES OF HYGAS® AND CO_2 ACCEPTOR PROCESSES

	Raw Gas		Product Gas	
	HYGAS	CO_2 Acceptor	HYGAS	CO_2 Acceptor
Component	mol % (dry)			
CO	26.1	17.2	0.1	0.1
CO_2	24.1	3.9	0.1	0.1
H_2	30.6	66.1	4.4	4.7
CH_4	16.7	12.0	94.7	94.5
C_2H_6	1.3	--	--	--
NH_3	0.4	0.6	--	--
H_2S	0.2	--	--	--
N_2	0.2	0.2	0.7	0.6
B-T-X	0.4	--	--	--
Total	100.0	100.0	100.0	100.0

JET-FUEL COMPONENT FROM COAL

The proposal for this project does not require process designs for jet fuel, but rather for kerosene. This has been interpreted as allowing us to design processes for making wide-cut kerosene or to increase the supply of jet-fuel component.

Liquids derived from the direct conversion of coal by hydroliquefaction, pyrolysis, and extraction are composed mainly of aromatics, naphthenes, and cycloparaffins. The products are less desirable for jet fuels than the more paraffinic, petroleum-derived fuels.

The basic coal liquefaction process used in this study is the Consol Synthetic Fuel (CSF) Process, which produces heavy fuel oil, naphtha, and high-Btu gas. A published process design based on Pittsburgh seam coal.

is used here. The use of Eastern coal appears justified by summaries published for the H-COAL® Process that show greater advantages for central bituminous than for Western subbituminous coals.

A block flow diagram for the CSF Process plus an add-on refinery is shown in Figure ES-6. Coal is ground, dried, and slurried with a coal-derived solvent, then preheated and extracted at 680 K (765°F) in a staged vessel. Extract and vaporized solvent are sent to the solvent recovery and fractionation systems. Undissolved coal residue is sent to residue separators: Overflow goes to solvent recovery and underflow goes to low-temperature carbonization.

Solvent recovery is divided into two sections. After the vapor from the extraction section is condensed, the gaseous stream is sent to a gas cleanup section and the recovered solvent is returned to slurry-mix tanks. The hydroclone overflow from the residue separation is fractionated in a vacuum still. Light oil and light spent solvent are sent to low-sulfur fuel production.

The heavy oil (including spent solvent and fuel oil) is taken overhead from the fractionator, and a heavier cut from a side stream provides most of the recycle hydrogen-donor solvent for the extraction section. The bottoms (which contain the extract, residue, and tar) are sent to the extract hydroconversion system.

Low-Temperature Carbonization

The hydroclone underflow from the residue separation is pumped to the low-temperature carbonizer (LTC) where it is reacted with steam and air. The overhead product is quenched, and a gas stream and a solvent/tar stream are separated.

The solvent/tar stream is delivered to the tar-distillation section, and the gas stream is used as a plant fuel after sulfur removal. Char from the LTC section is delivered to the Bituminous Coal Research (BCR) gasification system for hydrogen production.

Tar Distillation and Extract Hydroconversion

The heavy liquids from the LTC section are vacuum-distilled in the tar-distillation section. The overhead product is heavy oil. The bottoms are sent

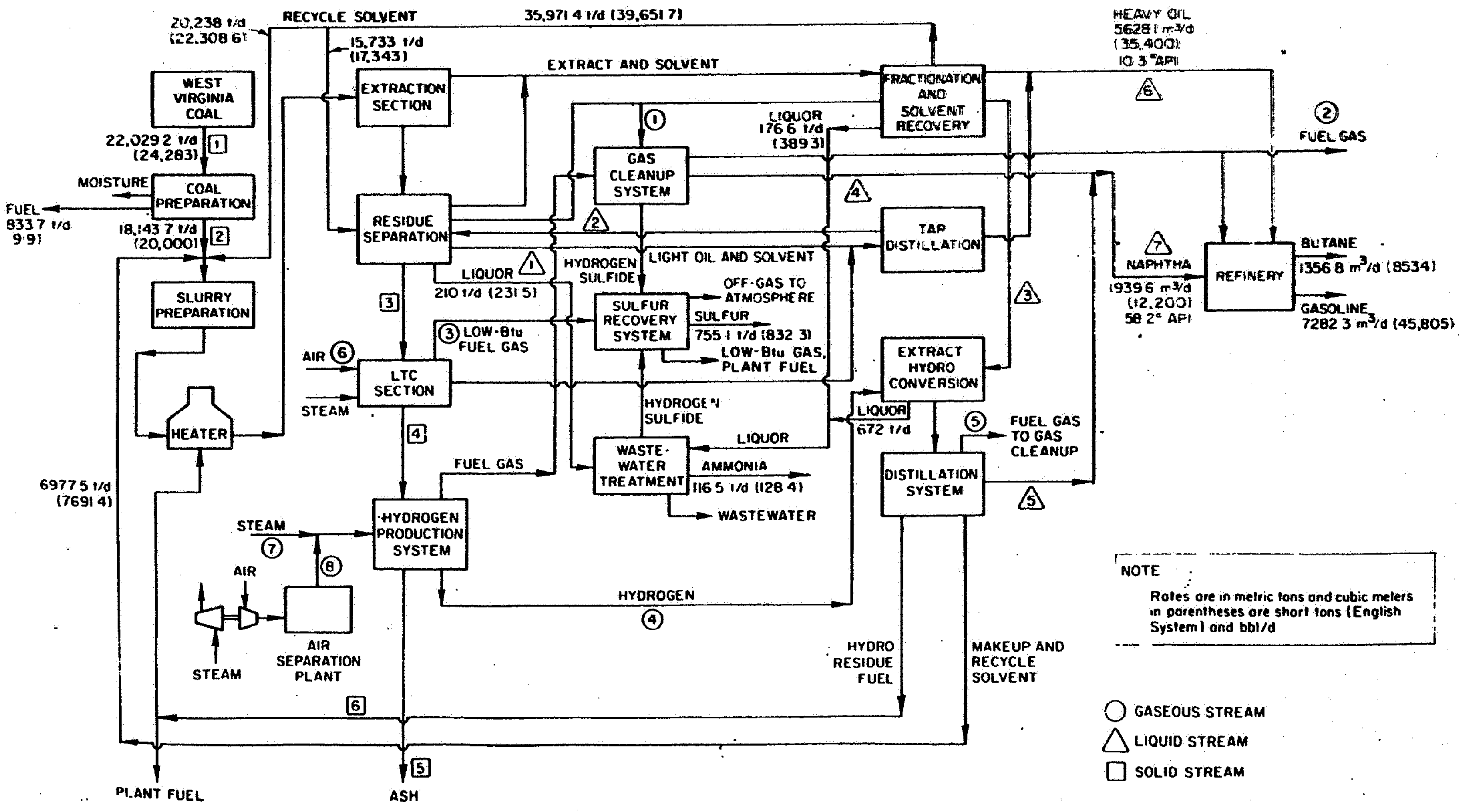

Figure ES-6. GASOLINE FROM COAL, 7282.4 M³/day (45,000 bbl/day) CONSOL SYNTHETIC FUEL PROCESS – LIQUEFACTION OF COAL BY CONSOL SYNTHETIC FUEL PROCESS WITH ADD-ON REFINERY TO YIELD AROMATIC GASOLINE to the residue-separation area and the extract from the solvent-recovery section is hydrotreated to produce the donor solvent and product oil. Extract hydrogenation is done in four stages operating at 20,786 kN/m^2 (3000 psig) and 700 to 714 K (800° to 825°F) in the presence of a cobalt-molybdenum-nickel catalyst. The overhead vapors are cooled to separate the hydrogen from the light oils. The recovery hydrogen is compressed and recycled back to the reactors. The hydrotreated liquid product is stabilized by removing C_4 and then fractionated. In the fractionator, the hydrogen-donor solvent is separated from the light-oil product. The hydrogen-donor solvent is sent to the slurry system for makeup solvent, and the light product oil is delivered to the add-on refinery to produce gasoline.

Gas Treatment and Sulfur Recovery

Fuel gas is produced at various sections of the plant. This gas is treated to remove carbon dioxide and hydrogen sulfide. Some of the fuel gas is used as a fuel for the plant and refinery operation, and the remaining gas is a by-product for sale. The hydrogen sulfide stream from the amine system is passed through the sulfur-recovery system.

Hydrogen Production

Hydrogen is made from synthesis gas generated by the BCR gasifier, in which char from the LTC section is gasified by steam and oxygen. The raw gas is passed through quench systems, CO-shift systems, the acid-gas removal section, and the CO removal section. Char from the LTC is gasified by steam and oxygen in a BCR suspension-type gasifier.

Products from the CSF Process are:

Heavy Fuel Oil	5,629 m^3/day (35,400 bbl/day)
Naphtha	1,940 m^3/day (12,200 bbl/day)
High-Btu Gas	78.59 TJ/day (74.49 X 10^9 Btu/day)

The overall efficiency for converting coal to the above products is 70%.

Jet Fuel From Coal Liquefaction

These products do not meet the product specifications for increased light distillate or wide-cut kerosene jet fuel. Two methods to achieve this objective have been considered in this study:

1. Conversion of CSF products to jet-fuel product
2. Conversion of these products to gasoline that is then imported into the octane pools of petroleum refineries. This added supply of motor gasoline component will reduce the amount of light distillate normally catalytically reformed and thus divert some of this distillate into jet fuel component.

Light Distillate for Jet Fuel

The refinery is shown as a single block in Figure ES-6. A more detailed flow diagram is given in the main body of the report (Figure 8). In the refinery, the heavy oil is hydrocracked. The products plus the naphtha are distilled to yield butane, light ends for gasoline, and a catalytic reformer feed. Products from the reformer are more butane and C_5+ gasoline.

The overall effect is shown in Figure ES-7. About 56% of the high-Btu gas from the CSF Process is utilized in the add-on gasoline plant as fuel and as feed for hydrogen manufacture required for hydrocracking. The products from the CSF-gasoline complex are gasoline, butane, and the remaining high-Btu gas. The latter two are sold as by-products; the 7282 m^3/SD (45,805 bbl/SD) of gasoline is conservatively split among six refineries. In a given refinery, the amount of light distillate normally fed to the catalytic reformer is reduced in the ratio of 1.195 for each volume of coal-derived gasoline imported. Because of yield losses during reforming, this is the rate of increased supply of jet fuel component (light distillate) resulting from the coal conversion to gasoline. Thus, an 8703 m^3/d (54,740 bbl/d) increase in jet fuel supply is achieved from the original coal liquefaction. On the basis of estimates of refinery energy consumption, increasing the output of jet fuel component does not reduce the overall refinery efficiency. Shipping the gasoline an average distance of 1609 km (1000 miles) uses energy amounting to only 0.6% of the coal fed to the CSF plant.

JET FUEL BY HYDROCRACKING

We are indebted to the Process Division of Universal Oil Products Co. (UOP) for giving us process requirements for upgrading the heavy fuel oil to jet fuel. A block flow diagram for the overall process is given in Figure ES-8. In this particular design the naphtha was not used for jet fuel. If naphtha had been used, jet fuel yields would have increased. However, this would not raise the overall efficiency but would lower it somewhat, since we have taken

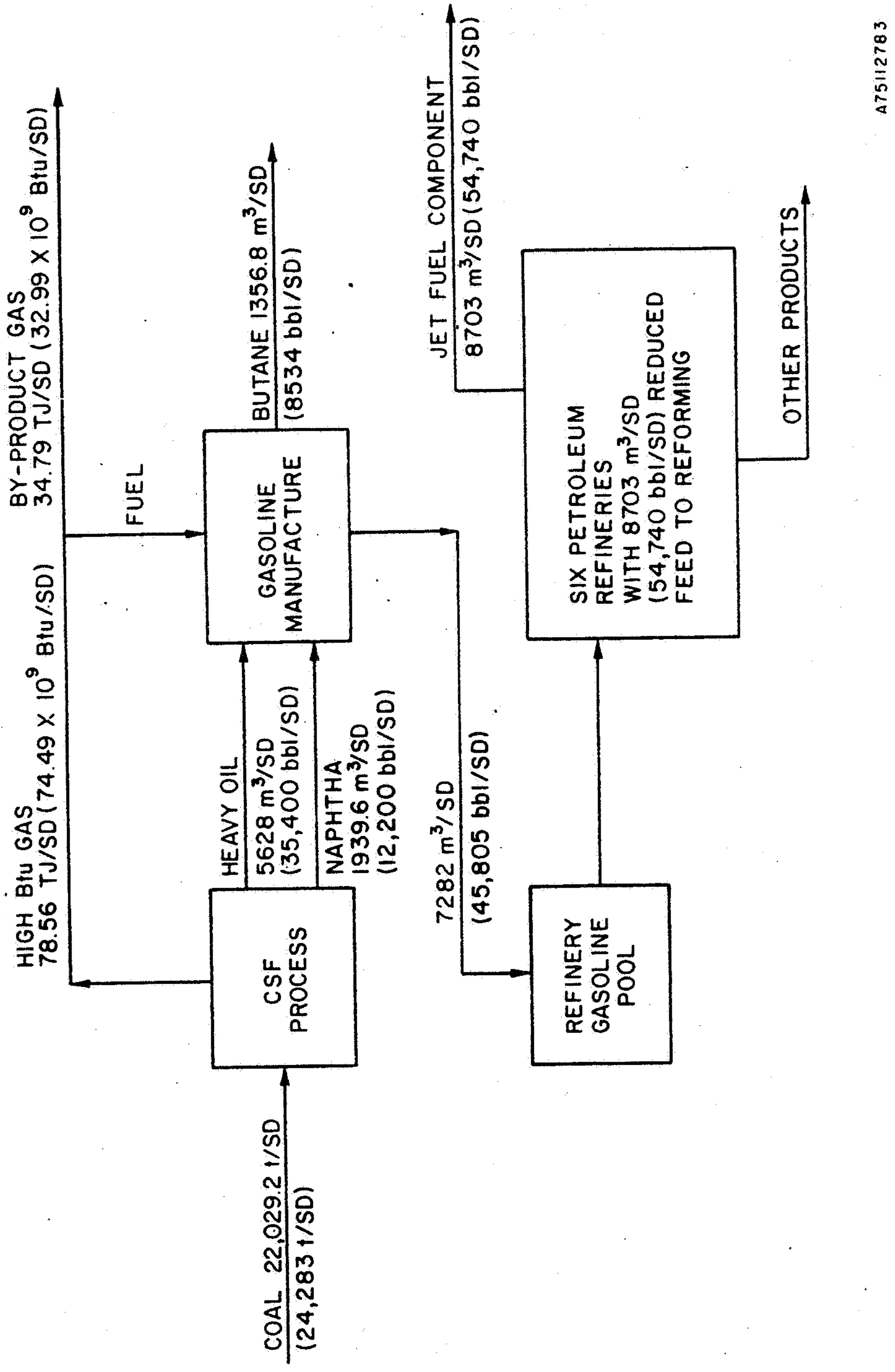

Figure ES-7. OVERALL EFFECT OF IMPORTING COAL-DERIVED GASOLINE INTO REFINERIES TO REDUCE THE AMOUNT OF LIGHT DISTILLATE FEED TO CATALYTIC REFORMING TO INCREASE JET FUEL COMPONENT

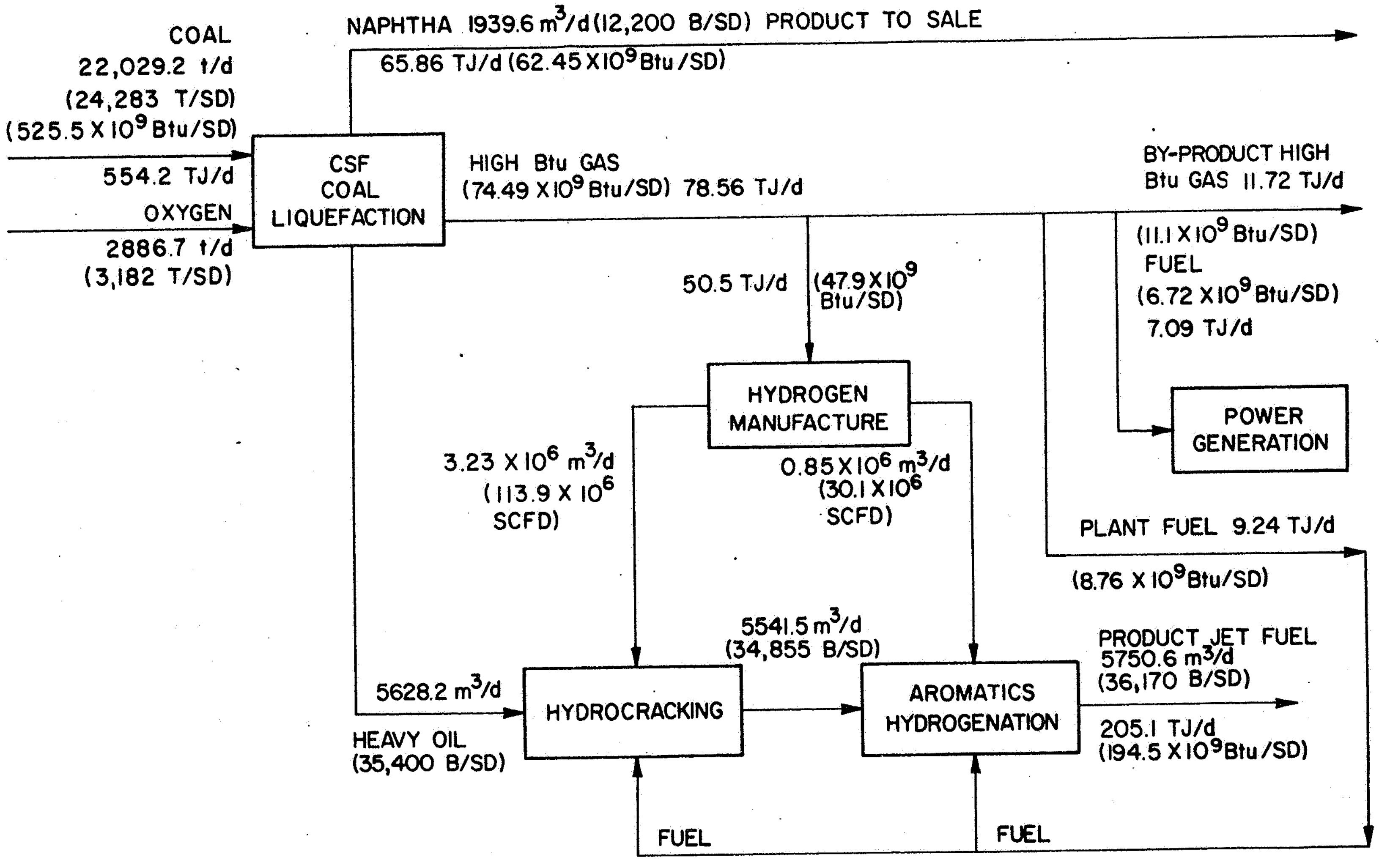

Figure ES-8. MANUFACTURE OF JET FUEL FROM COAL-DERIVED HEAVY FUEL OIL full credit for this stream as a by-product from the CSF Process. A two-step operation is used to convert 5628.2 m^3/SD (35,400 bbl/SD) of heavy oil to 5750.6 m^3/SD (36,170 bbl/SD) of final product. Most of the high-Btu gas from the CSF Process is used for hydrogen manufacturing and for fuel. The first step, hydrocracking, reduces the aromatics content from 83.5% to 39%, raises the API gravity from 15.7 to 41, and reduces the boiling range. Eighty-one percent of the hydrogen is consumed here. The second step is aromatics hydrogenation, in which the percentage of these compounds is reduced to a final product value of 17% and the smoke-volatility index is raised to 57.

OVERALL EFFICIENCY

One of the major tasks of this work is the comparison of overall efficiencies for the different coal-conversion processes. An overall summary of material requirements, yields, and overall efficiencies is shown in Table ES-4, which also appears as Table 53 in the main body of the report. All seven processes are compared here, and the overall efficiencies are calculated on the basis of product heating values as a percentage of coal heating value. Higher heating values (HHV) are used. Total coal input heating values range from 363.7 to 591.7 TJ/d (344.8 to 561 X 10^9 Btu/d).

Hydrogen Processes

The U-GAS Process shows the highest overall efficiency, 66.4%, compared to the Steam-Iron Process at 62.6%, and the Koppers-Totzek Process at 57%. However, with steam-iron, 18.0% of the coal feed is converted to electric power, a high-grade product that normally is generated at a low fuel conversion efficiency. It is essential that some use be made of the large amount of heat energy that remains in the spent producer gas in order to raise the overall efficiency from 44.6% for hydrogen alone to the final value. The generation of electricity is a way of upgrading this heat to a universally valuable product. If the heat energy input to the power plant could be sold as by-product, then the overall efficiency would be 81.5%.

Hydrogen by Koppers-Totzek gasification, although a commercially operating process, requires about 40% more oxygen than the U-GAS Process, primarily because of the much higher gasifier temperature. The lower oxygen

Table ES-4. SUMMARY OF PROCESS MATERIAL REQUIREMENTS, PRODUCT YIELDS, AND OVERALL EFFICIENCIES

Main Product	Hydrogen						Methane				Jet Fuel Component Via CSF Coal Liquefaction			
Gasifier	Koppers-Totzek		U-Gas		Steam Iron		HYGAS®		CO_2 Acceptor		Increased Refinery Output by Importing Gasoline from Coal			Jet Fuel by Hydrocracking Synfuel from coal
	kg/h	lb/h	kg/h	lb/h	kg/h	lb/h	kg/h	lb/h	kg/h	lb/h		kg/h	lb/h	
Reaction Coal (dry)	557,733	1,229,590	494,144	1,089,400	909,035	2,004,080	471,613	1,039,728	630,072	1,389,070	——	755,980	1,666,670	——
Fuel Coal (dry)	179,721	396,218	138,055	304,360	30,096	66,350	108,675	232,974	25,619	56,483	——	29,730	65,550‡	——
Total Coal (dry)	737,454	1,625,808	632,199	1,393,131	939,131	2,070,430	577,288	1,272,702	655,291	1,445,553	——	783,710	1,732,220	——
												H_2 Only		
Gasifier Steam	122,544	270,164	170,449	375,776	2,086,166	4,599,209	455,013*	1,003,131*	499,460	1,101,120	——	161,025	355,000	——
CO Conversion Steam	734,900	1,620,178	623,787	1,375,215	No CO Conversion		459,875	1,013,650	No CO Conversion			--	--	
	Metric tons/d	Short tons/d	Metric tons/d	Short tons/d	Metric tons/d	Short tons/d	Metric tons/d	Short tons/d	Metric tons/d	Short tons/d		Metric tons/d	Short tons/d	
Total Coal (as received)†	22,690	25,012	19,453	21,443	28,897	31,853	17,763	19,580	20,175	22,239	——	22,029	24,283	——
Oxygen	10,969	12,092	7,830	8,631	15,756	17,368	2,720	2,999	8,769	9,666	——	2,887	3,162	——
	(98% Purity)		(98% Purity)		(as Air)		(98% Purity)		(as Air)			99.5%		

	Koppers-Totzek			U-Gas			Steam Iron			HYGAS®			CO_2 Acceptor			Increased Refinery Output by Importing Gasoline from Coal			Jet Fuel by Hydrocracking Synfuel from coal		
	TJ/d	10^9 Btu/d	%	TJ/d	10^9 Btu/d	%	TJ/d	10^9 Btu/d	%	TJ/d	10^9 Btu/d	%	TJ/d	10^9 Btu/d	%	TJ/d	10^9 Btu/d	%	TJ/d	10^9 Btu/d	%
H. H. V. of Coal	464.6	440.5		398.3	377.7		591.7	561.0		363.7	344.8		413.1	391.7		554.2	525.5		554.2	525.5	
H. H. V. Main Product	264.0	250.3	56.8	263.9	250.2	66.2	263.9	250.2	44.6	254.7	241.5	70.0	264.4	250.7	64.0	310.4	294.3	45.8††	205.1	194.5	37.0
H. H. V. Naphtha	--	--	--	--	--	--	--	--	--	--	--	--	--	--	--	--	--	--	65.9	62.5	11.9
H. H. V. C_4	--	--	--	--	--	--	--	--	--	--	--	--	--	--	--	39.2	37.2	7.1	—	--	--
H. H. V. Power	--	--	--	--	--	--	106.1	100.7	18.0	--	--	--	10.0	9.5	2.4	--	--	--	--	--	--
H. H. V. Other Products**	0.8	0.8	0.2	0.7	0.7	0.2	—	--	--	14.3	13.6	4.0	2.9	2.7	0.7	44.4	42.1	8.0	21.3	20.2	3.8
H. H. V. of Other Coal to Products and Plant Efficiency	264.8	251.1	57.0	264.6	250.9	66.4	370.0	350.9	62.6	269.0	255.1	74.0	277.3	262.9	67.1	394.0	373.6	60.9	292.3	277.2	52.7

* Obtained from vaporized slurry water used in feeding coal to the gasifier.

† For hydrogen and methane, Montana subbituminous coal at 22% moisture; for jet fuel, Pittsburgh seam bituminous coal at 14.4% moisture.

‡ Fuel indicated is for CSF Process design only. Fuel for gasoline and jet fuel manufacture is obtained from high-Btu gas made in CSF Process.

** Ammonia and sulfur; high-Btu gas from CSF Process.

†† C_5+ gasoline. Increased refinery light distillate after importation of this gasoline is equivalent to 56% of coal HHV.

requirement and higher pressure of the U-GAS gasifier combine to save much energy and give an efficiency significantly higher than the 57% overall efficiency for Koppers-Totzek.

Methane Processes

By the use of water slurry feed for Montana coal, we have achieved the highest design efficiency for the HYGAS Process reported to date, 74%. Of this figure, 4% is due to by-products, mostly B-T-X generated in the HYGAS reactor. The by-products are relatively easy to extract from the raw gas. The water used to slurry the coal is vaporized in the slurry dryer and provides the steam for the CO-shift reaction, eliminating the losses that would result from condensation and subsequent revaporization in a boiler. Although methanation, an exothermic reaction, is required as a final step in the process, it does not detract much from the overall efficiency because most of the heat is recovered for use in the process. Reduced oxygen consumption, process steam, and process horsepower are the major reasons for the higher efficiency for the conversion of coal to methane rather than hydrogen.

The alternative process, the CO_2 Acceptor Process, converts 64% of the feed coal to SNG product, 6% less than the HYGAS Process. The major reason for this is probably the larger amount of power needed to run the CO_2 Acceptor Process, 330,500 kW versus 102,000 kW for HYGAS. The former figure includes 169,000 kW to drive the regenerator air compressor. While power for the latter is recovered from the regenerator off-gas, there are losses due to inefficiencies in compressors and expanders. We have used a combined-cycle power recovery system, similar to one in the Steam-Iron Process, which generates 115 MW of by-product electric power, which represents 2.4% of the total coal input HV.

Jet Fuel Component

The overall process of using coal to manufacture gasoline that is then imported into the refinery gasoline pool, which allows more refinery output of light distillate, is more efficient that the direct conversion of coal to jet fuel. The overall efficiency for the CSF Process is 71%. This drops to 61% because of the energy required to upgrade the CSF products to gasoline.

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Importing 7282 m^3/SD (45,805 bbl/SD) of coal-derived gasoline results in an increase in the jet fuel component of 8703 m^3/d (54,740 bbl/d). When referred to the original coal, this results in an apparent increase in overall efficiency to 70% because of the larger volume of jet fuel produced. However, the increased volume of jet fuel causes a reduction in refinery LPG, coke, and gas production due to reduced catalytic reforming. The changes in products and refinery economics have been accounted for in the economic analysis, so that the unit cost of the increased light distillate can be referred to the full 8703 m^3/d (54,740 bbl/d). Direct conversion of CSF heavy oil to jet fuel shows an overall efficiency of only 53% because of the larger hydrogen and fuel requirements. In both these cases, the percent conversion to liquid products leaving the coal plant is less than for either of the two methane-making processes discussed below.

Liquid Fuels

Although it was not a required part of this study, because of the basic interest in aircraft fuels, we estimated overall efficiencies for the conversion of coal to liquid hydrogen and methane for the three hydrogen processes and for HYGAS. These results are given in Table ES-5. The requirements for hydrogen liquefaction are based on work done by the Linde Division of Union Carbide Corp. for another NASA project. The energy requirement for methane liquefaction is typical for LNG plants.

The additional energy load for liquefaction requires additional coal for producing hydrogen by the Koppers-Totzek and U-GAS Processes, while the Steam-Iron Process generates more than the necessary electric energy as a by-product. On the basis of using coal for the generation of hydrogen and electricity for liquefaction, and with the inclusion of by-product, the Steam-Iron Process shows the highest overall efficiency for the hydrogen processes.

The power requirement for methane liquefaction is just a fraction of that needed to liquefy hydrogen. The overall efficiency from coal to liquid is thus much higher for methane than for hydrogen: 66.7% versus 34% to 48.5%, depending on the hydrogen process. These numbers include by-products as well as liquid hydrogen and methane.

Table ES-5. LIQUEFACTION OF HYDROGEN AND METHANE FROM 250 BILLION Btu/DAY PRODUCT GAS DERIVED FROM COAL GASIFICATION

	Hydrogen			Methane
Process	Koppers-Totzek	U-GAS™	Steam-Iron	HYGAS®[d]
Feed to Liquefaction Plant, mol/hr				
H_2	71,443	73,072	80,767	1,250
CH_4	4,252	3,722	1,245	26,776
CO	77	76	84	29
CO_2	4	8	84	29
N_2 + Ar	977	646	2,220	192
H_2O	10	8	11	4
Total	76,763	75,524	84,411	28,280
Product Liquefied, mol/hr[a]	62,656	64,084	70,833	24,362
Product Fuel Value, 10^6 Btu/hr	7,707.9	7,883.6	8,713.9	9,334.5
Power for Liquefaction, kW	682,096	697,642	771,115	102,800
Fuel for Liquefaction, 35% Efficiency, 10^6 Btu/hr	6,651.4	6,803.0	--	1,002.45
Excess Power, kW	--	--	458,385[b]	--
Less Fuel Gas Recovered, 10^6 Btu/hr	−2,361.7	−2,068.3	1,054.7	1,002.45
Additional Coal Required, 10^6 Btu/hr	4,289.7	4,734.7	--	--
Gasification Plant Coal, 10^6 Btu/hr	18,355.4	15,735.6	23,374.6	14,876
Total Coal, 10^6 Btu/hr	22,645.1	20,470.3	23,374.6	14,876
Overall Efficiency, Coal to Liquid H_2, or CH_4, %	34.0[e]	38.5[f]	48.5[c]	62.7 (plus 4.0% by-product)

[a] 87.7% of input hydrogen, 91% of input methane.

[b] Steam-Iron Process yields 1,229,500 kW of by-product power, more than enough for liquefaction.

[c] Includes liquid hydrogen, excess electric power, and recovered fuel gas — 11,33[illegible].1 X 10^6 Btu/h.

[d] Adjusted for 250 billion Btu/day.

[e] With 48% efficiency for conversion of fuel to electricity, overall efficiency is 37%.

[f] With 48% fuel to electricity conversion efficiency, overall efficiency is 42.3%.

ECONOMICS OF COAL CONVERSION

Table ES-6 presents an overall economic summary for five of the processes. The original plan was to present costs for three processes — one in each category. However, after the attractive overall efficiencies for the Steam-Iron Process were presented, funding was added to cost this process. As cost information was available for both of the alternative liquid processes, we decided to include these. The table shows only the major quantities. Additional details can be obtained from the portion of the text describing a given process.

The project plan calls for mid- 1974 costs and the use of the private industry financial method for calculating product price. We have used the discounted cash flow method of the Supply-Technical Advisory Task Force-Synthetic Gas-Coal of the Federal Power Commission (FPC). This method is based on 100% equity financing and 12% DCF. Other details are given in the text.

In summary Table ES-6, daily fuel production includes not only the major product but significant by-product fuels and, for steam-iron, by-product electric power. Total plant investment and capital requirements are indicated separately. Capital requirements include interest during construction, start-up cost, and working capital. When capital requirements are expressed on a unit basis, HYGAS (methane) and increased refinery light distillate (resulting from importation of coal-derived gasoline) show the lowest unit cost. The highest unit cost is the direct conversion of coal-derived heavy oil to jet fuel.

Annual operating costs are summarized as coal, other operating costs, by-products, and capital charges. We have shown product prices for coal at both 28.4¢ and 56.8¢/GJ (30¢ and 60¢/million Btu), representing the basic costs for the Western coals used for hydrogen and methane and for the Eastern coals used for liquids in this study. Figure ES-9 shows the effect of coal cost on product prices. At the lower cost, coal represents less than 25% of the total revenue for the U-GAS and HYGAS Processes. At the higher cost it represents about 40% of the revenue for the liquids. Over the range of coal costs presented, liquids are the most expensive products. For hydrocracking, this is due mainly to the small amount of product. For the alternative liquid

Table ES-6. SUMMARY OF ECONOMICS OF COAL CONVERSION TO FLUID FUELS; 25-YEAR LIFE — MID-1974 COSTS

Product	Hydrogen		Methane	Lt. Distillate for Jet Fuel; CSF Coal Liquefaction to Gasoline	Jet Fuel by Conversion of Heavy Oil from the CSF Process
Process	U-Gas	Steam-Iron [b]	Hygas [c]	For Refinery Import [d]	[e]
Daily Fuel Production, TJ (10^9 Btu)	263.9 (250.2)	370.1 (350.9)	267.2 (253.3)	384.4 (364.47)	282.69 (268.03)
Total Plant Investment, \$$10^6$[a]	409.32	480.51	330.10	455.88	487.35
Total Capital Requirement, \$$10^6$	539.79	532.58	435.92	616.55	654.57
Unit Capital, \$/Daily GJ (10^6 Btu)	2045 (2157)	1709 (1803)	1631 (1721)	1604 (1591)	2316 (2442)
Annual Operating Costs and \$$10^6$					
Coal, (¢/GJ) (¢/10^6 Btu)	37.218 (28.44) (30)	55.286 (28.44) (30)	33.984 (28.44)(30)	103.572 (56.8) (60)	103.572 (56.8) (60)
Other operating Costs	38.075	52.090	31.333	50.144	52.265
Gross Operating Costs	75.293	107.376	65.317	153.715	155.837
Byproduct Credits	-0.250	-95.932	-7.922	-36.459	-43.786
Net Subsidy to Oil Refineries	-	-	-	+32.799	--
Net Operating Cost	75.043	10.444	57.395	150.046	112.051
Capital Changes	103.050	120.745	83.262	118.266	125.466
Annual Revenue Required	178.095	131.189	140.657	268.312	237.517
25-Yr. Avg. Major Product Price					
Coal at 28.4¢/GJ	\$2.05/GJ	\$1.51/GJ	\$1.68/GJ	\$2.10/GJ	\$2.72/GJ
(30¢/10^6 Btu)	(\$2.17/10^6 Btu)	(\$1.60/10^6 Btu)	(\$1.77/10^6 Btu)	(\$2.21/10^6 Btu)	(\$2.87/10^6 Btu)
" " 56.8¢/GJ	\$2.52/GJ	\$2.15/GJ	\$2.11/GJ	\$2.64/GJ	3.53/GJ
60¢/10^6 Btu)	(\$2.65/10^6 Btu)	(\$2.28/10^6 Btu)	(\$2.22/10^6 Btu)	(\$2.78/10^6 Btu)	(\$3.72/10^6 Btu)

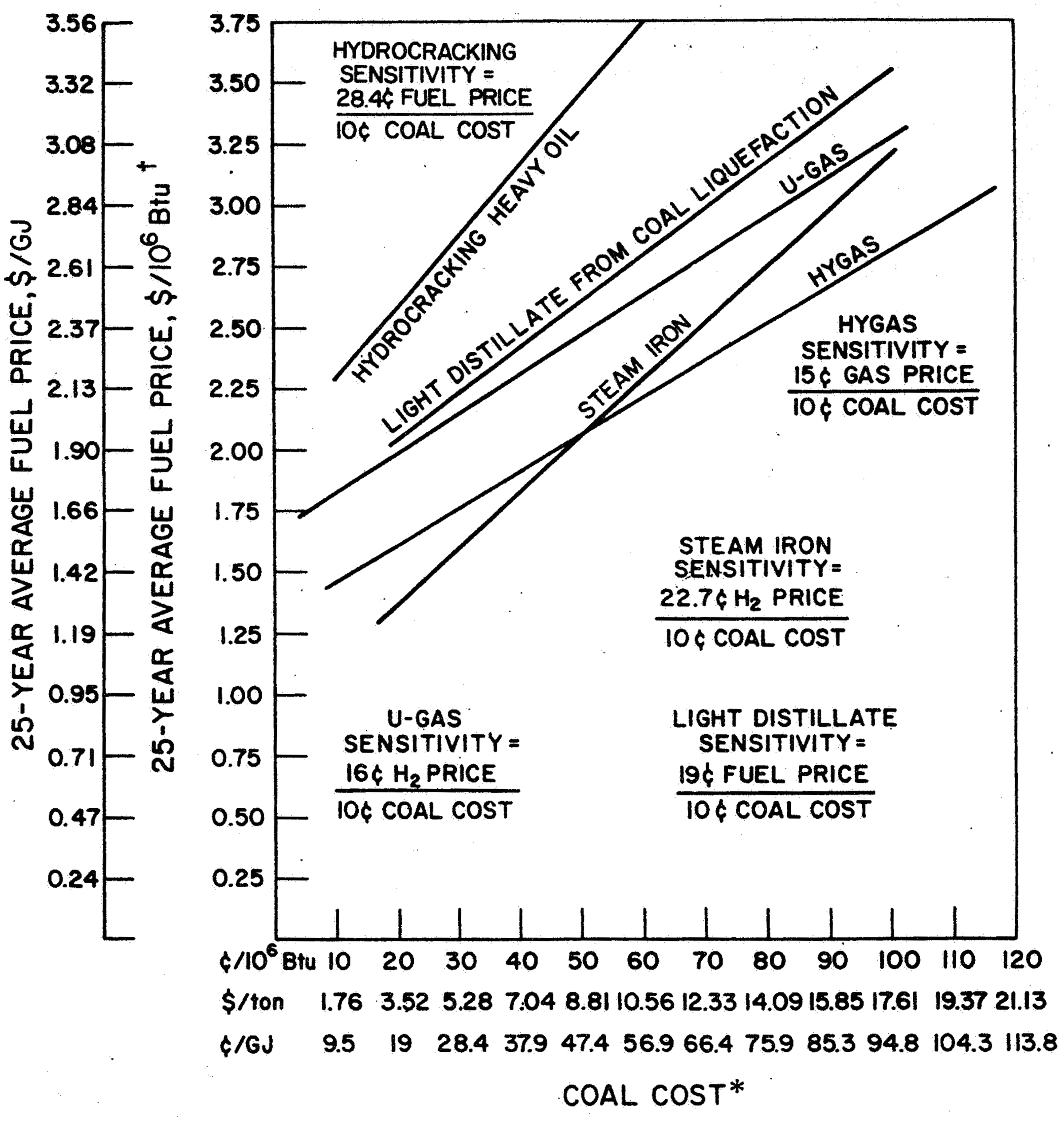
a. - Includes contingencies at 15% and contractor's charges at 15%.

b. - Includes 250.2 X 10^9 Btu/d hydrogen and 100.7 X 10^9 Btu/d by-product electric power.

c. - Includes 241.5 X 10^9 Btu/d pipeline gas and 11.77 X 10^9 Btu/d B-T-X liquid.

d. - Includes 294.3 X 10^9 Btu/d light distillate, 37.2 X 10^9 Btu/d Butane, and 33.0X10^9 Btu/d high Btu gas.

e. - Includes 194.47 X 10^9 Btu/d Jet Fuel, 62.45 X 10^9 Btu/d Naphtha, and 11.11 X 10^9 Btu/d high Btu gas.



* 26,251 kJ/kg (11,290 Btu/lb) DRY H.V., 22% MOISTURE AS RECEIVED.

† Major product.

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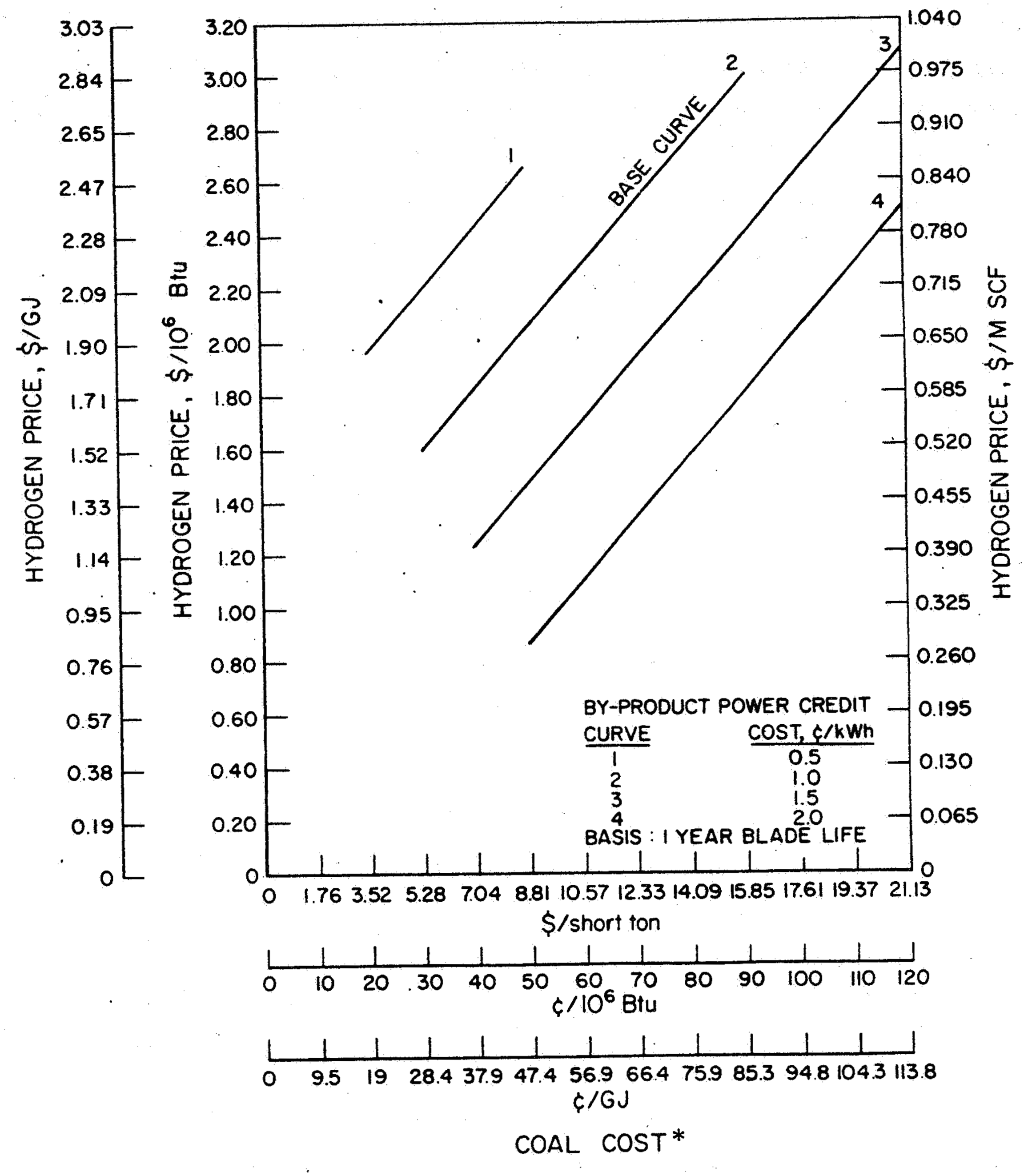
Figure ES-9. EFFECT OF COAL COST ON FUEL PRICE BY VARIOUS PROCESSES case, the higher cost is due mainly to the net subsidy paid to oil refiners to compensate for the decrease in revenue resulting from making more light distillate and less gasoline. In our study, gasoline was assumed to be worth 1.05¢/liter (4¢/gal) more than light distillate. The sensitivity figures shown for each process are a function of the ratio of major product to total coal input, expressed in the same energy units.

Hydrogen produced by the Steam-Iron Process differs because of the large by-product credit resulting from the sale of 1229 MW of power. The line in Figure ES-9 is based on setting the value of by-product power at 1¢/kWhr. Because of the great leverage exerted by the unit value of power, the effect of both coal cost and electric power by-product value are shown in a separate figure, Figure ES-10. A variation of 0.5¢/kWhr changes the hydrogen price by 56.8¢/GJ(60¢/million Btu).

Comparative economics for the U-GAS and HYGAS Processes indicate that methane from coal costs less than hydrogen from coal, if a major amount of methane is made in the gasifier. Methane formation supplies some of the heat for gasification, greatly reducing the oxygen requirement. Spent hydrogasifier char is used to generate hydrogen. If coal were first gasified (to $CO + H_2$) and all methane made by methanation, then this advantage would be lost. Compared with the U-GAS Process, the hydrogasification (HYGAS) reactor system costs more because of its greater size, its complexity, and its high operating pressure. However, the difference in reactor system costs is outweighed by the much higher costs for the oxygen supply, purification, and the associated utilities necessary for hydrogen produced by the U-GAS Process.

Although the Steam-Iron Process demonstrates hydrogen production at a good advantage, we also expect these advantages to benefit methane from coal if the Steam-Iron Process is used to supply hydrogen for HYGAS.

The direct conversion of coal-derived heavy oil to jet fuel is much less attractive than the indirect route of using coal-derived gasoline to supply part of the refinery work, so that the latter can increase the supply of jet fuel. The main reason is the difference in final product output. The annual revenue for direct conversion is actually a little less than for the alternative method, but this difference is outweighed by the much larger output for the alternative methods as reflected in the much lower unit capital cost.



*26,251 kJ/kg (11,290 Btu/lb) DRY HHV, 22% MOISTURE AS RECEIVED.

Figure ES-10. EFFECT OF COAL COST ON HYDROGEN PRICE, FROM MONTANA SUBBITUMINOUS COAL BY THE STEAM-IRON PROCESS

CONCLUSIONS

1. Coal can be converted to gaseous hydrogen and methane and to normally liquid hydrocarbon fuels without a serious energy loss. Overall coal conversion efficiencies ranged from 57% for the least efficient hydrogen process to 74% for methane production.

2. Based on the processes analyzed in this study, the conversion of coal to methane is more efficient and less expensive than its conversion to hydrogen. This is not true where all of the methane is generated by the methanation of synthesis gas (CO + H_2). However, if the major part of the methane is made during the gasification step, then the amount of oxygen or other heat source used can be reduced to a fraction of that needed to make hydrogen only.

3. The results of this study for methane gas and hydrocarbon liquids manufacture do not support a generalized conclusion as to which product can be made more efficiently from coal; this depends on the liquid product. For the more refined liquid products (such as gasoline), methane produced from Western coal by the HYGAS Process shows an advantage.

4. When all three products are produced as liquids, then the order of decreasing overall efficiency is: liquids (at normal ambient conditions), methane, and hydrogen. This statement also depends on the liquid product, as noted in Item 3. The greatest loss of efficiency occurs in the liquefaction of hydrogen, which, for the same heating value of gas liquefied, requires about seven times more energy to liquefy than does methane.

5. The major items of cost in the conversion of coal are capital investment and coal. Although Montana coal, at a base cost of 28.4¢/GJ (30¢/million Btu) is used for hydrogen and methane, and Pittsburgh seam coal, at 56.8¢/GJ (60¢/million Btu) is used for liquids, when product prices are compared at equal coal costs, HYGAS and steam-iron show the lowest costs. At the lower coal cost the range of major product prices is $1.57/GJ ($1.60/million Btu) for steam-iron hydrogen to $2.72/GJ ($2.87/million Btu) for jet fuel. At the higher coal cost the range is $2.11/GJ ($2.22/million Btu) for HYGAS to $3.53/GJ ($3.72/million Btu) for jet fuel. Methane and hydrogen are priced as gas at 6996 kN/m^2 (1000 psig).

6. The Steam-Iron Process rejects a large amount of heat in the form of hot, spent producer gas. This heat is advantageously used to generate a large amount of by-product electricity. Because of its high market value, this electricity exerts a strong leverage on the hydrogen price. At bus-bar by-product credits of 1.5¢ to 2.0¢/kWhr, the Steam-Iron Process shows the lowest product price of any of these processes. However, further work should be done to determine the economics for the other processes under conditions of joint power and major product production, at total coal inputs the same as for the Steam-Iron Process, in order to avoid a biased conclusion.

7. The advantages of the Steam-Iron Process can be used to supply hydrogen for the HYGAS Process. For the same output of product gas heating value as methane, only about 40% as much hydrogen is needed as when hydrogen is the product, so the effect of by-product power will be less. Further work is needed on this process.

8. Although methane shows cost and efficiency advantages over hydrogen when the two are made from coal, the results do not preclude the use of hydrogen as fuel. Hydrogen can also be made from water using other energy sources, such as nuclear or solar energy, whereas methane requires a source of carbon. As coal becomes more costly, or its use reserved for chemicals, hydrogen from nonfossil sources should become more important.

9. Although jet fuel can be made from coal, it requires that the high aromatic content of the liquefaction products be reduced at a large loss in overall efficiency. It appears preferable to use the aromatic materials as the gasoline component imported into the petroleum products system. This, in turn, will permit the diversion of catalytic reformer feedstock into the supply of jet fuel. Optimization of this method should be investigated.

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STUDY OF THE CONVERSION OF COAL TO HYDROGEN, METHANE, AND LIQUID FUELS

Introduction and Project Scope

The shortage of U.S. supplies of petroleum and natural gas has greatly stimulated research and developmental work on the conversion of coal to fluid fuels. Since aircraft jet fuel is only one of the fuels derived from petroleum, the shortage of this basic resource plus the increasing demands for other petroleum products may limit the supply of jet fuel from this source.

The abundance of U.S. coal resources and the ability to convert coal to many other fuels have prompted the consideration of various alternative fuels. As a fuel, hydrogen has shown a potential advantage for use in the aircraft itself. It is the purpose of this study to compare overall efficiencies and costs on a consistent, comparable basis for the conversion of coal to hydrogen, methane, and kerosene or wide-cut jet fuel.

Processes for Coal Conversion

There are many coal conversion procedures, some available commercially and others currently under development. We have selected processes that are expected to be available for the production of alternative fuels in the period when hydrogen-fueled aircraft might actually be in use, 1985 to 2000. The scope of this work covers the manufacture of the above three fuels at a coal conversion plant located close to the coal mine. Products leave the plant at pipeline pressure, 6996 kN/m^2 (1000 psig).

Plant Capacity

A recent study made by Linde for airport supply of liquid hydrogen was based on 2268 t/d (2500 short tons/d). This is equivalent to 324.8 TJ/d (308 billion Btu/d). Another study, which considered the supply situation for the San Francisco airport, was based on an input of 6000 MW of electric power to generate hydrogen — 389 TJ/d (369 billion Btu/d) — equivalent to a 518.9 TJ/d (492 billion Btu/d) energy input and, at 75% overall efficiency, 389 TJ/d (369 billion Btu/d) of hydrogen output. (These numbers are at the general level of output of standard pipeline gas-from-coal plants, 263.7 TJ/d (250 billion Btu/d). Since the different sections of synthetic natural (pipeline) gas plants consist of parallel trains, we are probably beyond the point where economies of scale are significant. Since the heating value per SCF of hydrogen is one-third that of methane, the level where economy of scales has no effect is probably less. Therefore, for the purpose of this study, plant capacity for the manufacture of hydrogen or pipeline gas from coal will be approximately 263.7 TJ/d (250 billion Btu/d). For kerosene from coal, the equivalent output would be obtained from about 6757 m^3/d (42,500 bbl/d).

Product Quality

This is governed partly by the requirements for liquefaction.

1. Hydrogen — A minimal 92% purity; the remainder can be H_2, Argon, CO, CH_4, and CO_2. It is preferable to reduce the CO_2 to 0.1% or less because, while the cost of removing the first four components is mainly in extracting the last few ppm, the cost of removing CO_2 is more dependent on the total amount removed.
2. Methane — A typical pipeline gas from coal would be satisfactory, i.e., a composition of 92% to 94% CH_4, and the rest H_2, N_2, Ar, CO, and CO_2, with the two carbon oxides at the 0.1% level.

Kerosene or Wide-Range Jet Fuel

Although jet fuel is the final desired product, the proposal for this study did not contemplate the inclusion of facilities to produce a specification product, and no specifications were proposed. The project plan calls for an estimate of the energy requirements (overall efficiency) for converting coal to syncrude and upgrading the crude to a kerosene-type product. We have considered refinery light distillate or wide-cut kerosene as suitable products for this part of the study.

Type of Coal

The scope of this study does not include the effect of different coals on a particular product or process. For purposes of comparability, it is desirable to keep coal variations to a minimum; however, we have not specified a single plant location and are not bound to a particular coal supply. Thus, there is also a good rationale for considering the coal feed best suited for the particular process considered.

Montana subbituminous coal was selected for the manufacture of hydrogen and methane because of its greater reactivity and nonagglomerating characteristics during gasification, its abundance, its low-sulfur content, and its low cost compared with Eastern coals. For liquefaction, as discussed in another section of this report, Eastern coal appears to have process advantages over Western coal. The fact that Eastern coal costs more presents no problem in comparability because we have presented product prices as a function of coal cost, so it is possible to make comparisons at the same coal cost. Tables 1 and 2 present analyses for these two coals.

Table 1. ANALYSIS OF MONTANA SUBBITUMINOUS COAL USED IN THE PROCESS DESIGNS FOR HYDROGEN AND METHANE

	As Received	Dry
Proximate Analysis	wt %	wt %
Moisture	22.0	--
Volatile Matter	29.9	37.7
Fixed Carbon	42.6	54.6
Ash	6.0	7.7
Total	100.0	100.0
Ultimate Analysis (Dry)		
Carbon	67.70	
Hydrogen	4.61	
Nitrogen	0.85	
Oxygen	18.46	
Sulfur	0.66*	
Ash	7.72	
Total	100.00	

Dry Heating Value = 26,251 kJ/kg (11,290 Btu/lb)
As Received HHV† = 20,475 kJ/kg (8806 Btu/lb)

* This is equivalent to 0.503 kg/GJ (1.17 lb $SO_2/10^6$ Btu), which is less than the Federal standard, 0.516 kg/GJ (1.2 lb $SO_2/10^6$ Btu) and the Montana standard, 0.86 kg/GJ (2.0 lb $SO_2/10^6$ Btu).

† Based upon maximum moisture content of 22.0 weight %.

Table 2. ANALYSIS OF PITTSBURGH SEAM COAL USED IN CSF COAL LIQUEFACTION PROCESS DESIGN

	As Received	Dry
	wt %	
Proximate Analysis		
Moisture	14.4	--
Volatile Matter	33.4	39.07
Fixed Carbon	40.9	47.71
Ash	11.3	13.22
Total	100.0	100.00
Ultimate Analysis (Dry)		
Hydrogen	4.90	
Carbon	68.97	
Nitrogen	1.28	
Oxygen	7.34	
Sulfur	4.29	
Ash	13.22	
Total	100.00	

MF Gross Heating Value = 29,390 kJ/kg (12,640 Btu/lb)(Dulong)
As Received HHV* = 25,158 kJ/kg (10,820 Btu/lb)

* Based upon a maximum moisture content of 14.4 weight %.

Project Tasks

There are two major areas involved in this study — analyses of process efficiency and process economics. These tasks have been subdivided as follows:

1. Process Efficiency Analysis — Process designs were developed for seven different methods of converting coal to liquid fuels. To have a reasonable measure of overall efficiency, one must go into considerable detail in developing material balances, energy balances, and determining utility requirements. Efficiencies for converting coal to the different products have been developed for the seven processes considered in this study.

   a. Hydrogen — Three processes were evaluated, based on different gasifier technologies: 1) Koppers-Totzek suspension gasification, 2) U-GAS fluidized-bed gasification, and 3) the continuous Steam-Iron Process.

b. Methane — Two processes were evaluated, again based on different gasification technologies: 1) the HYGAS Process and 2) the Consolidation Coal CO_2 Acceptor Process.

c. Kerosene or Jet-Fuel Component — For this product, two methods of upgrading syncrude are evaluated, both based on the same coal liquefaction process, the Consolidation Coal Synthetic Fuel Process (CSF). A high-aromatic-content gasoline is produced by one method; this gasoline is exported to petroleum refineries into the gasoline pool. This reduces the required amount of catalytic reforming and permits increased output of light distillate (jet-fuel component). The second method uses the synthetic crude heavy fuel oil as feedstock to a two-step UOP processing scheme of hydrocracking, followed by aromatic saturation to yield wide-range jet-B fuel.

2. Process Economics — We originally planned to determine process economics for one process in each of the product categories. The U-GAS and HYGAS Processes were selected for hydrogen and methane. Both processes are oxygen-based, use fluidized beds, and are new IGT processes, thus giving comparability and avoiding possible bias in costing some other firm's process. After the design of the Steam-Iron Process was completed, the sponsor requested that we include a cost estimate for this process in the economics work. For the jet-fuel component, we elected to present economics for the case in which the manufacture of light distillate from petroleum refineries is increased by importing coal-derived aromatic gasoline. This offers an attractive way of incorporating the desirable high-octane, aromatic components of liquids from coal into motor gasoline, while releasing paraffinic distillate material for use as jet fuel. Economics for the second method (jet fuel) became available late in the project.

To estimate process economics, it is necessary to size the various equipment items or sections of the plant to estimate the total plant investment. The project plan stipulates that mid-1974 costs and a private investor financing method be used.

As a contract requirement, measurements have been expressed in the International System of Units (Systeme International d'Unites or SI units). Since most of these units are relatively unfamiliar, and all our calculations were made using the British system, the more familiar British units are shown in parentheses. Table 3 shows relationships between the two systems of units.

Table 3. TABLE OF INTERNATIONAL UNITS AND THEIR CORRESPONDING UNITS IN CONVENTIONAL BRITISH SYSTEM* BASED ON NASA SP-7012 REPORT "THE INTERNATIONAL SYSTEM OF UNITS"

To SI Unit	From British Unit	Multiply by Conversion Factor	SI Units, Symbols
Kelvin	°F	(°F + 459.67) X 5/9	K
Kilonewtons/meter^2	psig	(psig + 14.7) 6.894757	kN/m^2
Kilojoule/kilogram	Btu/lb	2.32517	kJ/kg
Kilogram/hour	lb/h	0.45359237	kg/h
Kilogram-mole/hour	lb-mol/h	0.45359237	kg-mol/h
Metric ton/day	short tons/d	0.9071847	t/d
Metric ton/day	long tons/d	1.016046908	t/d
Gigajoule/hour	10^6 Btu/h	1.05468	GJ/h
Terajoule/day	10^9 Btu/d	1.05468	TJ/d
Meter^3/hour	gpm	0.2271247	m^3/h
Meter^3/day	gal/d	0.00378518	m^3/d
Megajoule/meter^3	Btu/SCF	0.037246	MJ/m^3
Meter^3/day	SCF/d	0.02831685	m^3/d
Kilogram/meter^3	lb-m/ft^3	16.01846	kg/m^3
Kilowatt	hp	0.7457	kW
Meters^3	Petroleum, barrel (42 gal)	0.15898 73	m^3
Gigajoules/meter^3	10^6 Btu/bbl	6.6336	GJ/m^3
Liter	gal	3.7854	Liter
Joule	Btu	1054.68	J

Prefix With SI Units

k	kilo	10^3
M	Mega	10^6
G	Giga	10^9
T	Tera	10^{12}

*Note:

The process design calculations were made using the conventional British units and then converted to the International System of Units (SI) as required by the contract. Since most people are not familiar with the SI units, the quantities expressed in parentheses in the flow sheets, tables, and text are represented in British units. For example, for an SI unit of K, the corresponding number is (°F).

264-TJ/d (250.3 Billion Btu/d) HYDROGEN FROM MONTANA SUBBITUMINOUS COAL BY THE KOPPERS-TOTZEK PROCESS

The Koppers-Totzek Process employs the partial oxidation of pulverized coal in suspension with oxygen and steam. This design is based on the use of Montana subbituminous coal for gasification. The coal composition is given in Table 1. The conversion of coal to hydrogen is a complex process requiring many operations. These operations can be grouped into four major sections:

1. Coal storage and preparation
2. Coal gasification to produce synthesis gas
3. Manufacture of hydrogen from raw synthesis gas
4. Utilities and other offsites.

Figure 1 shows the processing steps and Tables 4 and 5 give the compositions of the solid and gaseous streams. The various sections of the plant are discussed below.

Coal Storage and Preparation

The hydrogen plant is assumed to be located near a coal mine that will provide at least a 25-year supply of coal. Raw coal is brought from the mine by trucks to a storage area where provision is made to accumulate a 30-day supply of coal broken to 3.8 cm x 0 (1-1/2 inch x 0) size in such a way that it is uniformly distributed in the storage pile. This plant requires a continuous flow of 22,690 t/d (25,012 short tons/d) raw coal from the mine (22% moisture). Because of the uniform distribution of coal to the storage piles, the composition of the coal feed to the plant approaches a more uniform condition even though it varies from the mine. About 24.4% of the feed coal is used as boiler and dryer fuel.

In combination grinder-dryer mills, coal is dried to 2% moisture and pulverized to 70% through 200 mesh. The pulverized coal is conveyed to the storage bins before feeding to the gasifiers.

Generation of Synthesis Gas

The pulverized coal is continuously discharged into a mixing nozzle where it is entrained in oxygen and low-pressure steam. The moderate temperature and high burner velocity prevent the reaction of the coal and the oxygen until entry into the gasification zone of the Koppers-Totzek gasifier. The quantities of coal, steam, and oxygen required for the gasifiers are shown below.

Table 4. COMPOSITION OF SOLID STREAMS FOR 264 TJ/d (250.3 X 10^9 Btu/d) HYDROGEN FROM MONTANA SUBBITUMINOUS COAL BY THE KOPPERS-TOTZEK PROCESS

Stream	Ⓐ		Ⓑ		Ⓒ	
Stream Name	Plant Coal Feed		Gasifier Feed		Slag	
Component	wt %	kg/h	wt %	kg/h	wt %	kg/h
C	67.70	499,257	67.70	377,585	--	--
H	4.61	33,997	4.61	25,711	--	--
O	18.46	136,134	18.46	102,958	--	--
N	0.85	6,268	0.85	4,741	--	--
S	0.66	4,867	0.66	3,681	--	--
Ash	7.72	56,931	7.72	43,057	--	--
Subtotal	100.00	737,454	100.00	557,733		21,523
Moisture		208,000		11,382		--
Total		945,454		569,115		21,523
Total, lb/h		2,084,370		1,254,684		47,450

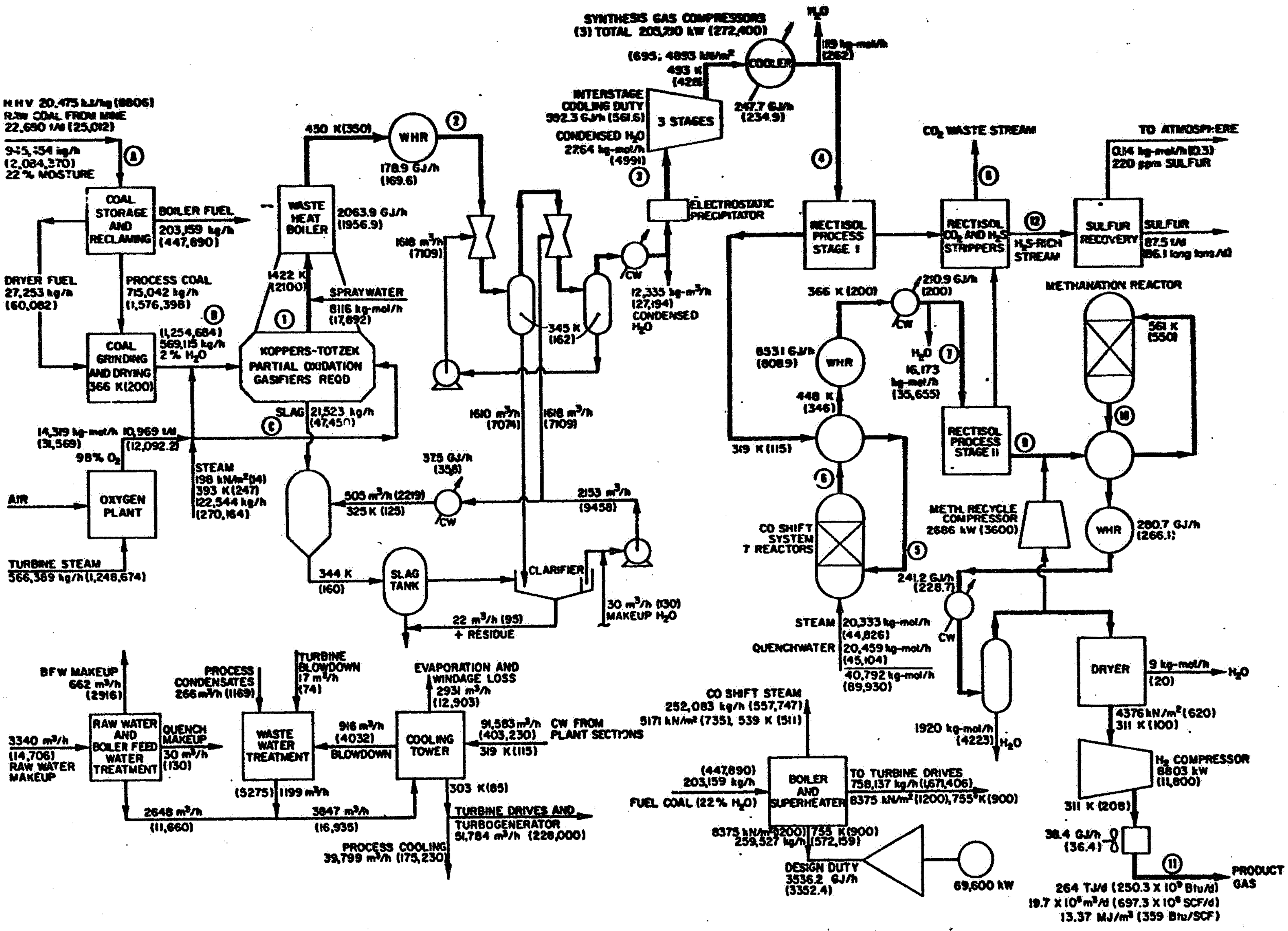

Figure 1. PROCESS FLOW DIAGRAM FOR 264 TJ/D (250.3 X 10⁹ Btu/d) HYDROGEN FROM MONTANA SUBBITUMINOUS COAL BY THE KOPPERS-TOTZEK GASIFICATION (Numbers in Parentheses Have Conventional British Units, See Table 3 for Units and Conversion Factors)

Table 5. PROCESS FLOW STREAMS FOR 264 TJ/d (250.3 X 10^9 Btu/d) HYDROGEN FROM MONTANA SUBBITUMINOUS COAL BY THE KOPPERS-TOTZEK GASIFICATION PROCESS

Stream No.	①		②		③		④		⑤		⑥	
Stream Name	Gasifier Effluent		Gas to Venturi Scrubber		Synthesis Gas Compressor Feed		Rectisol I Feed		CO-Shift Feed		CO-Shift Effluent	
Temperature, K (°F)	1772 (2730)		356 (181)		311 (100)		319 (115)		461 (370)		505 (449)	
Pressure, kN/m² (psig)*	144 (6.2)		141 (5.7)		127 (3.7)		4859 (690)		4824 (685)		4652 (660)	
	mol %	kg-mol/h	mol %	kg-mol/h	mol %	kg-mol/h	mol %	kg-mol/h	mol %	kg-mol/h	mol %	kg-mol/h
Component												
CO	51.28	26,553	44.33	26,553	55.29	26,553	58.17	26,553	58.87	26,553	2.44	2,087
CO_2	8.83	4,574	7.64	4,574	9.52	4,574	10.02	4,574	9.43	4,253	33.58	28,719
H_2	26.75	13,855	23.13	13,855	28.85	13,855	30.35	13,855	30.72	13,855	44.81	38,321
H_2O	12.04	6,234	23.96	14,349	5.16	2,478	0.21	96	--	--	18.65	16,326
CH_4	--	--	--	--	--	--	--	--	--	--	--	--
N_2 + Ar	0.88	455	0.76	455	0.95	455	1.00	443	0.98	443	0.52	443
H_2S + COS	0.22	114	0.19	114	0.24	114	0.25	114	--	--	--	--
O_2	--	--	--	--	--	--	--	--	--	--	--	--
Total	100.00	51,785	100.00	59,900	100.00	48,029	100.00	45,645	100.00	45,104	100.00	85,896
Total, lb-mol/h		114,164		132,056		105,884		100,632		99,437		189,367

Stream No.	⑦		⑧		⑨		⑩		⑪		⑫	
Stream Name	Rectisol II Feed		CO_2 Waste Stream From Rectisol II		Methanation Feed		Methanation Effluent		H_2 Product Gas		H_2S Stream to Sulfur Recovery	
Temperature, K (°F)	319 (115)		319 (115)		319 (115)		755 (900)		333 (140)		319 (115)	
Pressure, kN/m² (psig)*	4617 (655)		108 (1)		4583 (650)		4376 (620)		6996 (1000)		147 (6.6)	
	mol %	kg-mol/h	mol %	kg-mol/h	mol %	kg-mol/h	mol %	kg-mol/h	mol %	kg-mol/h	mol %	kg-mol/h
Component												
CO	2.99	2,087	0.38	124	4.84	1,963	0.10	35	0.10	35	--	--
CO_2	41.19	28,719	89.17	28,717	--	2	49 ppmv	2	50 ppmv	2	71.95	321
H_2	54.96	38,321	0.40	128	94.07	38,193	88.19	32,406	93.07	32,406	--	--
H_2O	0.22	153	--	--	--	--	5.25	1,929	0.01	5	--	--
CH_4	--	--	--	--	--	--	5.25	1,929	5.54	1,929	--	--
N_2 + Ar	0.64	443	10.05	3,236	1.09	443	1.21	443	1.28	443	2.55	11
H_2S + COS	--	--	--	--	--	--	--	--	--	--	25.50	114
O_2	--	--	--	--	--	--	--	--	--	--	--	--
Total	100.00	69,723	100.00	32,205	100.00	40,601	100.00	36,744	100.00	34,820	100.00	446
Total, lb-mol/h		153,712		71,000		89,510		81,006		76,763		984

* kN/m² expressed in absolute units, i.e., 101.33 kilonewtons/meter² = 1 atm = 14.696 psia.

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	kg/h	lb/h
Coal (Dry Basis)	557,733	1,229,590
Steam	122,544	270,164
Oxygen (98% O_2)	10,969 t/d	12,092 short tons/d

The oxygen, steam, and coal react in a refractory-lined, horizontal, cylindrical vessel with conical ends at a slight positive pressure and at 2089 K (3300°F). The fixed carbon and volatile matter of the coal are gasified to produce raw synthesis gas and molten slag at 1772 K (2730°F). About 50% of the molten slag drops into a water-filled quench pot, forming a slag-H_2O slurry at 366 K (200°F). This slurry is cooled to 325 K (125°F) and sent to a slag-settling pond. The water is recycled to the slag quench pot. The rest of the slagged wet ash is carried out as fly ash with the gas and, after waste-heat recovery from the hot gas, recovered by water scrubbing and electrostatic precipitation. The composition of the raw gas from the gasifiers is as follows:

	Actual	Dry Basis
	mol %	
CO	51.3	58.3
CO_2	8.8	10.0
H_2	26.8	30.4
H_2O	12.0	--
CH_4	--	--
N_2 + Ar	0.9	1.0
H_2S + COS	0.2	0.3
Total	100.0	100.0

Pressure = 144 kN/m^2 (6.2 psig)
Temperature = 1772 K (2730°F)

Manufacture of Hydrogen From Raw Synthesis Gas

Synthesis gas leaving the gasifiers is cooled to 1422 K (2100°F) by direct quench with water that helps solidify the entrained slag droplets. The quenched gas is cooled to 355 K (180°F) in the waste-heat recovery boiler, where a significant amount of superheated steam at 8375 kN/m^2 (1200 psig) and 755 K (900°F) for turbine drives is generated. The cooled gas is washed

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with methanol already charged with CO_2 at 319 K (115°F) and 4859 kN/m^2 (690 psig). The higher operating pressure of the Rectisol system is favorable for the physical absorption of acid-gas in methanol. The regeneration of this solvent yields an H_2S-rich gas containing 25.5% H_2S, which yields 87.5 t/d (86.1 long tons/d) of sulfur in the sulfur recovery plant, and 250 ppmv of sulfur in the vent gas to the atmosphere.

After the CO shift, the effluent is cooled to 319 K (115°F) by waste-heat recovery and cooling water before going to the second-stage Rectisol system for CO_2 removal. The CO_2-rich stream is vented to the atmosphere since it contains less than 6 ppmv sulfur.

To reduce the CO to 0.1%, the effluent from the second stage Rectisol containing 4.8% CO is methanated. A single-stage recycle quench methanation system with a feed temperature of 561 K (550°F) and an effluent temperature of 755 K (900°F) is used. The methanation effluent is cooled to 325 K (125°F), with a portion of the gas used as a recycle quench stream. The product gas is dried in a standard glycol drying unit to 112 kg H_2O/Mm^3 (7 lb H_2O/million SCF gas).

The dried gas is compressed to 6996 kN/m^2 (1000 psig) in a single-stage product gas compressor and sent to the pipeline. As the heating value of hydrogen is approximately one-third that of methane (for each SCF of gas), the volume of H_2 product gas is approximately three times that of SNG for a comparable 264 TJ/d (250.3 X 10^9 Btu/d) plant. The composition of the H_2 product gas is given below:

	mol %
CO	0.1
CO_2	<50 ppm
H_2	93.1
CH_4	5.5
N_2 + Ar	1.3
Total	100.0

Pressure = 6996 kN/m^2 (1000 psig)
Temperature = 333 K (140°F)

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Utilities and Other Offsites

The conversion of coal to hydrogen requires large quantities of utilities like steam power and cooling water, which are summarized in Tables 6-8. The plant is designed to be self-sufficient, based on coal. The boiler duty is 3444.3 GJ/h (3265.7 X 10^6 Btu/h). It is designed to supply 5171 kN/m^2 (735 psig) saturated steam for the CO shift and 8375 kN/m^2 (1200 psig) steam for the plant power requirements.

Table 9 presents the overall process thermal efficiency. Approximately 56.8% of the higher heating value of feed coal is converted to higher heating value in the product gas; 0.2% goes to the sulfur product. The overall energy balance is given in Table 10.

In summary, 264 TJ/d (250.3 X 10^9 Btu/d) of H_2 produced from Montana subbituminous coal using Koppers-Totzek gasifiers operating at 143 to 150 kN/m^2 (6 to 7 psig) and 1755 to 2089 K (2700° to 3300°F) requires 22,690 t/d (25,012 short tons/d) of 22% moisture coal and converts 56.8% of the HHV of coal to the HHV of the H_2 product and 0.2% of the HHV to sulfur.

Table 6. SUMMARY OF STEAM REQUIREMENTS FOR 264 TJ/d (250.3 X 10^9 Btu/d) HYDROGEN FROM MONTANA SUBBITUMINOUS COAL BY THE KOPPERS-TOTZEK PROCESS.

Steam Required	kg/h	lb/h	GJ/h	10^6 Btu/h
High Pressure 8378 kN/m^2 (1200) Turbine Steam 755K (900)	1,667,089	3,675,297	5220.8	4950.3
High Pressure 5171 kN/m^2 (735) Saturated CO Shift Steam	366,315	807,585	998.8	947.1
Subtotal	2,033,404	4,482,882	6219.6	5897.4
Low Pressure				
Gasifier 198 kN/m^2 (14), 393K (247)	122,545	270,164	326.8	309.9
Rectisol 446 kN/m^2 (50), 421K (298)	40,572	89,447	86.0	81.5
Waste-water Treatment 446 kN/m^2 (50), 421K (298)	54,431	120,000	115.3	109.3
Deaerator 143 kN/m^2 (6), 383K (230)	119,094	262,557	252.3	239.2
Subtotal	336,642	742,168	780.3	739.9
Total	2,370,046	5,225,050	6999.9	6637.3
Steam Generated				
From Waste Heat Recovery	--	--	402.0	381.2
Preheat Boiler Feed Water	145,069	319,822	328.4	311.4
Low-Pressure Jacket Steam	191,573	422,346	432.3	409.9
Low-Pressure Steam from CO Shift Waste Heat	114,232	251,838	266.6	252.8
High-Pressure Steam from Methanation	678,621	1,496,100	2126.5	2016.3
High-Pressure Steam from Waste Heat Boiler				
Subtotal	1,129,495	2,490,106	3555.8	3371.8
From Boiler Superheater	252,083	555,747	587.6	557.2
From Boiler Superheater	988,468	2,179,197	2856.5	2708.5
Total	2,370,046	5,225,050	6999.9	6637.3

Table 7. SUMMARY OF POWER REQUIREMENTS FOR 264 TJ/d (250.3 X 10^9 Btu/d) HYDROGEN FROM MONTANA SUBBITUMINOUS COAL BY THE KOPPERS-TOTZEK PROCESS

Component	hp	kW
Coal Drying and Reclaiming	3,000	2,238
Coal Crushing	8,837	6,592
Air Separation Plant	199,017*	148,467*
Oxygen Compressor	4,100*	3,059*
Synthesis Gas Compressor	272,400*	203,210*
Rectisol Refrigeration	24,000*	17,904*
Rectisol Motive Power	24,000	17,904
Methanator Recycle Gas Compressor	3,600*	2,686*
H_2 Compressor	11,800*	8,803*
Cooling Water Pumps and Fans	27,000	20,142
Boiler Feed Water Pumps	8,000	5,968
Waste-Water Treatment	2,000	1,492
Other Miscellaneous	10,000	7,460
Total	597,754	445,925
Power Generation		
*From Steam Driven Turbines	514,917	384,129
Net from Turbo Generator	82,837	61,796
Total	597,754	445,925

Table 8 . SUMMARY OF COOLING WATER REQUIREMENTS FOR 264 TJ/d (250.3 X 10^9 Btu/d) HYDROGEN FROM MONTANA SUBBITUMINOUS COAL BY THE KOPPERS-TOTZEK PROCESS

Component	m^3/h*	gpm*
Gas Quenching and Scrubbing System	8,829	38,873
Air Separation Plant (In Addition to Steam Condensors)	10,547	46,437
Rectisol System	1,000	4,400
Interstage Cooling in Syngas Compressor	12,060	53,100
Slag Cooling	882	3,884
CO-Shift System	3,021	13,300
Methanation System	3,460	15,234
For Steam Turbine Condensors	51,784	228,000
Total	91,583	403,228

* Cooling water temperature 303°-319°K (85°-115°F).

Table 9. KOPPERS-TOTZEK PROCESS EFFICIENCY FOR 264 TJ/d (250.3 X 10^9 Btu/d) HYDROGEN PLANT USING MONTANA SUBBITUMINOUS COAL

	kg/h	lb/h
Reactor Coal (Dry Basis)	557,733	1,229.590
Fuel Coal (Dry Basis)	179,721	396,218
Total	737,454	1,625,808
	GJ/h	10^6 Btu/h
HHV Total Coal*	19,359.1	18,355.4
HHV Product Gas	10,999.0	10,429.0
% Converted to Product Gas	56.8	
HHV Sulfur	37.0	32.0
% Converted to Sulfur	0.2	
% Converted to Products	57.0	

* At 26,251 kJ/kg (11,290 Btu/lb) dry basis.

Table 10 . SUMMARY OF OVERALL ENERGY BALANCE FOR 264 TJ/d (250.3 X 10^9 Btu/d) HYDROGEN FROM MONTANA SUBBITUMINOUS COAL BY THE KOPPERS-TOTZEK PROCESS
(Basis: 289 K (60 °F) Liquid H_2O)

Heat In	GJ/h	10^6 Btu/h	%
Coal to Gasifier	14,641.2	13,882.1	75.6
Coal to Boiler and Dryer	4,717.9	4,473.3	24.4
Total	19,359.1	18,355.4	100.0
Heat Out			
Hydrogen-Rich Gas	10,997.8	10,427.6	56.8
Sulfur	33.7	32.0	0.2
Heat Lost Through Cooling Water	6,447.5	6,113.2	33.3
Air Cooling	61.5	58.3	0.3
Coal Drying	558.0	529.1	2.9
Boiler Stack Gases, etc. (85% efficiency)	624.1	591.7	3.2
Heat Losses Through Hot Condensate and Deaerator Losses	90.3	85.6	0.5
Heat Recovery Losses	208.3	197.5	1.1
Other Miscellaneous Unaccounted	338.0	320.4	1.7
Total	19,359.1	18,355.4	100.0

PROCESS DESCRIPTION OF 263.9 TJ/D (250 Billion Btu/d) HYDROGEN PLANT FROM MONTANA SUBBITUMINOUS COAL BY THE U-GAS PROCESS

The U-GAS Process utilizes single-stage fluidized-bed gasifiers. The major advantages of this type of operation include —

1. High reaction rates are achieved because of good gas-solids contact.
2. The bed temperature is uniform and can be easily controlled.
3. The high mass of carbon in the fluid bed assures reducing conditions, giving good product gas and ensures the conversion of sulfur to H_2S, which is readily removed.
4. The ash removal system is unique because it allows recycling and subsequent gasification of fines and removal of only low-carbon ash. This system results in a high-carbon conversion and, therefore, higher efficiency.

Our process designs for hydrogen are based on Montana subbituminous coal, which is noncaking, and so pretreatment is unnecessary. The composition of this coal is shown in Table 1.

For the description that follows, refer to the flow diagram (Figure 2) and the material balances in Tables 11 and 12. Coal containing 22% moisture is received from the nearby mining area at the rate of 19,453 t/d (21,443 short tons/d). After breaking to 3.8 cm x 0 (1-1/2 in. x 0), the coal is placed in the storage-reclaiming area where 30 days of storage is maintained. Provisions are made to store the coal in such a way that it is uniformly distributed in the storage pile. Because of this provision, the reclaimed coal fed to the gasifiers approaches a uniform composition even though the composition of coal varies from the mine. In addition to 15,204 t/d (16,760 short tons/d) of process coal (22% moisture) for the gasifiers, 485 t/d (535 short tons/d) of wet coal (22% moisture) are required for the coal-drying system; the boiler requires 3763 t/d (4148 short tons/d) of coal (22% moisture).

Coal is simultaneously crushed to 0.6 cm x 0 (1/4 in. x 0) and dried to 10% moisture before being conveyed to the reactor feed system. Lockhoppers have been chosen for this feed system because they have been commercially proved in systems operating at this reactor pressure, 2411 kN/m^2 (335 psig). Coal is dropped from a feed bin into an open lockhopper. After being isolated by the lockhopper valves, the lockhopper is pressurized to the reactor pressure with carbon dioxide from a recycle compressor. The lockhopper discharge valve is then opened and the contents flow by gravity into

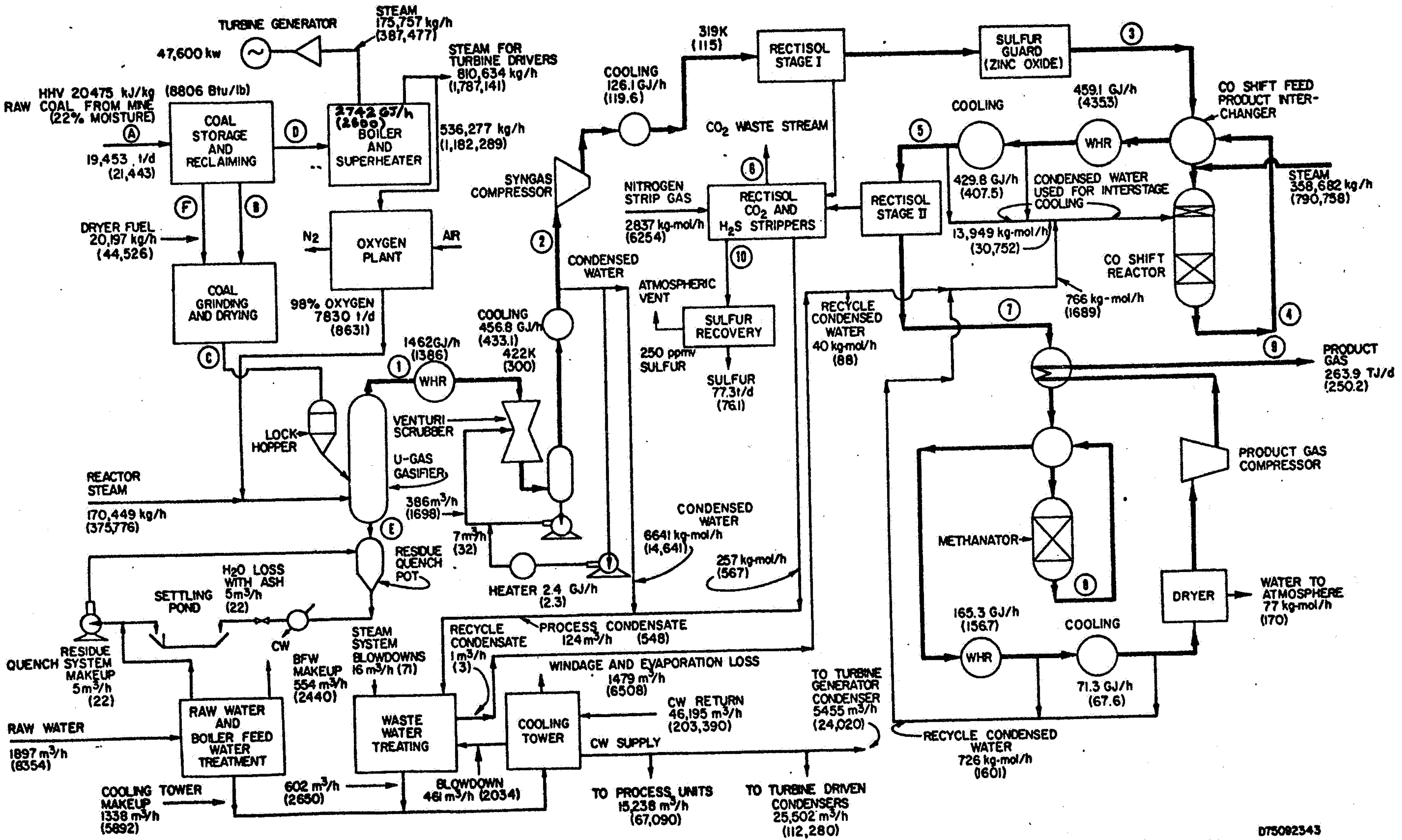

Fig. 2. PROCESS FLOW DIAGRAM FOR 263.9 TJ/d (250.2 X 10⁹ Btu/d) HYDROGEN FROM MONTANA SUBBITUMINOUS COAL BY THE U-GAS PROCESS (NUMBERS IN PARENTHESIS HAVE CONVENTIONAL BRITISH UNITS . SEE TABLE 3 FOR UNITS AND CONVERSION FACTORS)

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Table 11. COMPOSITION OF SOLID STREAMS FOR 263.9 TJ/d (250.2 X 10^9 Btu/d) HYDROGEN FROM MONTANA SUBBITUMINOUS COAL BY THE U-GAS PROCESS

Stream	Ⓐ		Ⓑ		Ⓒ		Ⓓ		Ⓔ		Ⓕ	
Stream Description	Coal From Mine		Process Coal		Dried Coal to Gasifier		Boiler Fuel		Gasifier Residue		Dryer Fuel	
	wt %	kg/h	wt %	kg/h	wt %	kg/h	wt %	kg/h	wt %	kg/h	wt %	kg/h
Component												
C	67.70	427,999	67.70	334,535	67.70	334,535	67.70	82,800	14.53	6,739	67.70	10,665
H	4.61	29,144	4.61	22,780	4.61	22,780	4.61	5,638	0.17	77	4.61	726
O	18.46	116,704	18.46	91,219	18.46	91,219	18.46	22,577	--	--	18.46	2,908
N	0.85	5,374	0.85	4,200	0.85	4,200	0.85	1,040	2.99	1,385	0.85	134
S	0.66	4,173	0.66	3,261	0.66	3,261	0.66	807	0.06	29	0.66	104
Ash	7.72	48,806	7.72	38,148	7.72	38,148	7.72	9,442	82.25	38,148	7.72	1,216
Subtotal	100.00	632,200	100.00	494,143	100.00	494,143	100.00	122,304	100.00	46,378	100.00	15,753
Water	--	178,313	--	139,374	--	54,905	--	34,496	--	--	--	4,443
Total	--	810,513	---	633,517	--	549,048	--	156,800	--	46,378	--	20,196
Total lb/h		1,786,875		1,396,665		1,210,443		345,684		102,245		44,526

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Table 12. PROCESS STREAMS FOR 263.9 TJ/d (250.2 X 10^9 Btu/d) HYDROGEN FROM MONTANA SUBBITUMINOUS COAL BY THE U-GAS PROCESS

Stream No.	①		②		③		④		⑤		⑥	
Stream Name	Gasifier Product		Synthesis Gas Compressor Feed		CO Shift Dry Feed		CO Shift Effluent		Rectisol Stage II Feed		CO_2 Waste from Rectisol CO_2 Stripping	
Temperature, K (°F)	1311	(1900)	311	(100)	319	(115)	511	(460)	319	(115)	319	(115)
Pressure, kN/m² (psig)	2413	(335)	2379	(330)	4826	(685)	4654	(660)	4619	(655)	103	(0.4)
	mol %	kg-mol/h	mol %	kg-mol/h	mol %	kg-mol/h	mol %	kg-mol/h	mol %	kg-mol/h	mol %	kg-mol/h
Component												
CO	43.31	21,481	50.01	21,481	50.62	21,481	1.23	945	1.55	945	0.38	107
CO_2	9.93	4,927	11.47	4,927	10.93	4,639	32.67	25,175	41.19	25,175	89.16	25,173
H_2	30.52	15,139	35.24	15,139	35.67	15,139	46.29	35,657	55.10	35,675	0.40	113
H_2O	13.63	6,759	0.28	118	--	----	18.28	14,088	0.23	139	---	------
CH_4	1.79	885	2.06	885	2.09	885	1.15	885	1.45	885	0.01	2
H_2S	0.20	101	0.24	101	0.1 ppm	--	--	---	--	--	---	------
N_2	0.20	101	0.23	101	0.21	90	0.12	90	0.15	90	10.05	2,837
Ar	0.42	203	0.47	203	0.48	203	0.26	203	0.33	203	---	------
Total	100.00	49,596	100.00	42,955	100.00	42,437	100.00	77,061	100.00	63,112	100.00	28,232
Total, lb-mol/h		(109,340)		(94,699)		(93,588)		(169,891)		(139,139)		(62,242)

Stream No.	⑦		⑧		⑨		⑩	
Stream Name	Methanation Feed		Methanation Effluent		Hydrogen Product Gas		H_2S Rich Stream to Sulfur Recovery	
Temperature, °K (°F)	319	(115)	719	(835)	333	(140)	319	(115)
Pressure, kN/m² (psig)	4516	(640)	4482	(635)	6998	(1000)	147	(6.6)
	mol %	kg-mol/h	mol %	kg-mol/h	mol %	kg-mol/h	mol %	kg-mol/h
Component								
CO	2.23	838	0.10	34	0.1	34	---	----
CO_2	50 ppm	2	<50 ppm	---	<50 ppm	----	72.00	288
H_2	94.64	35,562	92.15	33,145	94.25	33,145	---	----
H_2O	---	--	2.25	807	0.01	4	---	----
CH_4	2.35	883	4.69	1,688	4.80	1,688	---	----
H_2S	---	--	--	--	--	----	25.25	101
N_2	0.24	90	0.25	90	0.26	90	2.75	11
Ar	0.54	203	0.36	203	0.58	203	---	----
Total	100.00	37,578	100.00	35,967	100.00	35,164	100.00	400
Total, lb. mol/h		(82,845)		(79,295)		(77,524)		(882)

• kN/m² expressed in absolute units; i.e. 101.33 Kilo Newtons/meter² = 1 atm = 14.696 psia

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a continuous reactor feed bin. The lockhopper is then depressurized into a gas holder upstream of the recycle compressor; this cycle is continuously repeated.

Generation of Synthesis Gas

This plant design requires two gasifiers with a 9.4-m (31-ft) ID and a 20.9-m (68.5-ft) straight shell. These two units consume 7830 t/d (8631 short tons/d) of 98% oxygen combined with 170,449 kg/h (375,776 lb/h) of steam. The steam and oxygen serve as a fluidizing-gasifying medium for the 13,177 t/d (14,525 short tons/d) of gasifier coal feed. The gasifier feed quantities and steam required for CO-shift conversion are given below.

U-GAS GASIFIER FEED QUANTITIES
[For 264 TJ/d (250 X 10^9 Btu/d) Plant Utilizing Montana Subbituminous Coal]

Coal, kg/h (dry basis)	494,143 (1,089,399 lb/h)
Steam, kg/h	170,449 (375,776 lb/h)
Oxygen, t/d	7,830 (8,631 short tons/d)

Steam to Shift, kg/h
358,682 (790,758 lb/h)

The coal is gasified in a single-stage fluidized bed at a temperature of 1311 K (1900°F) and a pressure of 2413 kN/m^2 (335 psig). The reactor residence time is 80 minutes and the fluidizing velocity is 0.46 m/s (1.5 ft/s). Because the coal is injected below the fluidized-bed surface, methane formation is minimized; volatiles are decomposed because of the high temperature, 1311 K (1900°F). This gasifier concept is unique for its method of ash removal and its fines handling system. The design resolves the main disadvantage of coal gasification in a fluidized bed rich in carbon, the problem of selectively removing low-carbon content ash from the fluidized bed. At the same time as coal gasification, the ash is agglomerated into larger and heavier particles for selective separation from the bed. The method for removing the ash in this manner is described below. Part of the fluidizing steam-oxygen mixture enters the gasifier through a fluidizing grid that is sloped toward one or more inverted cones contained in the grid. The remaining gas flows upward at a high velocity through a throat at the cone apex. This

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creates a small, higher-temperature area above the cone within the fluidized bed. By properly selecting the ratio of oxygen to steam in the gas fed to the throat of the cone, the temperature above the cone throat can be controlled so that it is greater than that of the rest of the fluidized bed and near the ash-softening point for the coal. Because the ash in gasified coal particles is heated to near its softening point, the surface of the particles becomes sticky and the particles agglomerate. Ash agglomerates grow in size in the high-velocity, oxygen-rich jet until they are heavy enough to counter the drag force of the high-velocity gas stream from the throat of the inverted cone; at this point, they fall through the cone throat and out of the fluidized bed. Both the fluidized gasifier and ash agglomeration are being studied in a small pilot plant at IGT.

Fines (caused by attrition or from the previous crushing operation) preferentially carried upward and out (elutriated) of the fluidized bed are returned to the gasifier through cyclones. An internal cyclone leads directly to the fluidized bed and an external cyclone leads to the bottom of the gasifier. The fines from the external cyclone are recycled once to extinction. They are returned to the fluidized bed in a stream of steam and oxygen, rapidly gasified, and the ash is agglomerated with the normal bed-produced ash in the high-temperature zone above the inverted cone. The ash agglomerates fall into a water-filled quench pot where they form a slurry that is cooled and then depressurized across a valve before being sent to one of two slurry-settling ponds. The water is recycled to the slurry quench pot. The partially dried ash is reclaimed from the unused pond and disposed of in the mining area.

The raw gasifier product at 1311 K (1900°F) is cooled to 422 K (300°F) in a waste-heat recovery boiler that generates about 90% of the process steam requirements; the remaining steam is generated in the CO-shift waste-heat recovery unit.

Small dust particles (less than 5 microns) carried over from the cyclones are taken out in a venturi scrubber that has a high removal efficiency on particles of 1 to 2 microns. This step is necessary to completely remove particulate matter from the gas before compression.

Manufacture of Hydrogen From Synthesis Gas

A comparison of raw synthesis gas with the product gas is given below.

U-GAS HYDROGEN FROM COAL PLANT
(Montana Subbituminous Coal)

	Gasifier Raw Gas Composition	Product Gas Composition
	mol % (Dry Basis)	
CO	50.1	0.1
CO_2	11.5	<50.0 ppm
H_2	35.3	94.3
CH_4	2.1	4.8
N_2 + Ar	0.7	0.8
H_2S	0.3	--
Total	100.0	100.0

To increase the hydrogen yield, the CO is converted by the well-known shift reaction ($CO + H_2O \rightarrow CO_2 + H_2O$), which is carried out in a catalytic reactor. Commercial catalysts are available for operation at two temperature levels, 450 to 533 K (350° to 500°F) and 589 to 783 K (600° to 950°F). The low-temperature shift catalyst requires much less steam than a high-temperature shift system because of the more favorable equilibrium. In this case, 20,536 kg-mole/h (45,275 lb-mole/h) of CO are shifted. The low-temperature shift system saves 0.5 to 0.9 million kg/h (1.2 to 2 million lb/h) of shift steam, depending on whether the system used as a comparison is:

1. An exclusively high-temperature shift catalyst with no liquid water quench between stages, or
2. A combination high-temperature/low-temperature shift catalyst with liquid water quench between stages.

Also, because the low-temperature shift system decreases steam usage, the cost of the boiler is less, and, more importantly, the plant efficiency is raised 3% to 7%, again depending on the high-temperature shift design used as a comparison.

Waste-Water Treatment

The condensed water from the process contains H_2S, and is sent to a steam stripper, where the H_2S is removed and sent to the sulfur recovery unit. The processed water is then recycled to the CO-shift unit as quench and the remainder is used as makeup for the cooling tower.

Utilities and Other Offsite Facilities

The required utilities — steam, power, and cooling water — are summarized in Tables 13-15. Waste-heat recovery is used to the maximum extent practicable to generate steam and preheat the boiler feedwater. This cuts down on both boiler size and coal consumption. The power for the plant is derived from a coal-fired boiler that generates steam for process turbine drivers as well as steam for a turbine generator to supply the electrical power needs, 47,600 kW. This design makes the plant self-sufficient, based on the coal feed. Raw water is treated as required for cooling tower makeup and boiler feedwater makeup.

Process Efficiency and Plant Energy Balance

The process efficiency calculation is shown in Table 16. About 66.2% of the high-heating value in the coal is converted to high-heating value in the product gas. Waste-heat recovery units have been used to the maximum extent practicable to reduce the amount of heat lost to the cooling media. The total high-heating value of coal fed into the process is 16,596 GJ/h (15,736 million Btu/h), of which 2080 GJ/h (1972 million Btu/h), or 12.5%, is recovered with waste-heat recovery units. The product gas contains 10,995 GJ/h (10,425 million Btu/h) HHV, which is 66.2% of the HHV of the total coal.

The plant energy balance is shown in Table 17, which gives an accounting for the remaining heat input from coal.

In conclusion, the U-GAS Process utilizing a single-stage fluidized-bed gasifier operating at 2411 kN/m^2 (335 psig) and 1311 K (1900°F) produces 77.3 t/d (76.1 short tons/d) of sulfur and 264 TJ/d (250 billion Btu/d) of hydrogen from Montana subbituminous coal in an environmentally acceptable manner. The efficiency of conversion of total coal HHV to product gas and sulfur HHV is 66.4%.

Table 13. SUMMARY OF STEAM REQUIREMENTS FOR CONVERSION OF MONTANA SUBBITUMINOUS COAL TO 263.9 TJ/d (250.2 X 10^9 Btu/d) HYDROGEN BY THE U-GAS PROCESS

Steam Balances

	Steam Required		Steam Generated				
	50 psig, saturated						
	kg/h	lb/h	Sources	kg/h	lb/h	GJ/h	10^6 Btu/h
Recycled Condensate Heating	1,344	2,962	CO-Shift Waste Heat Recovery Unit (E-5)	70,789	156,063	160.7	152.4
H_2S Regenerator	40,801	89,951					
Waste-Water Stripper	28,644	63,150					
Total	70,789	156,063					
	385 psig - 800° F						
Gasifiers	170,451	375,780	Raw Gas Waste Heat Recovery Unit (E-2) (Superheating done in boiler)	170,451	375,780	481.3	456.3
	730 psig, saturated						
CO-Shift Steam	358,682	790,760	Raw Gas Waste Heat Recovery Unit (E-1)	377,652	832,581	877.9	832.4
Product Gas Dryer	18,970	41,821					
Total	377,652	832,581					
	1200 psig - 900° F						
Turbine Drivers	810,633	1,787,141	Boiler	986,390	2,174,618	2645.6	2508.4
Turbine Generator	175,757	387,477					
Total	986,390	2,174,618					

Table 14. SUMMARY OF POWER REQUIREMENTS FOR CONVERSION OF MONTANA SUBBITUMINOUS COAL TO 263.9 TJ/d (250 Btu/d) HYDROGEN BY THE U-GAS PROCESS

Section	Power Required	
Process Stage	kW	hp
Coal Storage and Reclaiming	1,940	2,600
Coal Grinding and Drying	5,789	7,760
Reactor Feed System*	2,984	4,000
Reactor Discharge System	112	150
Gas Cooling and Dust Removal	433	580
Syn-Gas Compressors*	33,152	44,440
Rectisol H_2S-CO_2 Removal		
a Motive Power	17,307	23,200
b) Refrigeration Units*	17,009	22,800
CO Shift	119	160
Methanation and Drying	52	70
Product Gas Compressors*	17,889	23,980
Oxygen Plant		
Air Compressors (4 required)*	105,775	141,790
Oxygen Compressors (4 required)*	35,741	47,910
Sulfur Recovery	82	110
Wastewater Treatment	366	490
Feedwater Pumps for Steam Turbine Drivers	3,193	4,280
BFW Feedwater Pumps for Turbine Generator	709	950
Cooling Water Pumps	10,011	13,420
Miscellaneous	7,460	10,000
Total	260,123	348,690
Steam Turbine Drivers		
Reactor Feed System	2,984	4,000
Syn-Gas Compressors	33,152	44,440
H_2S-CO_2 Refrigeration Compressors	17,009	22,800
Product Gas Compressors	17,889	23,980
Oxygen Plant		
Air Compressors	105,775	141,790
Oxygen Compressors	35,741	47,910
Subtotal	212,550	284,920
Power From Turbogenerator	47,573	63,770
Total	260,123	348,690

* Steam Turbine Drive.

Table 15. SUMMARY OF COOLING WATER REQUIREMENTS FOR THE CONVERSION OF MONTANA SUBBITUMINOUS COAL TO 263.9 TJ/d (250.2×10^9 Btu/d) HYDROGEN BY THE U-GAS PROCESS

Cooling Water	m^3/h	gpm
Reactor Feed System	186	820
Reactor Discharge System	522	2,300
Gas Cooling and Dust Removal	1,211	5,330
CO Conversion	513	2,260
Methanation	438	1,930
Rectisol	3,300	14,530
Oxygen Plant	8,847	38,950
Sulfur Recovery	86	380
Waste-Water Treating	134	590
Steam Turbine Driven Condensers	25,502	112,280
Turbine Generator Condenser	5,456	24,020
Total	46,195	203,390

Table 16. U-GAS PROCESS EFFICIENCY

Reactor Coal, kg/h (dry basis)	494,143	(1,089,399 lb/h)
Boiler Coal and Dryer Fuel, kg/h (dry basis)	138,057	(304,363 lb/h)
Total Coal	632,200	(1,393,762 lb/h)
HHV Total Coal, GJ/h*	16,596	(15,736 10^6 Btu/h)
HHV Product Gas, GJ/h	10,995	(10,425 10^6 Btu/h)
HHV Product Sulfur, GJ/h	30	(28.3×10^6 Btu/h)
% Converted to Product Gas =	66.2	
% Converted to Product Gas and Sulfur =	66.4	

* At 26,251 kJ/kg (11,290 Btu/lb).

The low-temperature shift catalyst cannot be used with sulfur-containing gas. Moreover, since acid-gases (H_2S and CO_2) must be removed at some stage of the process, it is desirable to remove H_2S prior to CO conversion and CO_2 after CO conversion.

Prior to H_2S removal, the gas is compressed to 4928 kN/m^2 (700 psig). This facilitates acid-gas removal because a physical absorption system has been chosen for the plant — higher partial pressures are favorable for absorption. The unit selected for acid-gas removal is a two-stage Rectisol plant similar to units that have been successful in commercial operation. The H_2S absorption system produces an H_2S-rich gas that is sent to the sulfur-recovery unit. This unit produces 77.3 t/d (76.1 long tons/d) of molten sulfur and has an atmospheric vent containing 250 ppmv sulfur.

After CO conversion, the CO_2 is removed by the second stage of the Rectisol unit. Some of this CO_2 stream is used as lockhopper gas; the remainder is vented to the atmosphere. The sulfur content is claimed to be less than 5 ppmv. A methanation unit is used to reduce the product gas CO concentration from 1.5% to 0.1%. This is more economical than converting all the CO by the shift reaction. The unit is a single-stage adiabatic reactor similar to those used in ammonia plants to remove carbon oxides. Water in the gas from the methanator is removed in a standard glycol drying unit. The dried gas is compressed to 6996 kN/m^2 (1000 psig) by the product gas compressor and enters the pipeline.

In terms of volumes of gas handled, this plant is larger than the standard 264 TJ/d (250 billion Btu/d) SNG plant. This plant produces 20 Mm^3/d (705 million SCF/d) of gas compared to about 7.1 Mm^3/d (250 million SCF/d) of gas for an SNG plant.

Table 17. OVERALL ENERGY BALANCE FOR CONVERSION OF MONTANA SUBBITUMINOUS COAL TO 263.9 TJ/d (250.2 X 10^9 Btu/d) HYDROGEN BY THE U-GAS PROCESS

[Basis: 289 K (60°F) Liquid H_2O]

	GJ/h	10^6 Btu/h
Input		
Gasifier Coal	12,971.2	12,299.3
Boiler and Dryer Fuel Coal	3,624.2	3,436.3
Total	16,595.4	15,735.6
Output		
Product Gas, HHV	10,994.8	10,424.8
Product Gas Enthalpy	45.8	43.4
By-Product Sulfur	29.8	28.3
By-Product Sulfur Enthalpy	11.5	10.9
Residue, HHV	232.1	220.0
Residue Enthalpy	1.6	1.5
Air-Cooling Heat Rejection	947.9	898.8
Cooling-Tower Heat Rejection	3,217.7	3,050.9
Boiler-Stack U-GAS Losses	481.6	456.6
Steam System Blow Down Cooling	17.5	16.6
CO_2 Removal System Atmospheric	64.5	61.2
CO_2 Removal System Atmospheric Vent-HHV Enthalpy	31.7	30.1
Coal-Drying System Vent Enthalpy	334.1	316.8
Product Gas Dryer Vent Enthalpy	3.6	3.4
Deaerator Vent Enthalpy	21.3	20.2
Waste-Heat Recovery Unit Losses	109.2	103.5
Miscellaneous and Unaccounted Losses	51.3	48.6
Total	16,596.0	15,735.6

263.9 TJ/d (250.2 Billion Btu/d) HYDROGEN FROM MONTANA SUBBITUMINOUS COAL BY THE STEAM-IRON PROCESS

In the Steam-Iron Process, the crushed and dried coal is reacted with steam and air to make producer gas that is used to reduce oxidized iron from the steam-iron reactor. Hydrogen is derived from water in accordance with reactions given below. The advantages of this process include —

1. High-purity hydrogen is produced.
2. No oxygen plant is required because air is used directly.
3. No CO shift is required.
4. A large amount of electric power is produced as a by-product.

Any type of coal, from bituminous to lignite, can be gasified in this process. For the caking coals, pretreatment with air is required at 644 to 700 K (700° to 800°F) to reduce the agglomerating tendency of coal.

Montana subbituminous coal was used for the gasification in this design, so no pretreatment is necessary. The composition of raw coal is given in Table 1. The processing steps required in the conversion of coal to hydrogen are shown in Figure 3. Tables 18 and 19 give the composition of solid and gaseous streams corresponding to the streams in Figure 3. The processing steps may be grouped in the following way:

1. Coal storage and preparation
2. Producer gas generator and steam-iron reactor
3. Upgrading of oxidizer effluent to H_2 product
4. Power generation from reductor off-gas using combined power cycle
5. Utilities and other offsites.

The description of these steps follows.

1. Coal Storage and Preparation

The hydrogen plant is assumed to be located near a coal mine that will provide a minimum 25-year supply of coal. Raw coal is brought from the mine by trucks to a storage area where provision is made to store a 30-day supply of coal broken to 3.8 cm x 0 (1-1/2 in. x 0) size. Coal is stored in such a way that it is uniformly distributed in the storage pile. This plant requires a continuous flow of 28,896 t/d (30,853 short tons/d) of raw coal from the mine (22% moisture). Because of the uniform distribution of coal to the storage piles, the composition of the coal feed to the plant becomes

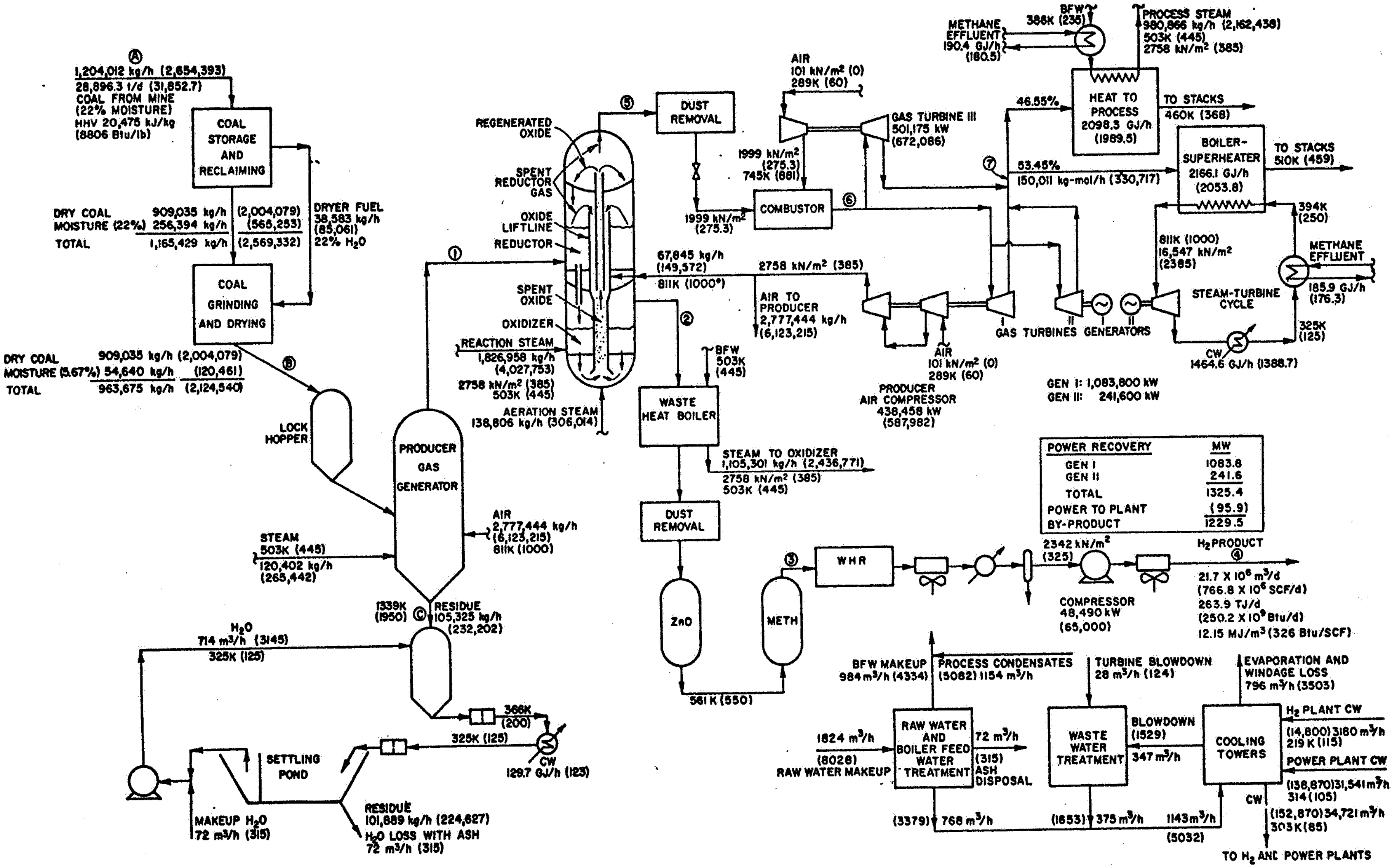

Fig. 3. PROCESS FLOW DIAGRAM FOR 263.9 TJ/d (250.2 X 10⁹ Btu/d) HYDROGEN FROM MONTANA SUBBITUMINOUS COAL BY THE STEAM-IRON PROCESS (NUMBERS IN PARENTHESIS HAVE CONVENTIONAL BRITISH UNITS. SEE TABLE 3 FOR UNITS AND CONVERSION FACTORS)

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Table 18 . COMPOSITION OF SOLID STREAMS FOR 263.9 TJ/d (250.2 X 10^9 Btu/d) HYDROGEN FROM MONTANA SUBBITUMINOUS COAL BY THE STEAM-IRON PROCESS

Stream	Ⓐ		Ⓑ		Ⓒ	
Stream Name	Plant Coal Feed		Coal-to-Gas Producer		Residue	
Component	wt %	kg/h	wt %	kg/h	wt %	kg/h
C	67.70	635,790	67.70	615,417	29.01	30,555
H	4.61	43,294	4.61	41,906	1.80	1,896
O	18.46	173,363	18.46	167,808	--	--
N	0.85	7,983	0.85	7,727	2.56	2,696
S	0.66	6,198	0.66	6,000	--	--
Ash	7.72	72,501	7.72	70,177	66.63	70,178
Subtotal	100.00	939,129	100.00	909,035	100.00	105,325
Moisture		264,883		54,640		--
Total		1,204,012		963,675		105,325
Total, lb/h		2,654,393		2,124,540		232,202

Table 19. PROCESS FLOW STREAMS FOR 263.9 TJ/d (250.2 X 10^9 Btu/d) HYDROGEN FROM MONTANA SUBBITUMINOUS COAL BY THE STEAM-IRON PROCESS

Stream No.	①		②		③		④		⑤		⑥		⑦	
Stream Name	Producer Gas		Oxidizer Effluent		Methanation Effluent		Hydrogen Product		Spent Producer Gas		Combustor Effluent		Gas Turbine Effluent	
Temperature, °K (°F)	1339 (1950)		1125 (1565)		597 (615)		333 (140)		1100 (1520)		1589 (2400)		921 (1198)	
Pressure, kN/m^2 (psig)*	2549 (355)		2515 (350)		2411 (335)		6996 (1000)		2515 (350)		1896 (2603)		114 (1.5)	
Composition	mol %	kg-mol/h	mol %	kg-mol/h	mol %	kg-mol/h	mol %	kg-mol/h	mol %	kg-mol/h	mol %	kg-mol/h	mol %	kg-mol/h
CO	27.44	42,117	0.54	554	0.04	38	0.10	38	8.76	14,103	--	--	--	--
CO_2	3.88	5,960	0.07	77	0.04	38	0.10	38	20.71	33,343	17.12	48,054	17.12	48,054
H_2	14.25	21,871	37.07	38,340	35.80	36,635	95.68	36,635	6.21	10,008	--	--	--	--
H_2O	4.28	6,563	61.34	63,443	62.59	64,038	0.01	5	16.00	25,759	13.24	37,169	13.24	37,169
CH_4	0.40	617	0.01	10	0.55	565	1.48	565	0.38	607	--	--	--	--
H_2S	0.12	188	--	--	--	--	--	--	0.12	188	--	--	--	--
N_2	49.63	76,169	0.97	1,007	0.98	1,007	2.63	1,007	47.82	77,012	64.54	181,122	64.54	181,122
O_2	--	--	--	--	--	--	--	--	--	--	5.03	14,123	5.03	14,123
SO_2	--	--	--	--	--	--	--	--	--	--	0.07	188	0.07	188
Total	100.00	153,485	100.00	103,431	100.00	102,321	100.00	38,288	100.00	161,020	100.00	280,656	100.00	280,656
Total, lb-mol/h		338,375		228,027		225,579		84,411		354,989		618,740		618,740

* kN/m^2 expressed in absolute units, i.e., 101.33 kilonewtons/$meter^2$ = 1 atm = 14.696 psia.

more uniform, even though it varies from the mine. About 3.2% of the feed coal is used as the dryer fuel.

The plant coal feed is ground to 80% below No. 10 U.S.S. Sieve size and dried to 5.67% moisture content in combination grinder-dryer mills. A lockhopper feed system, which has been used successfully in commercial Lurgi plants at 2170 to 2859 kN/m^2 (300 to 400 psig), is used to feed coal to the producer. Coal is fed to a lockhopper from a storage bin. After closing the lockhopper valves, the vessel is pressurized to the reactor pressure by an inert gas. The lockhopper discharge valve is then opened and the contents of the lockhopper flow by gravity into a continuous producer feed bin. The lockhopper is then depressurized and the cycle is repeated. Inert gas is recovered, recompressed, and recycled.

2. Producer Gas Generator and Steam-Iron Reactor

The ground and dried coal is continuously discharged to the fluidized-bed producer gas generator operating at 1339 K (1950°F) and 2549 kN/m^2 (355 psig). A high percentage of the carbon is gasified in the fluidized-bed producer. The following gasification reactions take place in the producer —

$$C + H_2O \rightarrow CO + H_2 - \text{Heat}$$

$$C + \underbrace{O_2 + N_2}_{\text{air}} \rightarrow CO + N_2 + \text{Heat}$$

Small quantities of CO_2, CH_4, and H_2S are also formed. The quantities of coal, steam, and air required for the producer and the steam-iron reactor are shown in Table 20. The residue from the producer is cooled to 367 K (200°F) in a water-filled quench tank. The residue-water slurry at 325 K (125°F) is depressurized and sent to a slurry-settling pond. The water is recycled to the quench pot and the residue is removed from the pond periodically.

The steam-iron reactor consists of an oxidizer and a reductor. A stream of iron oxide is cyclically reduced with producer gas in the reductor then reoxidized by the decomposition of steam in the oxidizer, or hydrogen-forming reactor. Solids circulation rate is about 27.2 million kg/h (60 million lb/h). The following reactions take place in the steam-iron reactor section.

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Table 20. STEAM-IRON GASIFICATION SYSTEM FEED QUANTITIES FOR 263.9 TJ/d (250.2 Billion Btu/d) HYDROGEN PLANT USING MONTANA SUBBITUMINOUS COAL

	Producer	Steam-Iron Reactor	Total Steam-Iron Gasification System
Coal, kg/hr (dry basis)	909,035	--	909,035
Coal, lb/hr (dry basis)	2,004,079	--	2,004,079
Steam, kg/hr	120,402	1,965,764	2,086,166
Steam, lb/hr	265,442	4,333,767	4,599,209
Air, kg/hr	2,786,515	67,845	2,854,360
Air, lb/hr	6,143,215	149,572	6,292,787

The composition of the producer gas is as follows:

	mole %
CO	27.4
CO_2	3.9
H_2	14.3
H_2O	4.3
CH_4	0.4
H_2S	0.1
N_2	49.6
Total	100.0

Temperature = 1339 K (1950°F)

Pressure = 2549 kN/m^2 (355 psig)

(Continued from page 45)

Reductor

$$Fe_3O_4 + CO \rightarrow 3\,FeO + CO_2$$

$$Fe_3O_4 + H_2 \rightarrow 3\,FeO + H_2O$$

Oxidizer

$$3\,FeO + H_2O \rightarrow Fe_3O_4 + H_2$$

In addition to the producer gas, additional amounts of steam and air are required for the steam-iron reactor; these quantities are listed in Table 20. Two effluent streams, one from the oxidizer and one from the reductor, are available for further processing. The compositions of both streams are shown below.

	Reductor Off-Gas		Oxidizer Effluent	
	Actual	(Dry)	(Dry)	Actual
	mole %			
CO	8.8	10.4	1.4	0.5
CO_2	20.7	24.7	0.2	0.1
H_2	6.2	7.4	95.9	37.1
H_2O	16.0	- -	- -	61.3
CH_4	0.4	0.5	- -	- -
N_2 + Ar	47.8	56.9	2.5	1.0
H_2S + COS	0.1	0.1	- -	- -
Total	100.0	100.0	100.0	100.0

	Reductor Off-Gas	Oxidizer Effluent
Pressure	2515 kN/m^2 (350 psig)	2512 kN/m^2 (350 psig)
Temperature	1100 K (1520°F)	1125 K (1565°F)

The oxidizer effluent, which contains primarily H_2 and steam, is upgraded to H_2 product, whereas reductor off-gas (spent producer gas) is used for power recovery.

3. Oxidizer Effluent Upgrading

The oxidizer effluent contains very small quantities of CO and CO_2, and no H_2S. This eliminates the processing steps of the CO-shift and acid-gas removal, so only methanation is required for upgrading the gas to the hydrogen product. Since a temperature of 561 K (550°F) for the methanation reactor feed is desirable, the effluent is cooled to this temperature in a waste-heat boiler, which generates about 50% of the steam required for the steam-iron reactor. Before methanation, dust particles are removed by cyclone separators and electrostatic precipitators. The gas passes through a zinc oxide bed (as a precautionary step) to prevent poisoning of the methanation catalyst by sulfur compounds that can be carried by the iron oxide from the reductor to the oxidizer.

The amounts of CO and CO_2 methanated are such that the final H_2 product contains 0.1% maximum of both CO and CO_2. The methanation reactor is a single-stage adiabatic reactor. The effluent at 597 K (615°F) is used in waste-heat recovery, then cooled to 311 K (100°F), dried in a glycol dryer to 112 kg/Mm^3 gas (7 lb H_2O/million SCF gas), compressed in a product gas compressor to 6996 kN/m^2 (1000 psig), cooled to 333 K (140°F), and finally sent to the pipeline. The composition of the product gas is as follows:

	mol %
CO	0.1
CO_2	0.1
H_2	95.7
CH_4	1.5
N_2 + Ar	2.6
Total	100.0

4. Power Generation From Reductor Off-Gas

The carbon monoxide and hydrogen in the spent producer gas are not completely converted in regenerating the iron oxide. The heating value plus the sensible heat of this stream represent 54% of the heating value of the input coal to the process. Part of this energy, 15% of the coal fuel value, is utilized in the plant to compress air for producer gas generation and to generate process steam. The remainder must either be utilized or wasted. There is the potential for generating a large amount of electric power, far more than is needed for the hydrogen manufacture alone.

We were able to obtain the services of UTRC of United Technologies Corp. for estimating the amount and cost of this power recovery. Most of the power is obtained by expansion of the gas through gas turbines and, therefore, the concentration of iron oxide particles carried over from the reductor must be reduced as much as possible. We have included in our design a two-stage cyclone system followed by electrostatic precipitators. The cleaned gas is available to the power recovery system at 2517 kN/m^2 (365 psia) and 1100 K (1520°F).

Two combined-cycle systems were studied:

1. Expansion to 1331 kN/m^2 (193 psia) with power recovery, followed by combustion and a combined-cycle system

2. Combustion at 1999 kN/m^2 (290 psia) followed by the combined cycle.

The second system showed slightly better power recovery and lower cost. This scheme, then, is the one selected for the design and analysis, and is shown in Figure 4. In this system, combustion air is compressed in Compressor A then introduced into the combustor, where it combines with the spent producer gas. Excess air maintains the exhaust temperature at 1589 K (2400°F). The combustion pressure of 1999 kN/m^2 (290 psia) results from allowing a 517 kN/m^2 (75 psi) pressure drop in the fuel meter, manifold, nozzle, and related components, as was assumed for Scheme I. A pressure drop of 103 kN/m^2 (15 psi) during the combustion process was also assumed. The temperature of 1589 K (2400°F) is beyond current technology, but by the time a steam-iron plant would be built, advanced technology will probably have made operation feasible at 1589 K (2400°F).

The heat content of this gas, heating value plus sensible heat, is 3.2 MJ/m^3 or 2688 kJ/kg (85 Btu/SCF or 1156 Btu/lb). UTRC believes utilization of this gas is possible. In addition to conventional combustors, catalytic combustors offer encouraging possibilities.

The exhaust gas flows are divided into three streams at the burner exit; each stream is proportional to the work required in the turbine expanders. In Turbine C, the net output power is used only to drive Compressor A, and the unit is completely self-contained. Turbine D, which drives Compressor B, the source of pressurized air for the Steam-Iron Process, is similarly self-contained; Power Turbine E is the only turbine component intended to drive an electric generator. All turbine components expand to a common discharge pressure of 114 kN/m^2 (16.5 psia), which is sufficient to cover losses in the subsequent heat exchanger units. The common exhaust gas stream is then divided into two streams, one of which is used to transfer process heat to water in a heat-to-process heat exchanger; the second exhaust stream heats water in the steam turbine system. A second heat exchanger in the heat-to-process stream, which recovers heat from a portion of the methanation effluent, is shown in Figure 4. Additional heat from the methanation effluent stream is used to preheat the water from the condenser discharge in the steam turbine system. Flow rate restrictions and thermodynamic temperature limits in the steam cycle dictate that the discharge temperature from the water preheater must not exceed 394 K (250°F).

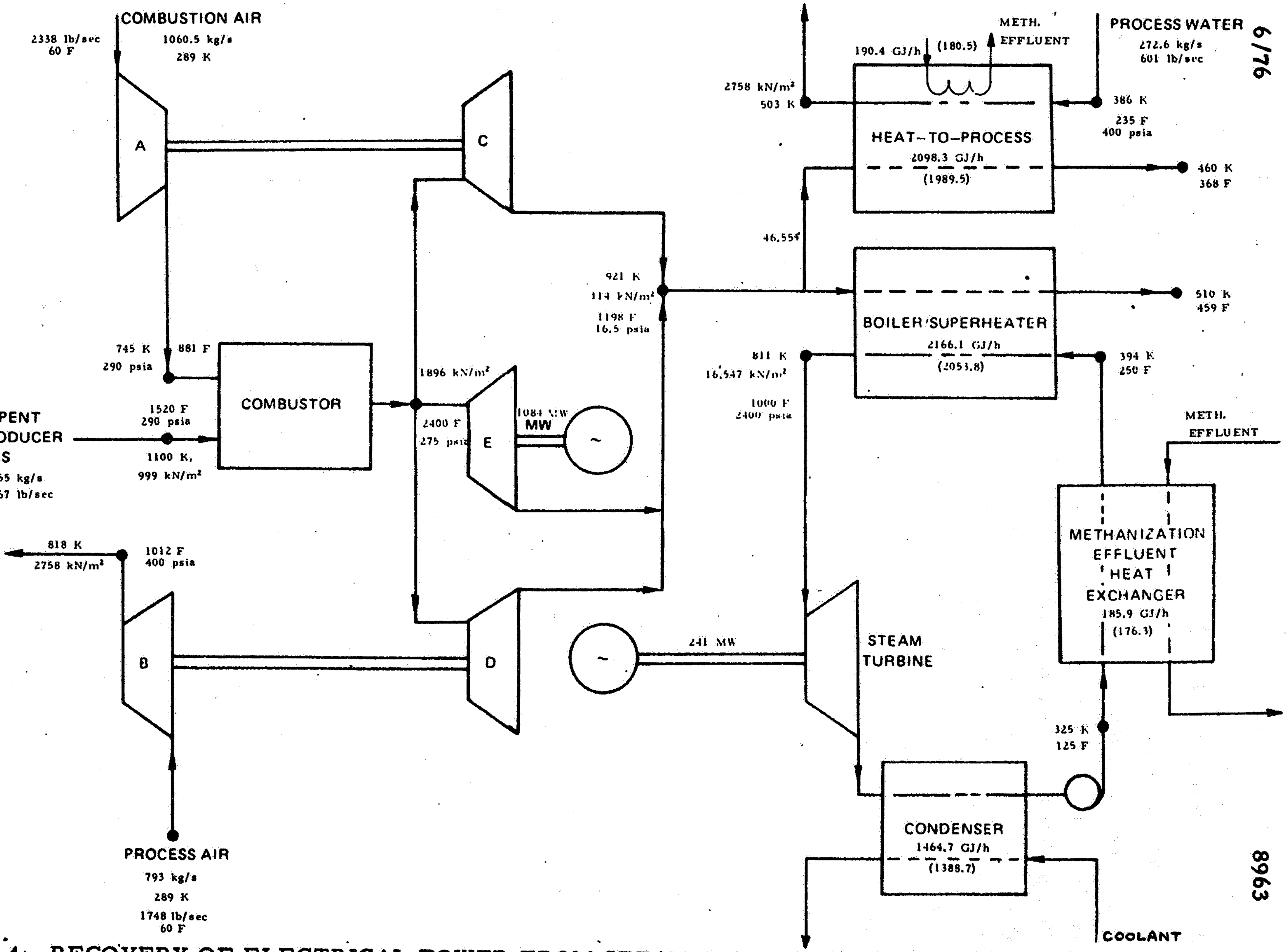

Fig. 4. RECOVERY OF ELECTRICAL POWER FROM STEAM-IRON PROCESS FOR 263.9 TJ/d (250.2×10^9 Btu/d) HYDROGEN FROM MONTANA SUBBITUMINOUS COAL

A tremendous amount of shaft power, 2.05 million kW (2.7 million total hp), is extracted from the products of combustion of the spent producer gas in expanding from 1896 to 114 kN/m^2 (275 to 16.5 psia).

About half of this power is used to supply the power needed to compress the air for producer gas generation and the air for the combustor. The rest of the shaft energy powers a generator that produces 1084 MW of electricity. After expansion to 114 kN/m^2 (16.5 psia), the gas is at a temperature of 921 K (1198 °F) and still contains substantial energy. This energy is then recovered in waste-heat boilers, with 46.5% transferred to process steam and 53.5% fueling the steam-turbine cycle. An additional 241 MW of electricity are generated, bringing the total to 1325 MW — a major power plant. About 96 MW are used locally in the hydrogen plant, leaving 1229 MW for export as a by-product. The heating value of the product hydrogen represents only 44.6% of the total heating value of the coal fed to the plant. The heat equivalent of the by-product electricity adds another 18% to give a total heating value output of 62.6%. If one considers the heat input necessary to generate this amount of electricity as a by-product, then the overall efficiency is substantially higher. For this combined cycle, the by-product heat energy input is 36.9% of the total plant coal. Adding this figure to 44.6% gives a total of 81.5%. Since we are taking by-product credit for electric power, which is worth much more than heat, then the heat equivalent of the electricity output appears to be the more appropriate figure.

UTRC's complete report is included as Appendix of this report.

5. Utilities and Other Offsite Facilities

The conversion of coal to hydrogen requires large quantities of utilities such as steam, power, and cooling water; these requirements are summarized in Tables 21-23. The plant is designed to be self-sufficient based on coal. Because reductor off-gas can provide all the power and steam needs of the plant, as shown in Figure 4 and Table 21, no separate coal-fired boiler is required. Table 24 summarizes the overall process efficiency and Table 25 summarizes the overall plant energy balance, giving an account of the remaining heat input from coal.

Table 21. SUMMARY OF STEAM REQUIREMENTS FOR 263.9 TJ/d (250.2 X 10^9 Btu/d) HYDROGEN FROM MONTANA SUBBITUMINOUS COAL BY THE STEAM-IRON PROCESS

Steam Required	kg/h	lb/h	GJ/h	10^6 Btu/h
H_2 Plant				
High-Pressure, 2654 kN/m^2 (385)				
Producer, 503 K (445)	120,402	265,442	329.4	312.3
Oxidizer, 503 K (445)	1,826,958	4,027,753	4997.8	4738.7
Aeration, 503 K (445)	138,806	306,014	380.3	360.6
Subtotal	2,086,166	4,599,209	5707.5	5411.6
Power Plant				
Turbine Steam for Power Cycle, 16,547 kN/m^2 (2400) 811 K (1000)	731,170	1,611,953	2352.0	2230.1
Subtotal	731,170	1,611,953	2352.0	2230.1
Boiler Blowdown and Deaeration Losses			22.7	21.5
Total	2,817,336	6,211,162	8082.2	7663.2
Steam Generated				
Waste-Heat Recovery From Oxidizer and Methanation Effluents for Preheating BFW and Generating Process Steam	1,105,300	2,436,771	3817.8*	3619.9*
Combustor Effluent Waste Heat Boiler for Process Steam Generation	980,866	2,162,438	2098.3	1989.5
Combustor Effluent Waste-Heat Boiler-Superheater for Power Generation	731,170	1,611,953	2166.1†	2053.8†
Total	2,817,336	6,211,162	8082.2	7663.2

* About 47.5% of the duty is used for BFW preheat

† About 28.3% of the duty is used for Superheating turbine steam.

Table 22. SUMMARY OF POWER REQUIREMENTS FOR 263.9 TJ/d (250.2 X 10^9 Btu/d) HYDROGEN FROM MONTANA SUBBITUMINOUS COAL BY THE STEAM-IRON PROCESS

Component	kW	hp
H_2 Plant		
Coal Storage and Reclaiming	2,781	3,730
Coal Grinding and Drying	10,716	14,370
Lock Hopper System	6,465	8,670
Methanation Effluent Air Cooler	1,268	1,700
Product Gas Compressor	48,471	65,000
Product Gas Air Cooler	112	150
Residue Handling	746	1,000
Cooling Water Fans and Pumps	686	920
Boiler Feedwater Pumps	2,423	3,250
Power Recovery Section		
Boiler Feedwater Pumps	4,922	6,600
Cooling Water Fans and Pumps	6,838	9,170
Turbine Blades Cooling	2,297	3,080
Miscellaneous for Both Sections	8,202	11,000
Subtotal	95,927	128,640
Producer Air Compressor	438,458	587,982
Combustor Air Compressor	501,175	672,086
Total Power Required	1,035,561	1,388,708
Power Recovery From Spent Reductor Gas		
From Combustor Gas Expansion (on shaft)	2,045,551	2,743,129
Less Power for Producer Air Compression	(438,458)	(587,982)
Less Power for Combustor Air Compression	(501,175)	(672,086)
Net Power Available From Gas Turbine	1,105,918	1,483,061
Power Recovery From Generator (98%)	1,083,800	1,453,400
Power Recovery From Steam Cycle	241,600	323,991
Total	1,325,400	1,777,391
Electric Power to Plant	(95,927)	(128,640)
By-Product Power	1,229,473	1,648,751

Table 23. SUMMARY OF COOLING WATER REQUIREMENTS FOR 263.9 TJ/d (250.2 X 10^9 Btu/d) HYDROGEN FROM MONTANA SUBBITUMINOUS COAL BY THE STEAM-IRON PROCESS

Component	m^3/h	gpm
H_2 Plant *		
Lock Hopper System	409	1,800
Residue Handling	1,862	8,200
Methane Effluent Water Cooling	909	4,000
Subtotal	3,180	14,000
Power Plant System (From UTRC Report by Heat Balance)†	31,541	138,870
Total	34,721	152,870

* Cooling water temperature, H_2 plant, 303°-319°K (85°-115°F).

† Cooling water temperature, power plant, 303°-314°K (85-105°F).

Table 24. STEAM-IRON GASIFICATION – PROCESS EFFICIENCY FOR 263.9 TJ/d (250.2 Billion Btu/d) HYDROGEN PLANT USING MONTANA SUBBITUMINOUS COAL

	kg/h	lb/h
Reactor Coal, (dry basis)	909,039	2,004,079
Boiler and Dryer Coal (dry basis)	30,094	66,347
Total Coal, (dry basis)	939,133	2,070,426

	GJ/h	10^6/h
HHV Total Coal,*	24,653	23,374.5
HHV Product Gas	10,993.1	10,423.2
% Converted to Product Gas	44.6	
HHV of By-Product Power, (1,229,473 KW)	4,425.6	4,196.2
% Converted to By-Product Power	18.0	
HHV Total Products,	15,418.7	14,619.4
% Converted to Total Products	62.6	

* At 26,251 kJ/kg (11,290 Btu/lb) dry basis.

Summarizing, a 263.9 TJ/d (250.2 billion Btu/d) H_2 plant using Montana subbituminous coal for the steam-iron gasification process requires 28,896 t/d (31,853 short tons/d) of raw coal containing 22% moisture. Of the coal HHV, 44.6% is converted to HHV of H_2 product and 18.0% to by-product power.

Table 25. SUMMARY OF OVERALL PLANT ENERGY BALANCE 263.9 TJ/d (250.2 X 10^9 Btu/d) HYDROGEN FROM MONTANA SUBBITUMINOUS COAL BY THE STEAM-IRON PROCESS [BASIS: 289 K (60 °F) Liquid H_2O]

	GJ/h	10^6 Btu/h	%
Heat In			
Process Coal	23,862.7	22,625.5	96.8
Dryer Fuel	790.0	749.0	3.2
Total	24,652.7	23,374.5	100.0
Heat Out			
Product Gas	10,993.1	10,423.2	44.6
By-Product Power	4,425.6	4,196.2	18.0
Heat to Cooling Water, H_2 Plant	221.5	210.0	0.9
Heat to Cooling Water, Power Plant	1,464.6	1,388.7	5.9
Heat to Air Cooling	1,897.3	1,798.9	7.7
Dryer Off-Gas	565.9	536.6	2.3
Heating Value of Char Residue	1,270.1	1,204.2	5.2
Stack Gas at 460K (368)	1,479.2	1,402.5	6.0
Stack Gas at 510K (459)	1,947.4	1,846.4	7.9
Subtotal	24,264.7	23,006.7	98.5
Assumed Waste-Heat Recovery Losses	317.4	300.9	1.3
Subtotal	24,582.1	23,307.6	99.8
Heat Unaccounted for	70.6	66.9	0.2
Total	24,652.7	23,374.5	100.0

254.7 TJ/d (241.5 X 10^9 Btu/d) SUBSTITUTE NATURAL GAS (SNG) FROM MONTANA SUBBITUMINOUS COAL BY THE HYGAS STEAM-OXYGEN PROCESS

This design, based on the HYGAS Process, provides a reference overall efficiency and cost for manufacturing methane from coal for comparing hydrogen and methane processes. This process is being studied in a large pilot plant in Chicago under the sponsorship of ERDA and the A.G.A. The second methane design is based on the Consolidation Coal CO_2 Acceptor Process.

The HYGAS Process produces a gas of pipeline quality, containing 94.7% methane. The product composition is as follows:

	mol %
CH_4	94.7
H_2	4.4
CO	0.1
CO_2	0.1
N_2 + Ar	0.7
H_2O	<0.1
Total	100.0

The methane can be liquefied after pipeline transmission to the point of destination.

Crushed and dried coal is reacted with steam and oxygen to produce methane-rich raw gas in a fluidized-bed reactor. This raw gas is upgraded to pipeline quality in several steps following the reactor. Any type of coal, from bituminous to lignite, can be gasified in this process. For the caking bituminous coals, pretreatment with air at 644 to 700 K (700° to 800°F) is required to reduce the agglomerating tendency of the coal.

Since Montana subbituminous coal is used for this study, no pretreatment is necessary. The processing steps required in the conversion of coal to SNG are shown in Figure 5, and the compositions of various streams are shown in Tables 26 and 27. These steps can be grouped into the following sections:

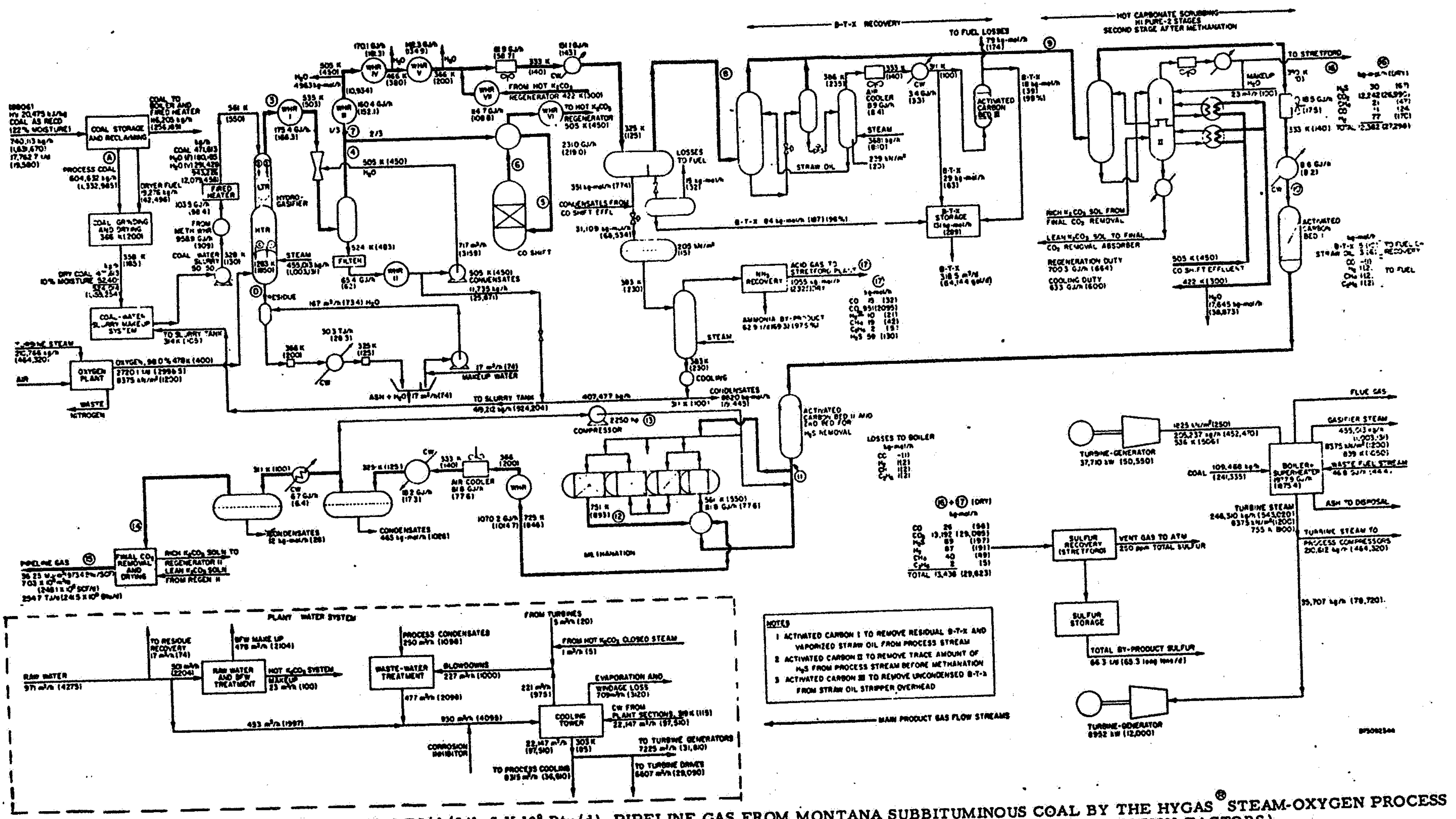

Fig. 5. PROCESS FLOW DIAGRAM FOR 254.7 TJ/d (241.5 X 10⁹ Btu/d) PIPELINE GAS FROM MONTANA SUBBITUMINOUS COAL BY THE HYGAS® STEAM-OXYGEN PROCESS (NUMBERS IN PARENTHESIS HAVE CONVENTIONAL BRITISH UNITS. SEE TABLE 3 FOR UNITS AND CONVERSION FACTORS)

Table 26. COMPOSITION OF SOLID STREAMS FOR 254.7 TJ/d (241.5 X 10^9 Btu/d) PIPELINE GAS FROM MONTANA SUBBITUMINOUS COAL BY THE HYGAS STEAM-OXYGEN PROCESS

Stream	Ⓐ		Ⓑ		Ⓒ		Ⓓ	
Stream Name	Process Coal		Coal to Hydrogasifier		Hydrogasifier Char		Residue	
Component	wt %	kg/h	wt %	kg/h	wt %	kg/h	wt %	kg/h
C	67.70	319,286	67.70	319,286	80.51	166,998	14.36	6,390
H	4.61	21,741	4.61	21,741	0.94	1,940	0.17	74
O	18.46	87,061	18.46	87,061	---	---	---	---
N	0.85	4,003	0.85	4,003	0.68	1,411	3.17	1,411
S	0.66	3,113	0.66	3,113	0.32	669	0.42	189
ASH	7.72	36,409	7.72	36,409	17.55	36,409	81.88	36,409
SUBTOTAL	100.00	471,613	100.00	471,613	100.00	207,426	100.00	44,473
MOISTURE	---	133,019	--	--	--	--	--	--
TOTAL		604,632	100.00	471,613	100.00	207,426	100.00	44,473
TOTAL lb/hr		1,332,985		1,039,729		457,926		98,046

Table 27. PROCESS FLOW STREAMS FOR 254.7 TJ/d (241.5 X 10^9 Btu/d) PIPELINE GAS FROM MONTANA SUBBITUMINOUS COAL BY THE HYGAS STEAM-OXYGEN-PROCESS

Stream No.	①		②		③		④		⑤		⑥		⑦	
Stream Name	Oxygasifier Effluent		Hydrogasifier Effluent		Slurry Water Vaporizer Effluent		Water-Quench Effluent		CO-Shift Feed		CO-Shift Effluent		CO-Shift Bypass	
Temperature, K (°F)	1283 (1850)		811 (1000)		589 (600)		524 (483)		561 (550)		640 (692)		524 (483)	
Pressure, kN/m² (psig)*	8203 (1175)		8134 (1165)		8065 (1155)		7996 (1145)		7927 (1135)		7858 (1125)		7858 (1125)	
Component	mol %	kg-mol/h	mol %	kg-mol/h	mol %	kg-mol/h	mol %	kg-mol/h	mol %	kg-mol/h	mol %	kg-mol/h	mol %	kg-mol/h
CO	17.00	6,230	20.13	9,432	12.91	9,432	13.03	9,432	13.03	6,288	3.40	1,640	13.03	3,144
CO_2	15.53	5,694	18.65	8,738	11.97	8,738	12.07	8,738	12.07	5,826	21.70	10,474	12.07	2,912
H_2	23.98	8,791	23.68	11,093	15.19	11,093	15.32	11,093	15.32	7,395	24.95	12,043	15.32	3,698
H_2O	39.50	14,478	22.58	10,627	50.40	36,805	49.95	36,153	49.95	24,102	40.32	19,454	49.95	12,051
CH_4	3.95	1,449	12.86	6,027	8.25	6,027	8.33	6,027	8.33	4,018	8.33	4,018	8.33	2,009
C_2H_6	--	--	0.99	462	0.63	462	0.64	462	0.64	308	0.64	308	0.64	154
NH_3	--	--	0.34	158	0.22	158	0.22	158	0.22	105	0.22	105	0.22	53
H_2S	0.04	15	0.19	91	0.12	91	0.13	91	0.13	61	0.13	61	0.13	30
N_2 + Ar	--	--	0.18	84	0.12	84	0.12	84	0.12	56	0.12	56	0.12	28
B-T-X	--	--	0.30	138	0.19	138	0.19	138	0.19	92	0.19	92	0.19	46
Total	100.00	36,657	100.00	46,850	100.00	73,028	100.00	72,376	100.00	48,251	100.00	48,251	100.00	24,125
Total, lb-mol/h		80,814		103,288		160,999		159,563		106,375		106,375		53,188

Stream No.	⑧		⑨		⑩		⑪		⑫		⑬		⑭		⑮	
Stream Name	B-T-X Recovery Feed		Hot K_2CO_3 Feed		Hot K_2CO_3 Effluent		Methanation Feed		Methanation Wet Product		Methanation Recycle		Methanation Dry Product		Pipeline Gas	
Temperature, K (°F)	325 (125)		325 (125)		325 (125)		325 (125)		751 (893)		325 (125)		311 (100)		311 (100)	
Pressure, kN/m² (psig)*	7755 (1110)		7686 (1100)		7582 (1085)		7513 (1075)		7272 (1040)		7168 (1025)		7168 (1025)		6995 (1000)	
Component	mol %	kg-mol/h	mol %	kg-mol/h	mol %	kg-mol/h	mol %	kg-mol/h	mol %	kg-mol/h	mol %	kg-mol/h	mol %	kg-mol/h	mol %	kg-mol/h
CO	12.03	4,769	12.06	4,765	17.44	4,754	17.45	4,752	0.09	51	0.10	38	0.10	13	0.10	13
CO_2	31.36	12,425	31.30	12,363	1.00	272	1.00	272	1.19	664	1.30	500	1.31	164	0.10	13
H_2	39.70	15,729	39.82	15,729	57.43	15,652	57.46	15,651	3.97	2,218	4.36	1,671	4.36	547	4.42	547
H_2O	0.17	68	0.17	68	0.18	49	0.18	49	9.07	5,075	0.18	70	0.09	11	0.02	2‡
CH_4	15.16	6,006	15.19	6,001	21.94	5,979	21.94	5,977	85.09	47,580	93.42	35,849	93.47	11,731	94.68	11,731
C_2H_6	1.16	460	1.15	455	1.67	455	1.66	453	--	--	--	--	--	--	--	--
B-T-X	0.13	52	0.02	7†	0.03	7†	--	--	--	--	--	--	--	--	--	--
H_2S	0.08	32	0.08	30	--	--	--	--	--	--	--	--	--	--	--	--
N_2 + Ar	0.21	84	0.21	84	0.31	84	0.31	84	0.59	330	0.64	246	0.67	84	0.68	84
Total	100.00	39,625	100.00	39,502	100.00	27,252	100.00	27,238	100.00	55,918	100.00	38,374	100.00	12,550	100.00	12,390
Total, lb-mol/h		87,359		87,087		60,080		60,048		123,279		84,601		27,669		27,315

† Includes 3 kg-mol/h straw oil.

* kN/m² expressed in absolute units, i.e., 101.33 kilonewtons/meter² = 1 atm = 14.696 psia.

‡ 112 kg/Mm³ (7 lb/10^6 SCF).

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1. Coal Storage and Reclaiming
2. Coal Grinding and Drying
3. Coal-Water Slurry Feed System
4. HYGAS Reactor System
5. Dust Removal System
6. CO Conversion
7. B-T-X Recovery
8. Acid-Gas Removal System
9. Methanation and Drying
10. Waste-Water Treatment
11. Sulfur Recovery
12. Utilities and Other Offsite Facilities

1. Coal Storage and Reclaiming

The SNG plant is assumed to be located near a coal mine that will provide a minimum 25-year supply of coal. Raw coal is brought from the mine by truck to a storage area where provision is made to store a 30-day supply of coal broken to 3.8 cm x 0 (1-1/2 in. x 0) size. Coal is stored in such a way that it is uniformly distributed in the storage pile. This plant requires a continuous supply of 17,763 t/d (19,580 short tons/d) of the raw coal from the mine (22% moisture). Because of the distribution of coal to the storage piles, the composition of the coal feed to the plant approaches a uniform condition although it varies from the mine. About 15.7% of the feed coal is used as boiler fuel and 2.6% as the dryer fuel.

2. Coal Grinding and Drying

The plant coal feed is ground to less than No. 8 U.S.S. sieve size — 80% below sieve size No. 12. (Ten percent to 11% below 100 sieve size is desirable.) It is then dried to 10% moisture content in combination grinder-dryer mills. The dried coal at 366 K (200°F) is pneumatically conveyed to the slurry preparation section.

3. Coal-Water Slurry Feed System

In this section, the feed is mixed with water to form a coal-water slurr of 50:50 consistency. Large reciprocating pumps are used to pressurize the slurry to the reactor pressure. The slurry is preheated and the water is

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partially vaporized at 561 K (550°F) using recovered waste heat and a fired heater before entering the vaporizer section of the HYGAS reactor. The slurry preheat step is necessary because the heat in the gasifier effluent is not enough to vaporize all the slurry water.

4. HYGAS Reactor System (Hydrogasifier)

The preheated coal-water slurry at 561 K (550°F) comes in contact with the hydrogasifier effluent in the fluidized-bed vaporizer, where all the slurry water is vaporized at 589 K (600°F). The dried coal then drops from the vaporizer through a feed standpipe to the bottom of the low-temperature reaction zone (LTR) where, together with recycle char, it is picked up by the effluent gas from the high-temperature reactor (HTR) at 1211 K (1720°F). The LTR is a transport reactor with a 10-second residence time in which the char is devolatilized, which enriches the gases from the HTR. The enriched gases leave the LTR through a cyclone separator at 811 K (1000°F) and are used to vaporize slurry water. The char from the cyclone separator is divided into HTR feed and recycle feed by a splitter valve. The HTR feed drops into the HTR bed through a dipleg. Of the total methane made into the hydrogasifier, 8.8% is made in the LTR. All of the ethane, ammonia, B-T-X, and most of the hydrogen sulfide are made in the LTR.

In the HTR fluidized bed, char from the LTR reacts at 1211 K (1720°F) with the synthesis gas produced in the gasifier. The following equivalent reactions take place, with the methane-forming reaction supplying heat for the steam-carbon reaction:

$$\begin{array}{ll} C + 2H_2 \rightarrow CH_4 & \text{(Exothermic)} \\ CO + H_2O \rightarrow H_2 + CO_2 & \text{(Slightly Exothermic)} \\ C + H_2O \rightarrow CO + H_2 & \text{(Endothermic)} \end{array}$$

The HTR is designed for a char residence time of 29 minutes and gas velocity of 15.24 cm/s (0.5 ft/s). About 67.2% of the methane in the hydrogasifier effluent is generated in the HTR.

The char from the HTR is reacted with steam and oxygen in the fluid-bed steam-oxygen gasifier at 1283 K (1850°F) to produce the synthesis gas required for supplying hydrogen for hydrogasification. The equivalent reactions taking place in this section are:

$$\begin{array}{lll}
C + H_2O \rightarrow CO + H_2 & & \text{(Endothermic)} \\
C + O_2 \rightarrow CO_2 & & \text{(Exothermic)} \\
C + CO_2 \rightarrow 2CO & & \text{(Endothermic)} \\
C + 2H_2 \rightarrow CH_4 & & \text{(Exothermic)} \\
CO + H_2O \rightarrow CO_2 + H_2 & & \text{(Slightly Exothermic)}
\end{array}$$

The oxygasifier is designed for a char residence time of 17 minutes and a gas velocity of 30.48 cm/s (1 ft/s). About 24% of the total methane formed in the hydrogasifier is generated in the oxygasifier. The char from the oxygasifier bed is cooled from 1283 K (1850°F) to 866 K (1100°F) by heating the 455,013 kg/h (1,003,131 lb/hr) of incoming steam from 839 K (1050°F) to 866 K (1100°F) in a mixing zone. The oxygen used for gasification is 2720.1 t/d (2998.5 short tons/d), containing 98% O_2 and 2% N_2 + Ar. Oxygen is supplied to the gasifier at 8375 kN/m^2 and 478 K (1200 psig and 400°F) from two 1361 t/d (1500 short tons/d) air separation plants.

Two hydrogasifiers, each weighing over 1814 t (2000 short tons) are required for this plant. The feed quantities required for the gasifier operation are given below:

Input

Coal (Dry Basis)	471,613 kg/h (1,039,729 lb/h)
Steam	455,013 kg/h (1,003,131 lb/h)
Oxygen	2720.1 t/d (2,998.5 short tons/d)
Slurry Water	471,613 kg/h (1,039,729 lb/h)

Output

CO + H_2	18,170 kg-mol/h (40,057 lb-mol/h)
CH_4	6027 kg-mol/h (13,288 lb-mol/h)

Carbon Conversion to Gases = 98%

5. Dust Removal System

The effluent from the slurry water vaporizer at 589 K (600°F) is cooled to 535 K (503°F), or 11 K (20°F) above the dew point of the gas by waste-heat recovery. Small dust particles (less than 5 microns) carried over from the cyclones in the gasifier are taken out in a jet venturi scrubber that has a high efficiency for removing particles 1 to 2 microns in size. The composition of the gas stream after dust removal is given below.

COMPOSITION OF RAW GAS FROM HYDROGASIFIER

	mol %
CO	13.03
CO_2	12.07
H_2	15.32
H_2O	49.95
CH_4	8.33
C_2H_6	0.64
NH_3	0.22
H_2S	0.13
N_2 + Ar	0.12
B-T-X	0.19
Total	100.00

The system maintains the gas at a temperature above 524 K (483°F) so that most of the water is retained for the CO conversion (shift) reaction. This makes it unnecessary to generate the large amounts of steam in a boiler and raises the overall plant efficiency.

6. CO Conversion

In order to upgrade this gas to pipeline quality, it is necessary to adjust the H_2/CO ratio from 1.18 in the gasifier effluent to above 3.2 for methanation of CO, by the well-known CO-shift reaction —

$$CO + H_2O \rightleftharpoons CO_2 + H_2 \text{ (Exothermic)}$$

As the gasifier effluent contains both B-T-X and H_2S, neither the conventional high-temperature chromium-promoted iron oxide nor the low-temperature zinc- and copper-based CO-shift catalyst can be used. The catalyst used must be either Girdler's G-93 Co-Mo or an equivalent that can operate satisfactorily in the presence of oil and sulfur.

The effluent from the dust removal system is split into two streams, with two-thirds going to the CO-shift reactor and one-third used as a bypass stream for control purposes.

Based on laboratory tests, a minimum steam-to-dry gas ratio of 1 is required when oil is present in the gas. The operating temperature range is from 547 to 755 K (525° to 900°F). The CO-shift feed has a steam-to-dry gas ratio of 1. It is preheated to 561 K (550°F) by the shift effluent. The CO-shift effluent.temperature is 640 K (692°F).

The CO-shift effluent is cooled to 325 K (125°F) by preheating the boiler feed water, regenerating the hot potassium carbonate solution, and generating low-pressure steam. The condensed water containing H_2S, CO_2, NH_3, and B-T-X is sent to the waste-water treatment step (Step 10).

7. B-T-X Recovery

The CO-shift effluent at 325 K (125°F) contains 0.73% B-T-X, 52 kg-mole/h (114 lb-mole/h); this is a valuable by-product. The effluent is scrubbed with straw oil to remove most of the B-T-X. The remaining B-T-X, 5 kg/mole/h (10 lb-mole/h), and a small amount of straw oil, 3 kg-mole/h, (6 lb-mole/h) are recovered after the first stage of the hot carbonate acid-gas removal system in the activated carbon tower, which contains BPL-type activated carbon. B-T-X is recovered from straw oil by stripping with steam. The stripping steam and a portion of the B-T-X are condensed, separated, and the B-T-X is sent to storage. The uncondensed B-T-X and other vapors are passed over another activated carbon bed containing BPL-type activated carbon for recovery. The activated carbon beds operate on a 4-hour cycle with 2 hours for adsorption and 2 hours for regeneration of the bed with steam.

8. Acid-Gas Removal System

The effluent from the straw oil B-T-X recovery system at 325 K (125°F), which contains 31.3% CO_2 and 0.08% H_2S, goes to the first stage of the two-stage HIPURE hot carbonate acid-gas removal system. This process is a typical method used for acid-gas removal. It is not necessarily the optimum method; such a determination is beyond the scope of this study. This gas is scrubbed with hot potassium carbonate solution. The effluent leaving the absorber at the top contains about 1% CO_2 and about 5 ppm H_2S at 350 K (170°F).

The rich carbonate solution containing H_2S and CO_2 is depressurized to 1703 kN/m^2 (10 psig) and regenerated in a stripper operating at about 389 K and 170.3 kN/m^2 (240°F and 10 psig), with heat supplied by the CO-shift effluent stream in a reboiler.

Acid gases stripped from the hot carbonate solution are sent to a Stretford Process unit, where the H_2S is converted to sulfur. The lean solution from the stripper bottom is pumped to 8030 kN/m^2 (1150 psig), cooled, and returned to the top of the absorber. About 50% of the power required to pump the hot carbonate to the absorber pressure is recovered by depressurizing the rich solution through a hydraulic turbine. The other power is supplied by an electric motor drive.

The absorber effluent at 350 K (170°F) is cooled to 325 K (125°F) before going to the activated carbon and ZnO beds for trace H_2S removal, which is necessary to avoid poisoning the methanation catalyst.

9. Methanation and Drying

The purified gas, which contains no sulfur but 17.45% CO, passes over the high-nickel catalyst where most of the CO and some of the CO_2 react with H_2 to form methane by the following reactions:

$$CO + 3H_2 \rightarrow CH_4 + H_2O$$
$$CO_2 + 4H_2 \rightarrow CH_4 + 2H_2O$$

This step increases the heating value of the gas to 36.25 MJ/m^3 (973 Btu/SCF) for the final product and reduces the CO content to 0.1%, as required to meet the pipeline gas specifications. The methanation reaction is highly exothermic; therefore, it is essential to control the temperature to prevent catalyst deactivation. This is accomplished by using recycle quench methanation with four stages, each with an inlet temperature of 561 K (550°F) and a product-recycle to fresh-feed ratio such that the CO content in the feed to each stage is about 4%, which limits the maximum catalyst-bed temperature to 755 K (900°F).

The methanation wet product at 751 K (893°F) is cooled to 366 K (200°F) in a series of exchangers that utilize the waste heat for preheating the methanation 1st stage feed to 561 K (550°F) and preheating the coal-water slurry. Further cooling of the gas to 325 K (125°F) is accomplished by air and water cooling. The cooled gas is split into two streams: 68.6% of the gas is recycled to mix with the feed stream and the remaining gas goes through a final hot potassium carbonate solution (HIPURE Stage II) to reduce the CO_2 to 0.1%. The purified gas is cooled to 311 K (100°F) and then dried to the pipeline standard of 112 kg H_2O/Mm^3 (7 lb $H_2O/10^6$ SCF) in a glycol drier. The product pipeline gas leaves the plant at 6996 kN/m^2 (1000 psig).

10. Waste-Water Treatment

The condensates from the CO-shift effluent after the separation of B-T-X go through the Chevron waste-water treatment process. The acid gases and NH_3 are stripped from the condensed water by stream. The stripped gases go to the ammonia recovery system where 62.9 t/d (69.3 short tons/d) of ammonia are recovered as a by-product; the acid gases go to the sulfur-recovery system. The stripped water at 311 K (100°F) is used as makeup for the coal-water slurry feed and for the cooling tower.

11. Sulfur Recovery

The acid gases from the waste-water treatment system are combined with the acid gases from the hot carbonate stage, which contains the H_2O, for sulfur removal. Since the H_2S concentration in the sulfur recovery feed stream is only 0.7%, the Stretford Process for the recovery of sulfur is used. In this process, an aqueous solution of sodium carbonate reacts with H_2S to form sodium hydrosulfide by the following reaction —

$$H_2S + Na_2CO_3 \rightarrow NaHS + NaHCO_3$$

The hydrosulfide is oxidized to sulfur by sodium vanadate, which is also in solution, by —

$$NaHS + NaHCO_3 + 2NaVO_3 \rightarrow S + Na_2V_2O_5 + Na_2CO_3 + H_2O$$

The $Na_2V_2O_5$ is oxidized back to the pentavalent state by blowing with air containing anthraquinone disulfonic acid (ADA) or sodium anthraquinone disulfonate, which acts as an oxidizing catalyst in the regenerator —

$$Na_2V_2O_5 + 1/2\ O_2 \xrightarrow{ADA} 2NaVO_3$$

During the regeneration of the solution, tiny particles of sulfur collect as a froth on top of the solution. The froth is skimmed off the solution and filtered. The filtered sulfur amounts to 66.3 t/d (65.3 long tons/d) of by-product sulfur. The vent gases from the absorber contain about 250 ppmv of total sulfur.

12. Utilities and Other Offsite Facilities

The conversion of coal to pipeline gas requires large quantities of utilities such as steam, power, and cooling water; these requirements are summarized in Tables 28-30. The plant is designed to be self-sufficient based on coal. The boiler duty is 1978 GJ/h (1875 X 10^6 Btu/h). It is designed to supply 8375 kN/m^2 (1200 psig) steam for the gasifier and for the production of plant power requirements. Most of the low-pressure steam is generated through waste-heat recovery. A small coal-fired boiler with a duty of 103.5 GJ/h (98.5 X 10^6 Btu/h), is used to supplement waste-heat recovery in preheating the coal-water slurry to 561 K (550 °F).

The overall plant thermal efficiency is 74%; of this figure, 70% represents pipeline gas and the rest is in by-products like B-T-X, ammonia, and sulfur. Table 31 presents the overall process efficiency.

The overall heat balance summary, Table 32, gives the overall energy balance.

In summary, the production of 254.7 TJ/d (241.5 X 10^9 Btu/d) of pipeline gas from Montana subbituminous coal using the HYGAS Steam-Oxygen Process, with the reactor operating at 8135 to 8203 kN/m^2 (1165 to 1175 psig) and 811 to 1283 K (1000° to 1850°F), requires 17,763 t/d (19,580 short tons/d) of 22% moisture coal. In this design, 70% of the higher heating value (HHV) of the coal is converted to HHV of pipeline gas and 4% goes to by-products, for a total overall efficiency of 74%.

Table 28. SUMMARY OF STEAM REQUIREMENTS FOR 254.7 TJ/d (241.5 X 10^9 Btu/d) PIPELINE GAS FROM MONTANA SUBBITUMINOUS COAL BY THE HYGAS STEAM-OXYGEN PROCESS

Steam Required	kg/h	lb/h	GJ/h	10^6 Btu/h
High Pressure 8378 kN/m² (1200)				
Hydrogasifier 839K (1050)	455,013	1,003,130	1586.7	1504.5
Turbine 755K (900)	246,310	543,020	771.4	731.4
CO Shift Steam (Vaporized Slurry Feed H_2O + H_2O From Gasifier Effluent)	--	--	--	--
Subtotal	701,323	1,546,150	2358.1	2235.9
Low Pressure 1825 kN/m² (250)				
Steam to Turbines 536K (506)	205,237	452,470	559.4	530.4
Low Pressure 791 kN/m² (100)				
Hot Carbonate Regeneration Steam	22,975	50,650	47.0	44.6
Low Pressure 446 kN/m² p (50)				
Waste-water Treatment	74,784	164,870	158.4	150.2
Sulfur Removal	9,212	20,310	26.4	25.0
B-T-X Recovery	12,565	27,700	43.5	41.2
Subtotal	324,773	716,000	834.7	791.4
Total	1,026,096	2,262,150	3192.8	3027.3
Steam Generated				
Waste Heat recovered from CO Shift Effluent, Methanation Effluent, etc., for Boiler Feed Water Preheat and Low Pressure Steam Generation	324,773	716,000	1214.9	1151.9*
Steam Generation in the Boiler-Superheater	701,323	1,546,150	1977.9	1875.4†
Total	1,026,096	2,262,150	3192.8	3027.3

* About 53.8% of the duty is used for boiler feed water preheat.

† About 27.4% of the duty is used for superheating process and turbine steams.

Table 29. SUMMARY OF POWER REQUIREMENTS FOR 254.7 TJ/d (241.5 X 10^9 Btu/d) PIPELINE GAS FROM MONTANA SUBBITUMINOUS COAL BY THE HYGAS STEAM-OXYGEN PROCESS

Section	kW	hp
Coal Storage and Reclaiming	1,723	2,310
Coal Grinding and Drying	5,528	7,410
Slurry Feed System	4,834	6,480
Quench System, CO-Shift Effluent Air Cooling	194	260
Benzene Recovery (Straw Oil + Activated Carbon I + III)	1,775	2,380
Prepurification (Hot K_2CO_3 + Activated Carbon II)	9,758	13,080
Sulfur Recovery, Stretford	3,685	4,940
Methanation and Drying	1,738	2,330
Boiler Fuel Conveyor (Assumed)	149	200
Waste-Water Treatment	1,679	2,250
Oxygen Plant Air Compressor	36,748	49,260
Oxygen Compressor	18,464	24,750
Turbine Steam Feedwater Pumps	1,201	1,610
Boiler Feedwater Pumps	1,485	1,990
Oxygasifier Char Slurry-Water Pump	597	800
Cooling-Water Pumps	4,804	6,440
Miscellaneous	7,460	10,000
Total Power Required	101,814	136,480
Power From Turbines		
Oxygen Plant Air Compressors	36,748	49,260
Oxygen Compressor	18,464	24,750
Subtotal	55,212	74,010
Power From 250 psig steam turbogenerator	37,710	50,550
Power From 1200 psig steam turbogenerator	8,892	11,921
Total Power Generated	101,814	136,480

Table 30. SUMMARY OF COOLING WATER REQUIREMENTS FOR 254.7 TJ/d (241.5 X 10^9 Btu/d) PIPELINE GAS FROM MONTANA SUBBITUMINOUS COAL BY THE HYGAS STEAM-OXYGEN PROCESS

Section	m^3/h^*	gpm*
Oxygasifier Char Cooling	434	1,910
Straw Oil System	28	120
Benzene Recovery Activated Carbon I and III	711	3,130
Hot K_2CO_3	2,215	9,750
Trace H_2S Removal	273	1,200
CO-Shift Cooling	216	950
Condensates Cooling	284	1,250
Methanation	359	1,580
Final Gas Drying	34	150
Oxygen Plant, Compressor Coolers	3,416	15,040
Waste-Water Treatment	325	1,430
Subtotal	8,315	36,610
Steam Turbine Condensers	6,607	29,090
Turbogenerator, 250 psig Steam	6,082	26,780
Turbogenerator, 1200 psig Steam	1,143	5,030
Total Cooling Water	22,147	97,510

*Cooling water temperature 303-319° K (85-115° F).

Table 31. HYGAS STEAM-OXYGEN PROCESS EFFICIENCY FOR 254.7 TJ/d (241.5 X 10^9 Btu/d) PIPELINE GAS PLANT USING MONTANA SUBBITUMINOUS COAL

	kg/h	lb/h
Reactor Coal (Dry Basis)	471,613	1,039,728
Fuel Coal (Dry Basis)	105,675	232,974
	577,288	1,272,702
	GJ/h	10^6 Btu/h
HHV Reactor Coal*	12,380.2	11,738.3
HHV Fuel Coal*	2,774.1	2,630.2
HHV Total Coal*	15,154.3	14,368.5
HHV of Product Gas	10,611.8	10,061.6

% Converted to Product Gas = 70.0

HHV of By-Products	602.2	570.9

% Converted to By-Products = 4.0

Total Overall Efficiency = 74.0

* At 26,251 kJ/kg (11,290 Btu/lb) on a dry basis.

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Table 32. SUMMARY OF OVERALL ENERGY BALANCE FOR 254.7 TJ/d (241.5×10^9 Btu/d) PIPELINE GAS FROM MONTANA SUBBITUMINOUS COAL BY THE HYGAS STEAM-OXYGEN PROCESS
[Basis: 289 K (60°F) Liquid H_2O]

	GJ/h	10^6 Btu/h	% of Coal Input
Heat In			
Process Coal	12,380.2	11,738.3	81.7
Dryer Fuel	394.7	374.2	2.6
Boiler Fuel	2,379.4	2,256.0	15.7
Total	15,154.3	14,368.5	100.0
Heat Out			
Heating Value of Pipeline Gas	10,611.8	10,061.6	70.0
Sensible Heat of Pipeline Gas	9.9	9.4	0.1
Heating Value of B-T-X	517.6	490.7	3.4
Heating Value of NH_3	59.0	55.9	0.4
Heating Value of Oxygasifier Char	221.6	210.1	1.5
Heating Value of Sulfur By-Product	25.6	24.3	0.2
Heating Value of Stretford Off-Gas	96.2	91.2	0.6
Sensible Heat of Stretford Off-Gas	18.0	17.1	0.1
Heat Loss to Air Cooling	1,066.1	1,010.8	7.0
Heat Loss to Cooling Water	1,542.1	1,462.7	10.2
Heat Loss to Boiler Stacks	318.6	302.1	2.1
Heat Loss With Dryer Flue Gas	270.4	256.4	1.8
Assumed Waste-Heat Recovery Losses	141.3	134.0	0.9
B-T-X Losses	28.7	27.2	0.2
Ammonia Losses	1.4	1.3	--
Heat Loss With H_2O Vapor With Stretford Effluent	41.1	39.0	0.3
Subtotal	14,970.0	14,193.8	98.8
Heat Unaccounted for	184.3	174.7	1.2
Total	15,154.3	14,368.5	100.0

264.4 TJ/d (250.7 X 10^9 Btu/d) SUBSTITUTE NATURAL GAS (SNG) FROM MONTANA SUBBITUMINOUS COAL BY THE CONSOLIDATION COAL CO. (CONSOL) CO_2 ACCEPTOR PROCESS

This process is the second of the two process designs for manufacturing methane from coal that were analyzed. It is a novel process, designed to avoid the use of high-purity oxygen by generating heat in the gasifier through the reaction of CO_2 with calcined dolomite (the acceptor). CO_2 is supplied by gasification reactions plus the injection into the gasifier of material scrubbed from the regeneration of off-gas. The process is somewhat analogous to the Steam-Iron Process in that a stream of material circulates between reactor and regenerator, with air supplying the basic source of oxygen for burning the gasifier residual char to provide heat for regenerating the acceptor. The process is currently being studied in a large pilot plant at Rapid City, S.D., sponsored by ERDA and the A.G.A.

In the gasifier, the crushed and dried coal is reacted with steam in the presence of CO_2 and MgO-CaO to generate methane and synthesis gas. Heat is supplied by the reaction of CO_2 with dolomite, which is showered through a fluidized bed of char operating at a gasification temperature of 1139 K (1590 °F). The raw gas is upgraded to pipeline quality in several steps following the reactor. The spent dolomite is thermally regenerated to MgO-CaO at about 1311 K (1900°F) in the regenerator. At present, this process is applicable only to lignite and low-rank Western coals. For this study, we based the process design on Montana subbituminous coal using material and energy balances around the gasifier and the regenerator supplied by Consol, whose assistance in this work is gratefully acknowledged. Figure 6 shows the processing steps required for the conversion of coal to methane pipeline gas. The compositions of the solid and gaseous streams corresponding to the streams in Figure 6 are shown in Tables 33 and 34. The processing steps may be grouped in the following way:

1. Coal storage and reclaiming
2. Coal grinding and drying
3. Coal preoxidation
4. Lockhopper feed system
5. Gasification reactor

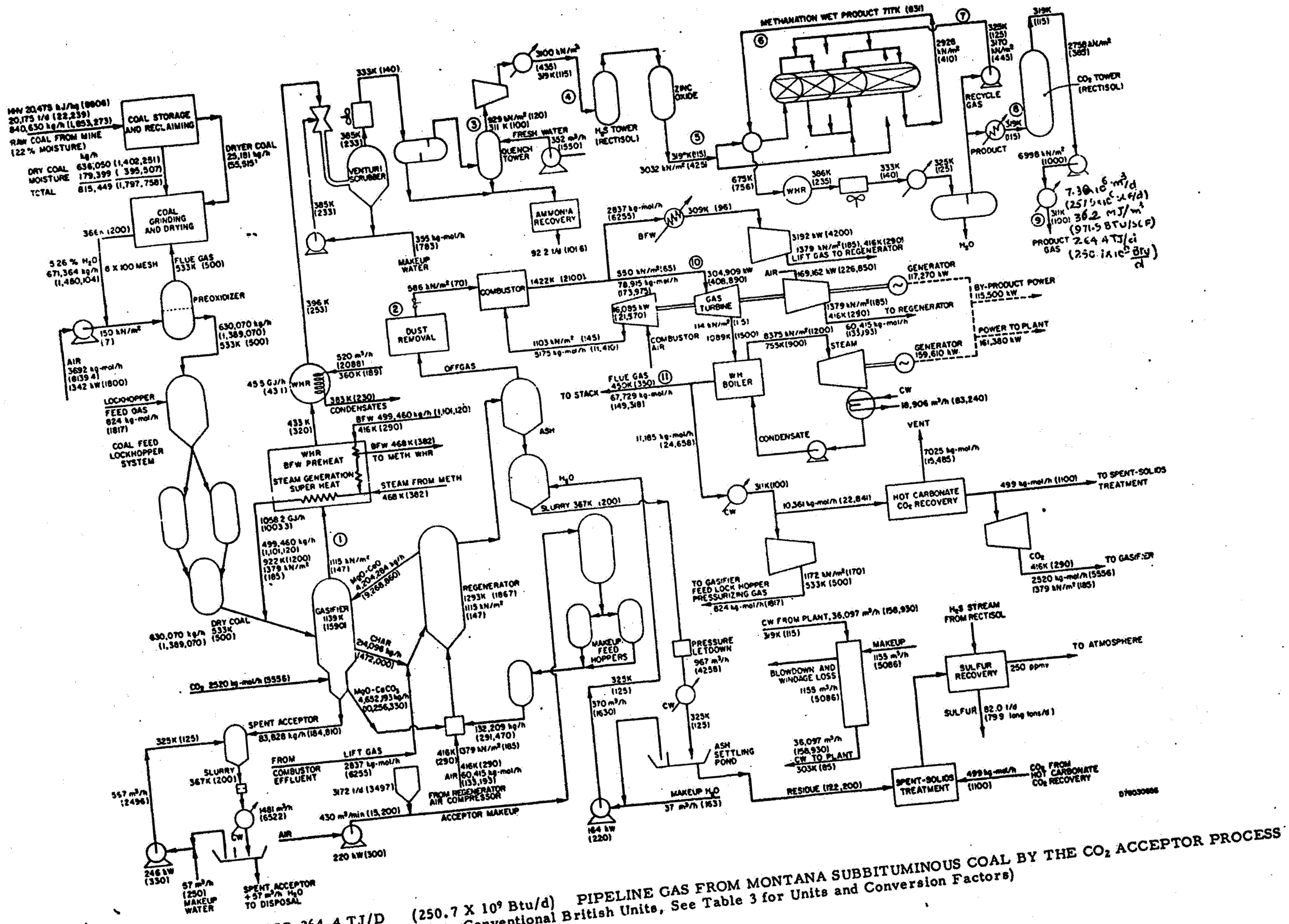

Figure 6. PROCESS FLOW DIAGRAM FOR 264.4 TJ/D (250.7 X 10^9 Btu/d) PIPELINE GAS FROM MONTANA SUBBITUMINOUS COAL BY THE CO_2 ACCEPTOR PROCESS (Numbers in Parentheses Have Conventional British Units, See Table 3 for Units and Conversion Factors)

Table 33. COMPOSITION OF SOLID STREAMS FOR 262.4 TJ/d (250.7 X 10^9 Btu/d) PIPELINE GAS FROM MONTANA SUBBITUMINOUS COAL BY THE CO_2 ACCEPTOR PROCESS

Stream	Ⓐ		Ⓑ		Ⓒ		Ⓓ	
Stream Name	Plant Coal Feed		Gasifier Feed		Char to Regenerator		Spent Char	
Component	wt %	kg/h	wt %	kg/h	wt %	kg/h	wt %	kg/h
C	67.70	443,903	67.62	426,053	76.03	162,772	5.87	3,252
H	4.61	30,227	4.38	27,597	0.50	1,066	-	-
O	18.46	121,041	18.68	117,697	-	-	-	-
N	0.85	5,573	0.85	5,419	-	-	-	-
S	0.66	4,328	0.67	4,221	0.55	1,175	5.58	3,094
Ash	7.72	50,619	7.79	49,083	22.92	49,083	88.55	49,083
Subtotal	100.00	655,691	100.00	630,070	100.00	214,096	100.00	55,429
Moisture		184,939	---	---	---	---	---	---
Total		840,630		630,070		214,096		54,429
Total lb/h		1,853,273		1,389,070		472,000		122,200

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Table 34. PROCESS STREAMS FOR 264.4 TJ/d (250.7 X 10^9 Btu/d) PIPELINE GAS (CH_4) FROM MONTANA SUBBITUMINOUS COAL BY THE CO_2 ACCEPTOR PROCESS

Stream No.	①		②		③		④		⑤		⑥	
Stream Name	Gasifier Overhead		Regenerator Off-Gas		Quench Effluent		Feed to H_2S Tower		Methanation Feed		Methanation Wet Product	
Temperature, K (°F)	1139 (1590)		1293 (1867)		311 (100)		319 (115)		319 (115)		717 (831)	
Pressure, kN/m² (psig)*	1115 (147)		1115 (147)		929 (120)		3100 (435)		3032 (425)		2928 (410)	
	mol %	kg-mol/h	mol %	kg-mol/h	mol %	kg-mol/h	mol %	kg-mol/h	mol %	kg-mol/h	mol %	kg-mol/h
Component												
CO	14.58	6,450	2.47	1,919	17.16	6,450	17.22	6,450	17.32	6,450	0.08	81
CO_2	3.27	1,445	32.90	25,512	3.84	1,445	3.85	1,442	3.65	1,361	0.89	856
H_2	56.17	24,855	0.04	34	66.11	24,855	66.36	24,855	66.74	24,855	4.24	4,082
H_2O	15.10	6,680	1.07	832	0.70	264	0.33	124	0.01	5	9.61	9,250
CH_4	10.16	4,497	--	--	11.96	4,497	12.01	4,497	12.07	4,497	84.64	81,490
NH_3	0.52	232	--	--	--	--	--	--	--	--	--	--
H_2S	0.02	9	< 0.01	3	0.02	9	0.02	9	--	--	--	--
N_2	0.18	78	63.49	49,239	0.21	78	0.21	78	0.21	78	0.54	520
SO_2	--	--	0.02	16	--	--	--	--	--	--	--	--
S_2	--	--	--	1	--	--	--	--	--	--	--	--
COS	--	--	0.01	5	--	--	--	--	--	--	--	--
Total	100.00	44,246	100.00	77,561	100.00	37,598	100.00	37,455	100.00	37,246	100.00	96,279
Total, lb-mol/h		97,545		170,992		82,889		82,574		82,112		212,259

Stream No.	⑦		⑧		⑨		⑩		⑪	
Stream Name	Methanation Recycle		Feed to CO_2 Tower		Product Gas		Combustion Gas to Expander		Stack Gas	
Temperature, K (°F)	325 (125)		319 (115)		311 (100)		1422 (2100)		450 (350)	
Pressure, kN/m² (psig)*	2825 (395)		2790 (390)		6996 (1000)		550 (65)		115 (2)	
	mol %	kg-mol/h	mol %	kg-mol/h	mol %	kg-mol/h	mol %	kg-mol/h	mol %	kg-mol/h
Component										
CO	0.09	59	0.09	12	0.10	12	--	--	--	--
CO_2	0.98	728	0.98	128	0.10	12	33.56	26,479	33.56	22,730
H_2	4.67	3,472	4.68	610	4.73	610	--	--	--	--
H_2O	0.46	342	0.33	43	--	--	1.06	839	1.06	718
CH_4	93.21	69,322	93.33	12,168	94.47	12,168	--	--	--	--
N_2	0.59	442	0.59	78	0.60	78	65.23	51,477	65.23	44,180
SO_2	--	--	--	--	--	--	0.03	24	0.03	20
O_2	--	--	--	--	--	--	0.12	96	0.12	81
Total	100.00	74,375	100.00	13,039	100.00	12,880	100.00	78,915	100.00	67,729
Total, lb-mol/h		163,969		28,745		28,396		173,975		149,318

* kN/m² expressed in absolute units, i.e., 101.33 kilonewtons/meter² = 1 atm = 14.696 psia.

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6. Dolomite regenerator
7. Gasifier effluent dust removal and syngas compression
8. H_2S removal
9. Methanation, drying, and final CO_2 removal
10. Regenerator off-gas power recovery system
11. Waste-water treatment
12. Sulfur recovery
13. Utilities and offsite facilities

1. Coal Storage and Reclaiming

The SNG plant is assumed to be located near a mine that will provide a minimum 25-year supply of coal. Raw coal is brought from the mine by the trucks to a storage area with the capability of storing a 30-day supply of material broken to 3.8 cm x 0 (1-1/2 in. x 0) size. Coal is stored in such a way that it is uniformly distributed in the storage pile. This plant requires a continuous supply of 20,175 t/d (22,239 short tons/d) of raw coal (22% moisture) from the mine. Because of the distribution of the coal in the coal storage pile, the feed to the plant has a more uniform composition even though it varies from the mine. About 3% of the feed coal is used as dryer fuel. Extra boiler fuel is not required because the steam and power requirements for the plant are met by the waste streams.

2. Coal Grinding and Drying

The plant coal feed is ground to 8 x 100 mesh size and dried to 5.26% moisture content in combination grinder-dryer mills, using coal fines plus flue gas from a fluidized-bed preoxidizer operating at 533 K (500°F). The ground and dried coal at 366 K (200°F) is conveyed to the preoxidizer with air at 150 kN/m^2 (7 psig).

3. Coal Preoxidation

Since subbituminous coal at process conditions forms coke in the gasifier near the feed point, preoxidation at the conditions described below will completely suppress the formation of coke.

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The preoxidizer operates at approximately atmospheric pressure and 533 K (500°F). Air is added to the vessel for burning a portion of the raw coal to maintain the preoxidizer temperature. The flue gas from the pre-oxidizer is used to dry coal, thus recovering thermal losses. The composition of the preoxidized coal is given below:

	wt %
Carbon	67.62
Hydrogen	4.38
Nitrogen	0.86
Oxygen	18.68
Sulfur	0.67
Ash	7.79
Total	100.00

HHV = 25,872 kJ/kg (11,127 Btu/lb)

4. Lockhopper Feed System

Because the reactor operates at a low pressure, 1115 kN/m^2 (147 psig), a lockhopper feed system, similar to the one used successfully in commercial Lurgi plants, is used to feed coal to the gasifier. Coal is fed to a lockhopper from a storage bin. The vessel is pressurized by the flue gas from the regenerator. The flue gas is vented to the atmosphere.

5. Gasification Reactor

The preoxidized coal at 533 K (500°F) is fed to the bottom of the gasifier, where the fixed carbon of the coal is gasified with steam after rapid devolatilization. The reaction steam is at 922 K (1200°F) and 1379 kN/m^2 (185 psig). The gasifier operates at 1115 kN/m^2 (147 psig) and 1139 K (1590°F). A circulating stream of dolomite supplies heat for the gasification reaction by the exothermic CO_2 acceptor reaction:

$$MgO\text{-}CaO + CO_2 \rightarrow MgO\cdot CaCO_3 \text{ (exothermic)}$$

In addition to the CO_2 generated in the gasifier, CO_2 from an external source is supplied for the acceptor reaction. About 61.8% of the carbon in the preoxidizer coal is gasified. The residual char and spent dolomite are transferred to the regenerator by the lift gas and the regenerator air, respectively. The composition of the gasifier effluent is represented by Stream 1 in Figure 6 and Table 34. It contains only a small portion of sulfur in the feed char as H_2S. The remaining sulfur goes to the regenerator either with spent dolomite as MgO-CaS or with residual char. Thirty-seven percent of the product gas methane is made directly in the gasifier; the remaining amount is produced by methanation. The feed quantities required for the gasifier operation are given below:

Component	kg/h	lb/h
Preoxidized Coal	630,072	1,389,070
Steam	499,460	1,101,120
Carbon Dioxide	110,912	244,520
Acceptor From Regenerator	4,204,284	9,268,860
Output	**kg-mol/h**	**lb-mol/h**
$CO + H_2$	31,305	69,016
CH_4	4,497	9,915

% carbon converted to gas = 61.8.

The CO_2 acceptor loses reactivity as it circulates between the gasifier and the regenerator. A portion of spent acceptor is withdrawn from the gasifier and replaced with fresh dolomite. The spent acceptor is cooled, slurried with water, and disposed to the residue settling pond.

The gasifier effluent contains a H_2/CO ratio of 3.85 and only 3.27% CO_2, indicating a sufficient amount of H_2 from the gasifier for the methanation of both CO and CO_2. Thus neither CO conversion nor CO_2 removal before methanation is required in this process. Some of the CO_2 in the effluent is removed with the H_2S in the Rectisol (Stage I) for H_2S removal, but the rest is needed to react with the excess hydrogen to make methane after all the CO is converted.

6. Dolomite Regenerator

The CO_2 acceptor regenerator operates at 1115 kN/m^2 (147 psig) and 1293 K (1867°F). The spent dolomite from the gasifier is calcined and the CO_2 acceptor reaction is reversed in the regenerator by the following reaction:

$$MgO\text{-}CaCO_3 \rightarrow MgO\text{-}CaO + CO_2 \text{ (endothermic)}$$

The heat for this reaction is supplied by burning the gasifier residual char with air. Ash and spent char are elutriated from the regenerator by the flue gas and collected via an external cyclone-lockhopper system. The ash is rich in calcium; at the reducing conditions in the regenerator, the calcium accepts most of the sulfur released from the acceptor. Consequently, three-fourths of the sulfur in the original feed coal is found in the regenerator ash stream. This ash cannot be disposed of as water slurry because H_2S would slowly evolve, creating a pollution problem. In order to avoid this problem, the ash is treated with CO_2 and H_2O in a simple chance reaction to convert the calcium and magnesium to the carbonate. Specifically the reactions are:

$$CaS + CO_2 + H_2O \rightarrow CaCO_3 + H_2S$$
$$CaO + CO_2 \rightarrow CaCO_3$$
$$MgO + CO_2 \rightarrow MgCO_3$$

The regenerator off-gas (Stream 2, Figure 6) at 1293 K (1867°F) and 1115 kN/m^2 (147 psig) has a large potential for power recovery, which is discussed below.

7. Gasifier Effluent Dust Removal and Syngas Compression

The gasifier effluent at 1139 K (1590°F) is cooled to 396 K (253°F), or 11 K (20°F) above the dew point, by the waste-heat recovery. Small dust particles (less than 5 microns) carried over from the cyclones in the gasifier are taken out in a jet venturi scrubber that has a high efficiency for removing particles 1 to 2 microns in size. It is necessary to remove dust particles from the gas before compression. The scrubber effluent is cooled to 311 K (100°F) by air and water cooling. The condensate containing dissolved ammonia and other gases is sent to waste-water treatment.

The clean gas at 311 K (100°F) and 930 kN/m^2 (120 psig) is compressed to 3204 kN/m^2 (450 psig) in order to facilitate acid-gas removal in a two-stage Rectisol system and methanation of CO. The higher pressure facilitates these steps.

8. H_2S Removal

The compressed syngas at 3204 kN/m^2 (450 psig) and 469 K (385°F) is cooled to 319 K (115°F) before going to the first stage of the Rectisol system for the selective H_2S removal. The H_2S-rich gas with 6.5% H_2S from the stripper is sent to the Claus sulfur recovery plant. The H_2S-free synthesis gas passes over a bed of ZnO to remove trace amounts of sulfur before methanation.

9. Methanation, Drying, and Final CO_2 Removal

The sulfur-free gas containing 17.32% CO and 3.65% CO_2 passes over a high-nickel catalyst where essentially all the CO and some of the CO_2 react with H_2 to form methane by the following reactions:

$$CO + 3H_2 \rightarrow CH_4 + H_2O$$
$$CO_2 + 4H_2 \rightarrow CH_4 + 2H_2O$$

This step increases the heating value of the gas to 36.2 MJ/m^3 (972 Btu/SCF) for the final product and reduces the CO content to 0.1%, as required by the pipeline gas specifications. As the methanation reaction is highly exothermic, it is essential to control the temperature to prevent catalyst deactivation. This is accomplished by using recycle-quench methanation with four stages, each with an inlet temperature of 561 K (550°F) and product-recycle to fresh-feed ratios such that the carbon monoxide content in the feed to each stage is about 4%. This limits the catalyst bed temperature to 755 K (900°F). This is analogous to the system used in the HYGAS design. The methanation wet product at 716 K (830°F) is cooled to 386 K (235°F) in a series of exchangers, utilizing the waste heat in preheating the first stage methanation feed to 561 K (550°F), as boiler feed-water preheat, and in steam generation. The product is further cooled to 325 K (125°F) by air and water cooling. The cooled gas is split into two streams; 85.1% of the gas is recycled to mix with fresh feed to the four methanation stages. The remaining gas is cooled to 319 K (115°F) and the residual CO_2 is removed.

The product gas from methanation contains 0.98% CO_2; this is reduced to 0.1% in the second stage of the Rectisol acid-gas removal system. The effluent (after CO_2 removal) is compressed to 7033 kN/m^2 (1005 psig), cooled to 311 K (100°F), and then dried to the pipeline standard of 112 kg/Mm^3 (7 lb/10^6 SCF) water in a glycol dryer; the dry gas leaves the plant at 6996 kN/m^2 (1000 psig).

10. Regenerator Off-Gas Power Recovery System

The regenerator off-gas at 1293 K (1867°F) and 1115 kN/m^2 (147 psig), which contains small quantities of CO, H_2, and sulfur compounds, also contains a large amount of energy which can be recovered in a combined cycle system, as is done in the Steam-Iron Process. The sulfur concentrations in the gas are low enough that the amount of SO_2 after combustion, based on total coal fuel value to the plant, is below the specified limit; thus, no sulfur removal system is specified.

After dust removal with cyclone separators and electrostatic precipitators, the off-gas is burned with a stoichiometric amount of air in a combustor. A pressure drop of 517 kN/m^2 (60 psig) is assumed for controlling the combustion. Also, a pressure drop of 5% in the combustor is assumed. About 3.5% of the combustor effluent is cooled, compressed to 1379 kN/m^2 (200 psig) and then used as lift gas. The remaining effluent at 1422 K (2100°F) and 550 kN/m^2 (65 psig) is expanded to 114 kN/m^2 (1.8 psig) and 1089 K (1500°F) in a gas turbine to recover power. A portion of this expansion power is used to drive the combustor and regenerator air compressors on the same shaft. The remaining power is converted to 117,270 kW of electric power in a generator. A maximum temperature range of 1255 to 1366 K (1800° to 2000°F) is probably the limit today, but gas turbines with inlet temperatures of 1589 K (2400°F) are expected to be available by the time this process becomes commercial. Since the steam-iron gas turbine operates at 1589 K (2400°F), 1422 K (2100°F) is certainly acceptable.

The expanded gas is used in a steam power cycle to generate 159,610 kW of electric power from steam at 8375 kN/m^2 (1200 psig) and 755 K (900°F) that was generated in a waste-heat boiler. A total of 276.9 MW of power is generated in addition to the shaft power used for air compression. From the total power generated, 161.4 MW is used within the plant for motor drives, etc., leaving 115.5 MW as by-product power.

The cooled combustor gas leaves the waste-heat boiler at 450 K (350°F). CO_2 is recovered from 13.1% of this gas by the hot carbonate process. This CO_2 is required in the gasifier and in the Chance reaction to recover H_2S from the regenerator residue. Another 1% of the stack gas is cooled, compressed, and used for the lockhopper feed system. The remaining cooled combustor effluent is vented to the atmosphere.

11. Waste-Water Treatment

The condensate from the gasifier effluent contains ammonia that is stripped with steam in a Chevron-type waste water treatment system. Ammonia is recovered as a by-product, and the water is used as cooling-tower makeup.

12. Sulfur Recovery

The H_2S-rich stream from Stage I of the Rectisol acid-gas removal system is combined with the H_2S stream from the Chance reaction. The concentration of H_2S in the combined gas is 15.4%. Thus, a Claus plant with a tail-gas clean-up system is used to recover 82.0 t/d (79.9 long tons/d) of sulfur. The vent gas from the Claus plant contains less than 250 ppm sulfur.

13. Utilities and Offsite Facilities

The conversion of coal to pipeline gas requires large quantities of steam, power and cooling water. These requirements are summarized in Tables 35-37. The coal-based plant is designed to be self-sufficient. The waste-heat recovery provides all the steam needs of the plant, and power recovered from the regenerator effluent (using the gas turbine and steam power recovery cycle) provides by-product power in addition to the power required in the plant.

The overall plant efficiency is 67.1% of which 64.0% is pipeline gas and the rest represents by-products such as power, ammonia, and sulfur. Table 38 presents the process efficiency. The overall energy balance summary (Table 39) gives the accounting for the remaining heat input from coal.

In summary, production of 264.4 TJ/d (250.7 X 10^9 Btu/d) pipeline gas from Montana subbituminous coal using the CO_2 Acceptor Process with the reactor operating at 1115 kN/m^2 (147 psig) and 1139 K (1590°F) requires 20,175 t/d (22,239 short tons/d) of 22% moisture coal. The design converts 64.0% of the HHV of coal to pipeline gas and 3.1% to by-products.

Table 35. SUMMARY OF STEAM REQUIREMENTS FOR 263.7 TJ/d (250 X 10^9 Btu/d) PIPELINE GAS FROM MONTANA SUBBITUMINOUS COAL BY THE CO_2 ACCEPTOR PROCESS

Steam Required			kg/h	lb/h	GJ/h	10^6 Btu/h
Pipeline Gas Plant	kN/m^2					
Gasifier Steam	1379 (185)	922 K (1200)	499,460	1,101,120	1864.8	1768.1
Rectisol Steam	345 (50)	421 K (298)	5,972	13,167	12.7	12.0
Hot K_2CO_3 System	791 (100)	443 K (337)	180,416	397,750	369.1	350.0
Ammonia Recovery	345 (50)	421 K (298)	62,510	137,810	132.5	125.6
Deaeration and Boiler Blowdown	159 (8)	386 K (235)	7,492	16,517	10.3	9.8
		Subtotal	755,850	1,666,364	2389.4	2265.5
Power Plant						
Turbine Steam for Power Cycle	8375 (1200)	755 K (900)	607,295	1,338,857	1917.3	1817.9
		Total	1,363,145	3,005,221	4306.7	4083.4
Steam Generated						
Waste-Heat Recovery From Gasifier Effluent			755,850	1,666,364	1058.2	1003.3
Waste-Heat Recovery From Methanation Effluent					1331.2	1262.2
Combustor Effluent Waste-Heat Recovery			607,295	1,338,857	1917.3	1817.9
		Total	1,363,145	3,005,221	4306.7	4083.4

Table 36. SUMMARY OF POWER REQUIREMENTS FOR 264.4 TJ/d (250.7 X 10^9 Btu/d) PIPELINE GAS FROM MONTANA SUBBITUMINOUS COAL BY THE CO_2 ACCEPTOR PROCESS

Component	kW	hp
Coal Storage and Reclaiming	1,939	2,600
Coal Grinding and Drying	7,457	10,000
Air Blower for Preoxidizer	1,439	1,930
Coal Feed Lock Hopper Gas Compressor	3,110	4,170
Compressor for CO_2 Feed to Gasifier	10,029	13,450
Lift Gas Compressor	3,132	4,200
Acceptor Makeup Air Blower	224	300
Venturi Scrubber, Effluent Air Cooling, and Quench Tower Feed Water	418	560
CO_2 Removal From Stack-Gas (Feed Compression and Solution Pumps) Hot Carbonate System	25,406	34,070
Methanation Feed Compressor and After Air Cooler	54,846	73,550
Methanation Recycle Compressor and Effluent Air Cooling	8,956	12,010
Product Gas Compressor	14,660	19,660
Rectisol System Motive Power	373	500
Rectisol System Refrigeration	9,396	12,600
Ash and Spent Acceptor Slurry Water Pump	410	550
Ammonia Recovery Condensates	1,417	1,900
Boiler Feed Water Pumps	447	600
Turbine Feed Water Pumps (for Power Generation)	2,446	3,280
Cooling Water Pumps	7,822	10,490
Miscellaneous	7,457	10,000
Total Power Required	161,384	216,420
Power Recovery From Reductor Off-Gas		
From Combustor Off-Gas Expansion	304,909	408,890
Less Combustor Air Compressor	(16,085)	(21,570)
Net Power Available From Gas Turbine	288,824	387,320
Less Power for Regenerator Air Compressor (on shaft)	(169,162)	(226,850)
Net for Power Recovery (on shaft)	119,662	160,470
Power Recovery (98%)	117,269	157,261
Power From Waste-Heat Steam Power Cycle	159,614	214,046
Total Power Generated	276,883	371,307
Electric Power to Plant	(161,384)	(216,420)
Total By-Product Power	115,499	154,887

Table 37. SUMMARY OF COOLING WATER REQUIREMENTS FOR 264.4 TJ/d (250.7 X 10^9 Btu/d) PIPELINE GAS FROM MONTANA SUBBITUMINOUS COAL BY THE CO_2 ACCEPTOR PROCESS

Component	m^3/h*	gpm*
Lock Hopper Gas Cooling After Compression	59	260
Methanation Feed Compressor	232	1,020
Product Gas Compressor	740	3,260
Methanation Effluent Cooler	522	2,300
Regenerator Air Compressor Interstage Cooling	5,167	22,750
Coal Feed Lock Hopper Gas Cooling Before and After Compression	114	500
CO_2 Removal System	7,230	31,830
Ash and Spent Acceptor Slurry Cooling	2,448	10,780
Ammonia Recovery Condensate Cooling	509	2,240
Rectisol System Cooling Water	170	750
Subtotal	17,191	75,690
Cooling Water for Power Generation Cycle	18,906	83,240
Total	36,097	158,930

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* Cooling water temperature 303-319 K (85°-115°F).

Table 38. PROCESS EFFICIENCY FOR PIPELINE GAS PLANT USING CO_2 ACCEPTOR PROCESS

Component	kg/h	lb/h
Reactor Coal (dry)	630,072	1,389,070
Fuel Coal (dry)	25,619	56,483
Total Coal (dry)	655,691	1,445,553
	GJ/h	10^6 Btu/h
HHV Total Coal	17,212.3	16,319.9
HHV of Product Gas	11,018.6	10,447.3
% Converted to Product Gas	64.0	
By-Product Power	415.8	394.1
% Converted to Power	2.4	
By-Product Ammonia + Sulfur	117.7	111.6
% Converted to By-Product	0.7	
Total Plant Efficiency	67.1	

Table 39. SUMMARY OF OVERALL ENERGY BALANCE FOR 264.4 TJ/Day (250.7 X 10^9 Btu/D) PIPELINE GAS FROM MONTANA SUBBITUMINOUS COAL BY THE CO_2 ACCEPTOR PROCESS
(BASIS: 289 K (60) LIQUID H_2O)

Heat In	GJ/h	10^6 Btu/h	%
Process Coal	16,696.7	15,831.0	97.0
Dryer Fuel	515.6	488.9	3.0
Total Heat In	17,212.3	16,319.9	100.0
Heat Out			
Pipeline Gas Sensible Heat	15.6	14.8	0.1
Pipeline Gas Heating Value	11,018.6	10,447.3	64.0
By-Product Power,	415.6	394.1	2.4
Sulfur,	31.3	29.7	0.2
Ammonia,	86.4	81.9	0.5
Heat to Cooling Water	2,514.4	2,384.0	14.6
Heat to Air Cooling	1,080.6	1,024.6	6.3
Heat Lost With Stack Gas	391.4	371.1	2.3
Heat Lost With Dryer Effluent	583.7	553.4	3.4
Heating Value + Sensible Heat of Ash and Spent Acceptor	141.5	134.2	0.8
Assumed Waste-Heat Recovery Losses	259.1	245.7	1.5
Heat Lost With Power Generation	23.3	22.1	0.1
Heat Loss With Sulfur Recovery Vent Gas	1.1	1.0	--
Assumed Loss in Power	28.6	27.1	0.2
Heat Loss in Regenerator	36.6	34.7	0.2
Heat Loss in Gasifier	55.7	52.8	0.3
Condensate Losses	91.5	86.8	0.5
Hot Carbonate Absorber Effluent	26.5	25.1	0.2
Heat Lost With Water Along With Ash Disposal	14.1	13.4	0.1
Subtotal	16,815.6	15,943.8	97.7
Heat Unaccounted for	396.6	376.1	2.3*
Total Heat Out	17,212.2	16,319.9	100.0

* The balance is 2.3% off, 1% of which can be accounted for from gasifier balance.

LIQUID FUELS FROM COAL

Introduction

The manufacture of hydrogen and methane from coal requires new technology and produces a new product for aviation fuel, which requires new methods of transportation and storage as well as new engines. The manufacture of liquid fuels from coal is directed toward developing a new source to supplement the fuel currently used in today's airplanes, which is derived from petroleum.

The proposal for this project requires that we set up process designs for kerosene, but not for jet fuel manufacture. Kerosene manufacturing has been interpreted as being a somewhat less stringent process than jet fuel manufacture, allowing us to design processes for increasing the supply of light distillate material or for making wide-cut kerosene or jet fuel.

Table 40 gives specifications for civil jet fuel and kerosene. We have considered jet-B or wide-cut kerosene as reasonable targets for the coal-derived liquid products required in this portion of the study.

Fischer-Tropsch (Synthol) Process

The conversion of coal to liquid fuels has been of interest for many years. Prior to World War II work was carried out in Germany on the hydrogenation of coal and the catalytic conversion of synthesis gas ($CO + H_2$) to liquid products. The catalytic conversion process is known by the generic term Fischer-Tropsch. This process converts synthesis gas(but not coal) directly to the desired products. Coal or some other fossil fuel is used to generate the synthesis gas. By the proper selection of catalysts and operating conditions, a very large variety of products can be made. These products include chemicals, subsitute natural gas (SNG), liquefied petroleum gas (LPG), gasoline, kerosene, diesel oil, and fuel oil. As is well known, a commercial plant (SASOL) has been in operation since 1955 in Sasolberg, South Africa. Synthesis gas, generated from coal, is converted in fixed-bed and entrained-bed reactors to yield a variety of products similar to those based on highly paraffinic mineral oils. The entrained-bed reactor is based on the Kellogg Synthol Process, which produces a wide spectrum of products but very little aromatics.

Table 40. CIVIL JET FUEL SPECIFICATIONS*

Properties	ASTM, D D-1655-70		IATA November 1969	
	JET A and JET A-1	JET B	Kerosene	Wide cut
Freezing Point,°F (°C) Max.	A- -36 (-38) A-1 -54(-48)	-56 (-49)	-58 (-50)	-58 (-50)
Distillation				
Initial Boiling Point,°F (°C)	-	-	-	-
10% Evaporated, °F(°C) Max.	400 (204)	-	400 (204)	-
20% Evaporated, °F(°C) Max.	-	290 (143)	-	290 (143)
50% Evaporated, °F(°C) Max.	450 (232)	370 (188)	450 (232)	370 (188)
90% Evaporated, °F(°C) Max.	-	470 (243)	-	470 (243)
Final Boiling Point,°F(°C) Max.	550 (288)	-	550 (288)	-
Residue, % vol. Max.	1.5	1.5	1.5	1.5
Loss, % vol. Max.	1.5	1.5	1.5	1.5
Flash Point, °F (°C) Min.	105 (41)	-	100 (38)	-
Max.	150 (66)	-	150 (66)	-
Reid Vapour Pressure, psi,Max	-	3	-	3
Specific Gravity, 60/60°F	0.7753-0.8299	0.7505-0.8017	0.775-0.830	0.751-0802
API Gravity	51-39	57-45	51-39	57-45
Viscosity at -30°F(-34.4°C)cSt.Max.	15	-	15	-
Aromatics, % vol. Max.	20	20	20	20
Olefins, % vol. Max.	-	5	-	-
Sulfur, % Wt. Max.	0.3	0.3	0.2	0.2
Mercaptan Sulfur,% Wt.Max.	0.003	0.003	0.001	0.001
or Doctor Test	Negative	Negative	Negative	Negative
Copper Strip Corrosion				
2 hrs. at 212°F (100°C)Max.	#1	#1	#1	#1
3 hrs. at 122°F	-	-	Open	Open
Silver Corrosion, Max.	-	-	Open	Open
Existent Gum, mg/100 ml. Max.	7	7	7	7
Thermal Stability-5 hr.at 6lb/hrs. 300/400°F(148.9/204.4°C)- ΔP in"Hg, Max.	12	12	3	3
Preheater Tube Deposit Rating	<3	<3	<3	<3
Heat of Combustion,Net BTU/lb.Min.	18,400	18,400	18,400	18,400
Aniline-Gravity Product Min.	-	-	5250	5250
Luminometer Number Min.	45	50	45	45
or Smoke Point, mm. Min.	25	-	25	-
or Smoke Volatility Index,Min.	-	54	-	54
or Naphthalenes, % vol Max.	3	-	3	-
Water Reaction,vol.change,ml.Max.	1	1	1	1
Interface Rating, Max.	+1b	+1b	1b	1b
Electrical Conductivity, picomho/m at time, place and temperature of delivery into aircraft	-	-	50-300	50-300
Total Acidity, mg KOH/g Max.	0.1	0.1	0.1	0.1
Additives				
Antioxidant	Option	Option	Option	Option
Metal Deactivator	Option	Option	Option	Option
Corrosion Inhibitor	By Agreement	By Agreement	By Agreement	By Agreement
Static Dissipator	By Agreement	By Agreement	Required	Required
Others	By Agreement	By Agreement		

* Source: Gardner, L. and Whyte, R. B., "Jet Fuel Specifications." Paper No. AGARD-CP-84-71 presented at Advisory Group for Aerospace Research and Development Conference Proceedings No. 84 on Aircraft Fuels, Lubricants, and Fire Safety, n.d.

Liquids derived from the direct conversion of coal by hydroliquefaction, pyrolysis, and extraction are composed mainly of aromatics, naphthenes, and cycloparaffins. Specifications for jet fuels derived from petroleum show a maximum aromatic content of 20% and smoke points of 25 or better (Table 40). Aromatics and naphthenes, which are produced by hydrogenation of aromatics, have low smoke points.

Thus, the Synthol Process is of interest in this application because the products would meet the specifications for jet fuel and kerosene better than products of coal liquefaction. We contacted the M.W. Kellogg Co. for information on the application of the Synthol Process to the production of wide-cut jet fuel. The Synthol plant built by Kellogg for SASOL was designed to produce gasoline and naphtha, which are considered the most desirable products. Information on the Synthol Process was obtained from Dr. B. G. Mandelik of the M.W. Kellogg Co. Kellogg believes that the Synthol Process can convert synthesis gas ($CO + H_2$) to the desirable products at an overall efficiency of as high as 80%. Based on the generation of synthesis gas by IGT's U-GAS Process, and using the same process coal feed and gasifier operation as in the hydrogen case, we calculated a production of 53,708 kg-mole/h (118,406 lb-mole/h) of synthesis gas containing 68.2% H_2 + CO in the ratio 2.4:1 as desirable for Synthol feed. The overall efficiency for generating this gas, converting it to jet fuel, and other products via the Synthol Process, and supplying some additional steam and power is 57.5%. This efficiency is significantly less than the overall efficiency indicated for the newer coal conversion processes. The Synthol section produces 3289 m^3/d (20,690 bbl/d of kerosene-type jet fuel plus 2068 t/d (2280 short tons/d) of other products.

Consol Synthetic Fuel Process

The newer coal liquefaction processes include the Consolidation Coal Co. Synthetic Fuel (CSF) Process, Solvent-Refined Coal (SRC), H-COAL,® COED, and Synthoil. None of these processes are in commercial operation; however, all but Synthoil have been carried into the pilot plant stage.

We selected the CSF Process as the basic coal conversion process for this study. The liquid products from this process are reported to be similar to those produced by Hydrocarbon Research Inc.'s (H-OIL) Process when using coal extract. The CSF Process produces some aromatics and less naphthenes than the HRI Process, but the sum of aromatics plus naphthenes is about the same, around 90%, with a very low paraffin content. The overall efficiencies of the processes are similar, around 70%.

This favorable comparison with H-OIL processing of coal extract plus the availability of a fairly detailed process design for the CSF Process design[3] prompted the choice of the latter as a basis for the coal-to-liquid fuel process in the present study. The design study[2] referred to concluded that the CSF Process as contemplated is technically feasible.

At the beginning of this project, we decided that we would not study the effect of different coals on a given process. However, we would select the coal or coal type that appeared to be the most advantageous for a particular process. Within the scope of the project, the choice is made between Eastern and Western coals. For purposes of comparability it is desirable to keep coal variation to a minimum. For the manufacture of hydrogen and methane, Montana subbituminous coal was selected because of its greater reactivity, abundance, low-sulfur content, nonagglomerating characteristics during gasification, and its lower cost compared to Eastern coals.

The process design for the CSF Process is based on Pittsburgh seam coal from Consolidation Coal Co.'s Ireland mine, which was used in the development of the process. An analysis of this coal is given in Table 2 in the first section of this report.

Based on published summaries of studies of the H-COAL Process, the use of Illinois bituminous coal gives more favorable results, from the standpoint of both overall efficiency and economics, than Wyoming subbituminous coal.[4] As the Illinois coal is more similar to Pittsburgh seam coal than to Wyoming subbituminous coal, it appears that Eastern bituminous coals should be more favorable for liquefaction than Western subbituminous coals. Therefore, the use of Eastern coal is a reasonably good basis for the CSF Process.

Description of CSF Process

The basic CSF Process design used for this section of the study is shown diagramatically in Figure 7. Tables 41 and 42 show solid, liquid, and gaseous streams for this design. IGT has utilized the products of the coal conversion plant in a refinery add-on to produce C_5+ aromatic gasoline..

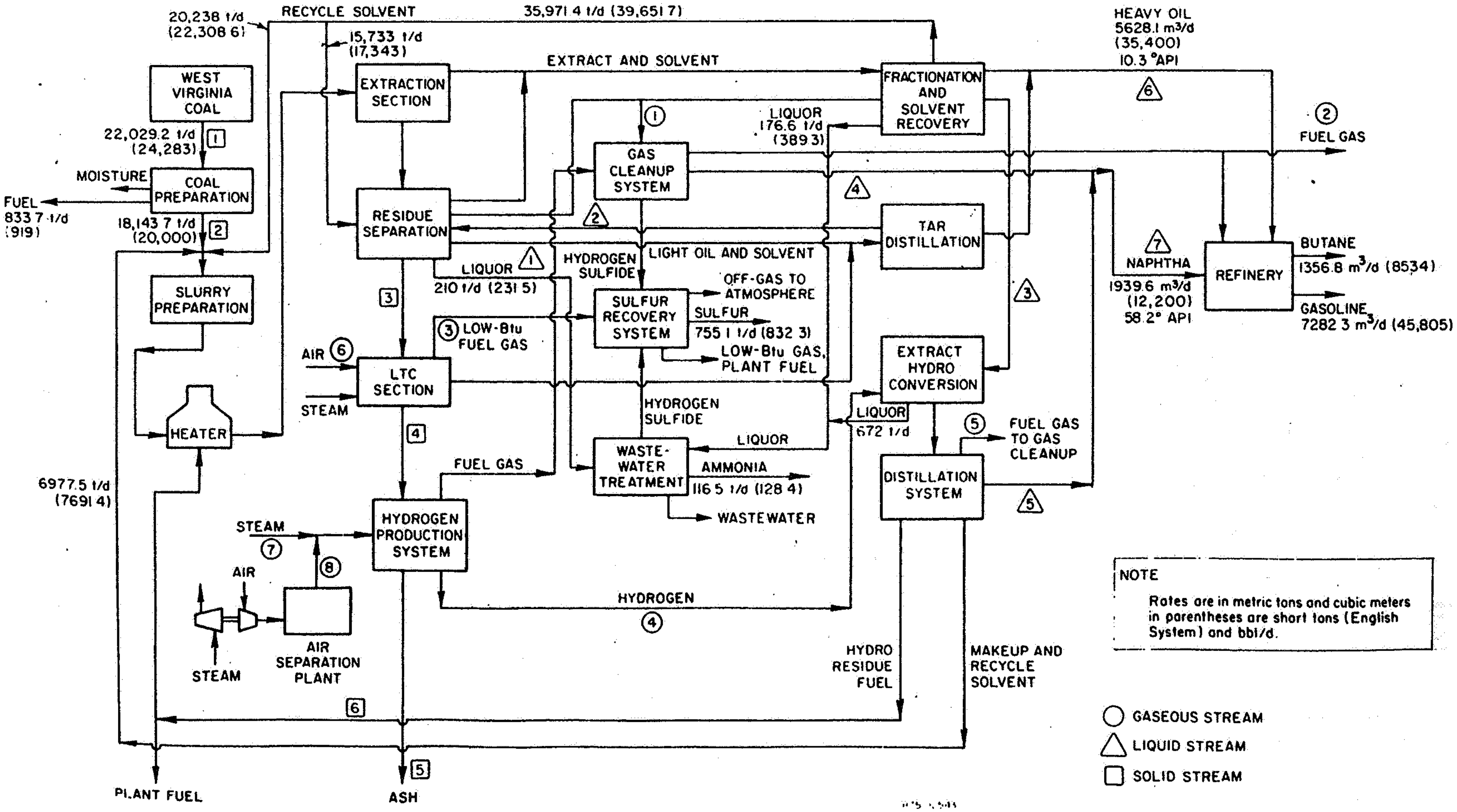

Figure 7. GASOLINE FROM COAL, 7282.3 m^3/d (45,800 bbl/d) CONSOL SYNTHETIC FUEL PROCESS – LIQUEFACTION OF COAL BY CONSOL SYNTHETIC FUEL PROCESS WITH ADD-ON REFINERY TO YIELD AROMATIC GASOLINE

Table 41. CONSOL SYNTHETIC FUEL PROCESS, SOLID AND LIQUID STREAMS

	Solid Stream					
Stream No.	1	2	3	4	5	6
Stream Description	Total Coal	Moisture-Free Coal to Extraction	Solvent, Extract, and Residue to Tar Distillation	Char From Low-Temperature Carbonization to Gasifier	Gasifier Ash	Hydrodistillation Residue
kg/h	917,876	755,980	280,000	255,407	121,517	87,755
lb/h	2,023,580	1,666,670	617,300	563,080	207,900	193,467

	Liquid Streams						
Stream No.	1	2	3	4	5	6	7
Stream Description	Light Oil Solvent	Solvent Plus Tar	Extract From Solvent Recovery	Butane From Gas Plant	Naphtha Hydrodistillate	Heavy Oil Product	Product Naphtha
kg/h	27,571	497,927	482,214	3421	56,888	233,335	60,309
lb/h	60,783	1,097,742	1,063,100	7541	125,417	514,417	132,958

Table 42. CONSOL SYNTHETIC FUEL PROCESS, GAS STREAMS

Stream No.	①	②	③	④	⑤	⑥	⑦	⑧
Stream Description	Gas Cleanup Feed (Mixed Streams)	Product Fuel Gas 1653 kN/m² (225 psig)	Low-Btu Fuel Gas 143 kN/m² (16 psig)	Process Hydrogen 7686 kN/m² (1100 psig)	Hydrogen Let Down Gas Mixed Streams 2170 & 10443 kN/m² (300 & 1500 psig)	Air to Low Temp. Carbonization 239 kN/m² (20 psig)	Steam to H_2 Plant Gasifier 8720 kN/m² (1250 psig)	Oxygen to Gasifier 8375 kN/m² (1200 psig)
CO	2.48	7.11	8.42	--	--	--	--	--
CO_2	15.82	--	12.18	--	--	--	--	--
H_2	4.95	30.91	--	96.02	34.07	--	--	--
H_2O	--	--	5.02	--	--	--	100.00	--
CH_4	20.79	38.73			39.08	--	--	--
C_2H_6	9.34	11.72	4.81†	3.03	11.07	--	--	--
C_3H_8	3.09	8.08		--	8.47	--	--	--
C_4H_{10}	6.80	9.02	--	--	--	--	--	--
H_2S	36.73	--	0.79	--	3.39	--	--	--
N_2	--	3.43	68.45	0.95	3.92	79.00	-- (Ar)	0.54
O_2	--	--	0.33	--	--	21.00	--	99.46
Total	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00
kg-mol/h	877.4	4046	6882	14,577	3543.4	5936	8,942	3754
lb-mol/h	1934.3	8919	15,173	32,137	7811.8	13,087	19,715	8,277

† Assumed C_2H_6.

This operation is discussed in more detail below. The products from the CSF plant proper are based on the conversion of 21,195 t/d (23,364 short tons/d as received, 20,000 dry) of process coal to the following products —

Heavy Fuel Oil:	10.3 °API, 5628 m^3/d (35,400 bbl/d)
Naphtha:	58 °API, 1939.6 m^3/d (12,200 bbl/d)
High-Btu Gas:	78.56 TJ/d (74.49 X 10^9 Btu/d), 34.7 MJ/m^3 (931.5 Btu/SCF)

The CSF Process extracts a de-ashed fuel from the coal that is hydrogenated to desulfurize the fuel and also to saturate the aromatics, in order to provide a solvent for the extraction step. Hydrogen is added to the coal by the hydrogen donor solvent, which is coal-derived, and regenerated and recycled within the process. The process description follows.

Coal Preparation and Extraction

The boiler fuel requirement is 834 t/d (919 short tons/d). The raw process coal (Pittsburgh seam coal), 21,195 t/d (23,364 short tons/d), is crushed in the hammer-mills to a size of about 5.08 to 0.95 cm (2 to 3/8 in.), then partially dried by contact with the flue gas. The partially dried coal is dried further in the fluid-bed dryers. Fines smaller than 14 mesh are recovered in multiple-stage cyclones and bag filters. The crushed coal is combined with the recovered fines and heated to 505 K (450°F) in fluidized-bed dryers to remove the remaining moisture. The preheated coal is then slurried with a coal-derived solvent and pumped at 1136 kN/m^2 (150 psig) through a tubular furnace, where it is heated to the extraction temperature of 680 K (765°F). Extraction occurs principally in a staged extraction vessel. The hot vapor from the extractor is sent to the solvent-recovery area and the slurry phase is sent to a residue-separation section.

Residue Separation and Solvent Recovery

The undissolved coal residue is removed from the slurry in the residue-separation section by two-stage hydroclones. Overflow from the first stage goes to the solvent-recovery section and the underflow passes to a second stage, the wash stage. The overflow from this stage is fed back to the first stage. The underflow is sent to the low-temperature carbonization system.

Solvent recovery is divided into two sections. After the vapor from the extraction section is condensed, the gaseous stream is sent to a gas cleanup section and the recovered solvent is returned to slurry mix tanks. The hydroclone overflow from the residue separation is fractionated in a vacuum still. Light oil and light spent solvent are sent to low-sulfur fuel production.

The heavy oil (including spent solvent and fuel oil) is taken overhead from the fractionator, and a heavier cut from a side stream provides most of the recycle hydrogen donor solvent for the extraction section. The bottoms, which contain extract, residue, and tar are sent to the extract hydroconversion system.

Low-Temperature Carbonization

The hydroclone underflow from the residue separation is pumped to the low-temperature carbonizer (LTC) where it is reacted with steam and air. The overhead product is quenched, and a gas stream and a solvent/tar stream are separated out. The solvent/tar stream is delivered to the tar-distillation section, and the gas stream is used as a plant fuel after sulfur removal. Char from the LTC section is delivered to the Bituminous Coal Research (BCR) gasification system for hydrogen production.

Tar Distillation and Extract Hydroconversion

The heavy liquids from the LTC section are vacuum-distilled in the tar-distillation section. The overhead product is heavy oil. The bottoms are sent to the residue-separation area. The extract from the solvent-recovery section is hydrotreated to produce the donor solvent and product oil. Extract hydrogenation is done in four stages that operate at 20,786 kN/m^2 (3000 psig) and 700 to 714 K (800° to 825°F) in the presence of a cobalt-molybdenum-nickel catalyst. The overhead vapors are cooled to separate the hydrogen and light oils. The recovered hydrogen is compressed and recycled back to the reactors. The hydrotreated liquid product is flashed to 136 kN/m^2 (5 psig). The fuel gas is sent to the gas treatment plant, and gas liquor is sent to the wastewater treatment plant to recover ammonia and hydrogen sulfide. The hydrotreated liquid product is stabilized by removing C_4 and then fractionated. In the fractionator, the hydrogen-donor solvent is separated from the light-oil product. The hydrogen-donor solvent is sent to the slurry system for makeup solvent, and the light product oil is delivered to the refinery to produce gasoline.

Gas Treatment and Sulfur Recovery

Fuel gas is produced at various sections of the plant. This gas is treated in the amine system to remove carbon dioxide and hydrogen sulfide. Some of the fuel gas is used as a fuel for the plant and refinery operation, and the remaining amount is a by-product for sale. The hydrogen sulfide stream from the amine system is passed through the sulfur-recovery system, which consists of a modified Claus plant and a Beavon tail-gas plant.

Hydrogen Production

Hydrogen is made from synthesis gas generated by the BCR Process. Char from the LTC is gasified by steam and oxygen in the BCR suspension-type gasifier. The raw gas is passed through the quench system, the CO-shift system, the acid-gas removal section, and a CO removal step.

Tables 43-45 summarize the utility requirements for the CSF Process. The plant is essentially self-contained except that in the original design,[3] 61,180 kW of electric power is purchased. In order to keep this design on the same basis as the others in this study, we have added equipment to supply this power onsite. This requires a coal-fired boiler and turbine-generator plus an added cooling water supply. Additional coal is calculated using a basis of 35% overall conversion efficiency from coal to electric power. The original design calls for steam turbine drives for oxygen supply and hydrogen compression. We have added the requirements for these services to the power summary for completeness only because the original design includes steam to drive the turbines as part of the utility balances. Table 46 is the overall energy balance.

Kerosene — Jet Fuel Manufacture

The products of the CSF Process, heavy fuel oil, naphtha, and high-Btu gas, do not meet the product specification for kerosene or jet fuel. Two ways are presented for increasing production of these fuels through coal liquefaction:

1. Modify the CSF products to get the desired ones. This will require a hydrocracking step and the generation of large amounts of hydrogen, which means expensive equipment and lowering the overall efficiency.
2. An alternate method involves the conversion of the products of coal liquefaction to gasoline and importing this material into the petroleum refinery octane pool. This gasoline would then allow release of

material in the kerosene boiling range from the catalytic reforming feedstock and, in this way, increase the supply of jet fuel components. Because the products of coal liquefaction are highly aromatic, they should produce a high-octane material, a desirable contribution to the refinery gasoline pool. Such aromatics are not desirable in jet fuel, as discussed above, but are satisfactory as a gasoline component.

Even if all the aromatics were hydrogenated to naphthenes (saturated cyclic compounds), the material might still be marginal in meeting smoke point specifications. A sample of decalin ($C_{10}H_{18}$) had a smoke point value of 21.6, which is below the civil jet fuel specifications of 25, but above the military JP-5 requirement of 19. However, this is one of the lighter naphthenic compounds. Heavier ones might give lower smoke point readings.

Gasoline From Coal for Export to Petroleum Refineries

The heavy fuel oil stream can be hydrocracked and combined with the naphtha to yield a C_5-478 K (C_5-400°F) gasoline that could be sold to conventional refineries to meet part of their gasoline outputs.

The first step in the process is the manufacture of gasoline from the products of the coal conversion plant in a refinery attached to the plant. One of IGT's engineers, Mr. R. Donald Oberle, a petroleum engineer with many years' experience in the petroleum industry, has made the calculations for the conversion of CSF liquid products to gasoline and also for the modification of a conventional petroleum refinery operation resulting from the importation of the coal-derived gasoline.

The coal product refinery is shown as an appendage to the CSF plant in Figure 7. More details are given in Figure 8. All three of the products from the coal plant are utilized. This process scheme was available to our study from earlier IGT work. This particular section does not represent as detailed a process design as the other parts of the work but represents conventional refinery practice. Overall results and net fuel requirements are based on generalized requirements for refinery operations.

The naphtha, 1939.6 m^3/d (12,200 bbl/d), is distilled to give three streams. About two-thirds of the feed yields a 355 to 478 K (180° to 400°F) stream for catalytic reforming. The C_5-355 K (C_5-180°F) is routed to the motor gasoline pool. Butane, 217.8 m^3/d (1370 bbl/d), is a fuel by-product.

Table 43. SUMMARY OF STEAM REQUIREMENTS FOR CONVERSION OF COAL TO 378.6 TJ/d (359 X 10⁹ Btu/d) LIQUIDS PLUS HIGH-Btu GAS BY THE CONSOL SYNTHETIC FUEL PROCESS

Steam Required	kg/h	lb/h
1250 psig, 1200°F		
BCR Gasifier	161,025	355,000
Used to Preheat Oxygen, then Depressured to 600 psia, 486°F	19,050	42,000
	180,075	397,000
1200 psig, 900° F		
Turbine Steam	237,180	523,900
600 psig, 750° F		
Low Temperature Carbonization	27,530	60,700
Hydrogenation Letdown and Absorption	42,320	93,300
Compression and Gas Treating Plant	29,760	65,600
Hydrogen Manufacture	342,915	756,000
Hydrogen Compression	96,615	213,000
Waste-water Treating	9,115	20,100
	548,255	1,208,700
150 psig, 366° F		
Low Temperature Carbonization	5,490	13,100
Tar Distillation	180	400
Gas Treating Plant	8,480	18,700
Hydrogen Manufacture	11,570	25,500
Hydrogen Compression	905	2,000
Wastewater Treating	78,015	172,000
	104,640	230,700
25 psig, 267° F		
Gas Treating Plant	75,750	167,000
Hydrogen Manufacture	45,360	100,000
	121,110	267,000
Steam Generated		
1200 psig, 1200° F		
Fired Boiler	180,075	397,000
1200 psig, 900° F		
Fired Boiler	237,180	522,900
Steam Generated	kg/h	lb/h
600 psig, 750° F		
Extraction Hydroconversion	130,770	288,300
Waste Heat Recovery	116,120	256,000
Fired Boiler	301,365	664,400
	548,255	1,208,700
150 psig, 366° F		
Solvent Recovery	58,150	128,200
Sulfur Plant	21,680	47,800
Fired Boiler	24,810	54,700
	104,670	230,700
25 psig, 267° F		
Solvent Recovery	171,590	378,300
Sulfur Plant	57,880	127,600
	229,470	505,900

Table 44. SUMMARY OF POWER REQUIREMENTS FOR CONVERSION OF COAL TO 378.6 TJ/d (359 X 10^9 Btu/d) LIQUIDS PLUS HIGH-Btu GAS BY THE CONSOL SYNTHETIC FUEL PROCESS

Section	kW	hp
Feed Preparation	12,202	16,363
Extraction	2,109	2,828
Residue Separation	1,273	1,707
Solvent Recovery	493	661
Low-Temperature Carbonization	1,710	2,293
Tar Distillation	237	318
Extract Hydroconversion	1,179	1,581
Hydrogenation Letdown	3,310	4,439
Hydrodistillation	1,533	2,056
Gas Recovery and Treating Plant	960	1,287
Hydrogen Manufacture	23,100	30,978
Hydrogen Compression	19,761	26,500
Oxygen Supply	60,983	81,780
Waste-Water Treating	104	139
Sulfur Plant	2,116	2,838
Steam Generation	4,117	5,521
Water Treating	800	1,073
Cooling Water	7,130	1,561
Miscellaneous	1,200	1,609
Total Power	144,317	193,532

Table 45. SUMMARY OF COOLING WATER REQUIREMENTS FOR CONVERSION OF COAL TO 378.6 TJ/d (359 X 10^9 Btu/d) LIQUIDS PLUS HIGH-Btu GAS BY THE CONSOL SYNTHETIC FUEL PROCESS

Section	m^3/h	gpm*
Residue Separation	727	3,200
Solvent Recovery	456	2,009
Low-Temperature Carbonization	1,043	4,590
Tar Distillation	3	13
Extract Hydroconversion	2,091	9,207
Hydrogenation Letdown	3,500	15,410
Hydrodistillation	40	175
Gas Recovery and Treating Plant	3,647	16,057
Hydrogen Manufacture	20,123	88,600
Hydrogen Compression	3,589	15,800
Waste-Water Treating	2,498	11,000
Sulfur Plant	368	1,620
Power Plant	7,371	32,452
Total	45,456	200,133

* 302°K to 316°K temperature rise (85° to 110°F).

Table 46. OVERALL ENERGY BALANCE FOR PROCESS STREAMS, CONSOL SYNTHETIC FUEL PROCESS
(Basis: 289 K (60°F), Liquid H_2O)

	TJ/d	10^9 Btu/d	10^6 Btu/h	% of Coal Input
In				
Process Coal	533.25	505.60	21,067	96.2
Fuel Coal	20.97	19.88	828	3.8
Total	554.22	525.48	21,895	100.0
Out				
High-Btu Gas	78.56	74.49	3,104	14.18
Naphtha	65.86	62.45	2,602	11.88
Fuel Oil	233.98	221.85	9,243	42.22
NH_3	2.62	2.48	103	0.47
Sulfur	7.00	6.64	277	1.26
H_2 Loss	1.32	1.25	52	0.24
Heat to Coolant	63.47	60.18	2,508	11.45
Heat Losses (by Difference)	101.39	96.14	4,006	18.30
Total	554.20	525.48	21,896	100.00

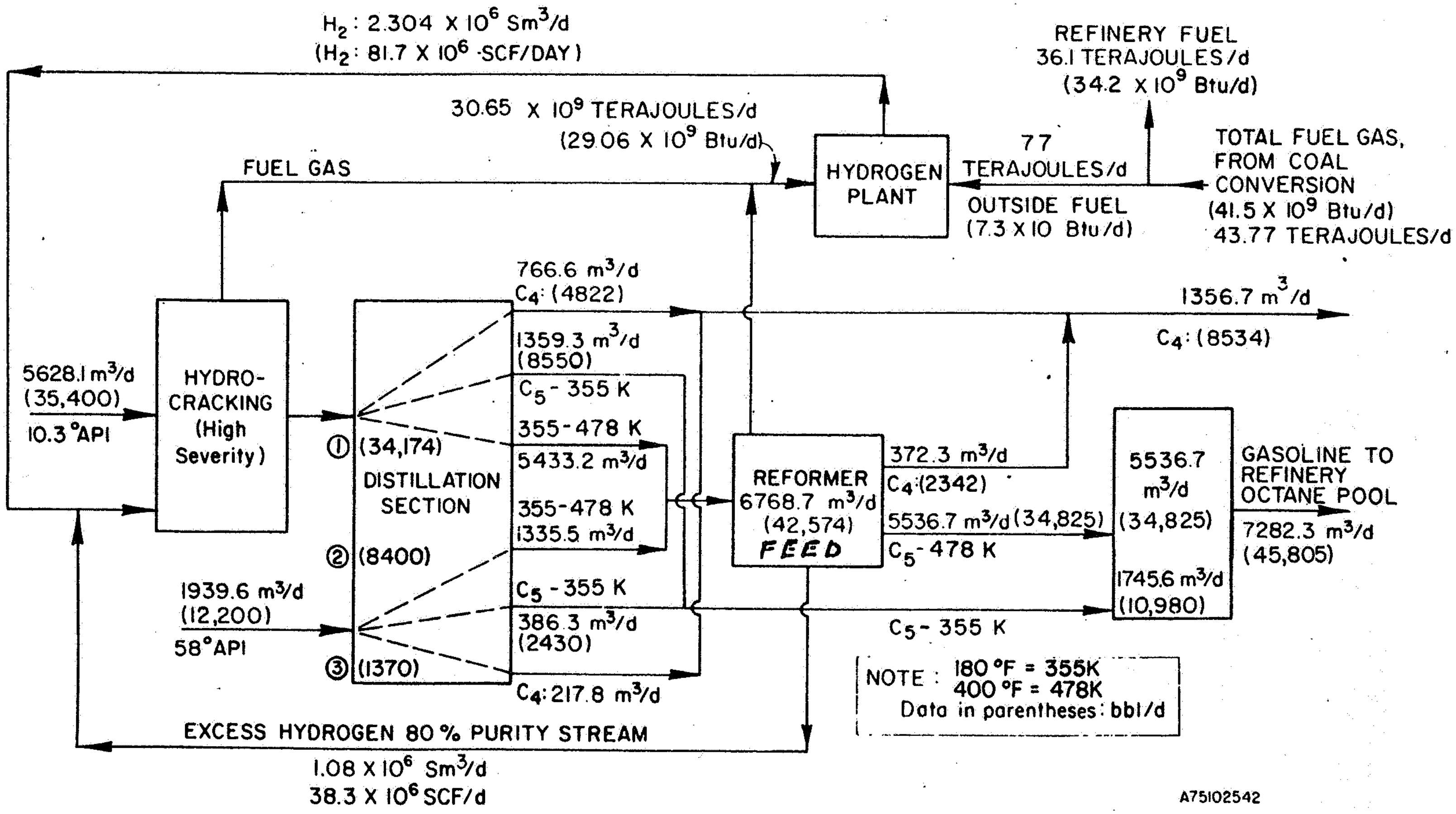

Figure 8. REFINERY BLOCK DIAGRAM – 7282.3 M^3/D (45,800 bbl/D) PLANT FOR UPGRADING LIQUID FUELS FROM COAL TO AROMATIC GASOLINE

The 10.3 °API, 478 to 672 K (400° to 750°F) boiling-range stream must undergo considerable hydrogenation to make it suitable for gasoline. It is first hydrocracked at high severity. The products are then separated into the same three boiling-range cuts as the naphtha. Combined 355 to 478 K (180° to 400°F) streams, 6768.7 m^3/d (42,574 bbl/d), are sent to catalytic reforming to increase the gasoline octane number. Reformate and C_5-355 K (C_5-180°F) streams are combined to yield 7282.4 m^3/SD (45,805 bbl/SD) of high-octane gasoline for export to petroleum refinery gasoline pools.

About 56% of the high-Btu gas from the CSF plant is utilized in the gasoline plant as fuel and feed for manufacturing hydrogen required for hydrocracking the heavy oil. Butane from hydrocracking and reforming is sold as a by-product. Products after the refining operation are —

Gasoline	C_5-478 K (C_5-400°F), 7282.3 m^3/d (45,805 bbl/d)
Butane	1356.8 m^3/d (8534 bbl/d)
High-Btu Gas	34.79 X 10^6 TJ/d (32.99 X 10^9 Btu/d)

The high-Btu gas is 44% of that generated by the CSF Process; the rest is used as fuel in the manufacture of the gasoline. Table 47 gives the product summary.

Light Distillate for Jet Fuel

The increase in jet fuel as a result of coal liquefaction is obtained from a modification of conventional refinery operation by reducing the amount of light distillate (422 to 478 K, 300° to 400°F) fed to the catalytic reformer (Figure 9). Calculations are based on a 23,848 m^3/SD (150,000 bbl/SD) petroleum refinery in which the feed to the catalytic reformer is 6677.5 m^3/SD (42,000 bbl/SD). Production of catalytic reformate is 5176.6 m^3/SD (32,560 bbl/SD). By reducing this feed by 1431 m^3 (9000 bbl), the reformate production, in turn, is reduced by 1197 m^3 (7530 bbl). The reduction of petroleum-derived reformate is made up by coal-derived reformate. Viewed another way, the reduction in demand for reformer gasoline releases light distillate feed as a suitable component of jet fuel. In this preliminary analysis, we assumed that the C_5+ gasoline derived from coal is a satisfactory material that can be imported to compensate for the reduction in reformate from the catalytic reformer. Since 7282 m^3/SD (45,805 bbl/SD) aromatic gasoline is available from the coal liquefaction complex, this will supply six 23,848 m^3/SD

Table 47. PRODUCTS SUMMARY FOR CONVERSION OF COAL BY THE CONSOL SYNTHETIC FUEL PROCESS PLUS UPGRADING OF LIQUID PRODUCTS TO GASOLINE REFORMATE

Products	Amount/Stream-Day	Amount/Yr	Assumed Unit Value at Coal Conversion Plant	Annual Value, $
C_5 – 400°F Reformate	7282.4 m^3 (45,805 bbl)	2,392,590 m^3 (15,048,940 bbl)	--	--
High-Btu Gas	34.79 TJ (32.99 X 10^9 Btu)	11.428 TJ (10,837 X 10^9 Btu)	$1.61/GJ ($1.70/10^6 Btu)	18,422,900
Butane	1,356.8 m^3 (8534 bbl)	445,708 m^3 (2,803,419 bbl)	3.17¢/liter (12¢/gal)	14,129,200
Sulfur	754.9 metric tons (743 long tons)	247,992 metric tons (244,075 long tons)	$9.84/metric ton ($10/long ton)	2,440,750
Ammonia	116.5 metric tons (128.4 tons)	38,265 metric tons (42,179 tons)	$38.6/metric ton ($35/ton)	1,476,270
Total By-Product Value				36,469,120
Increased Light Distillate From Refineries Importing C_5 – 400°F Reformate	8702.96 m^3 (54,740 bbl)	2,858,924 m^3 (17,982,090 bbl)	--	--

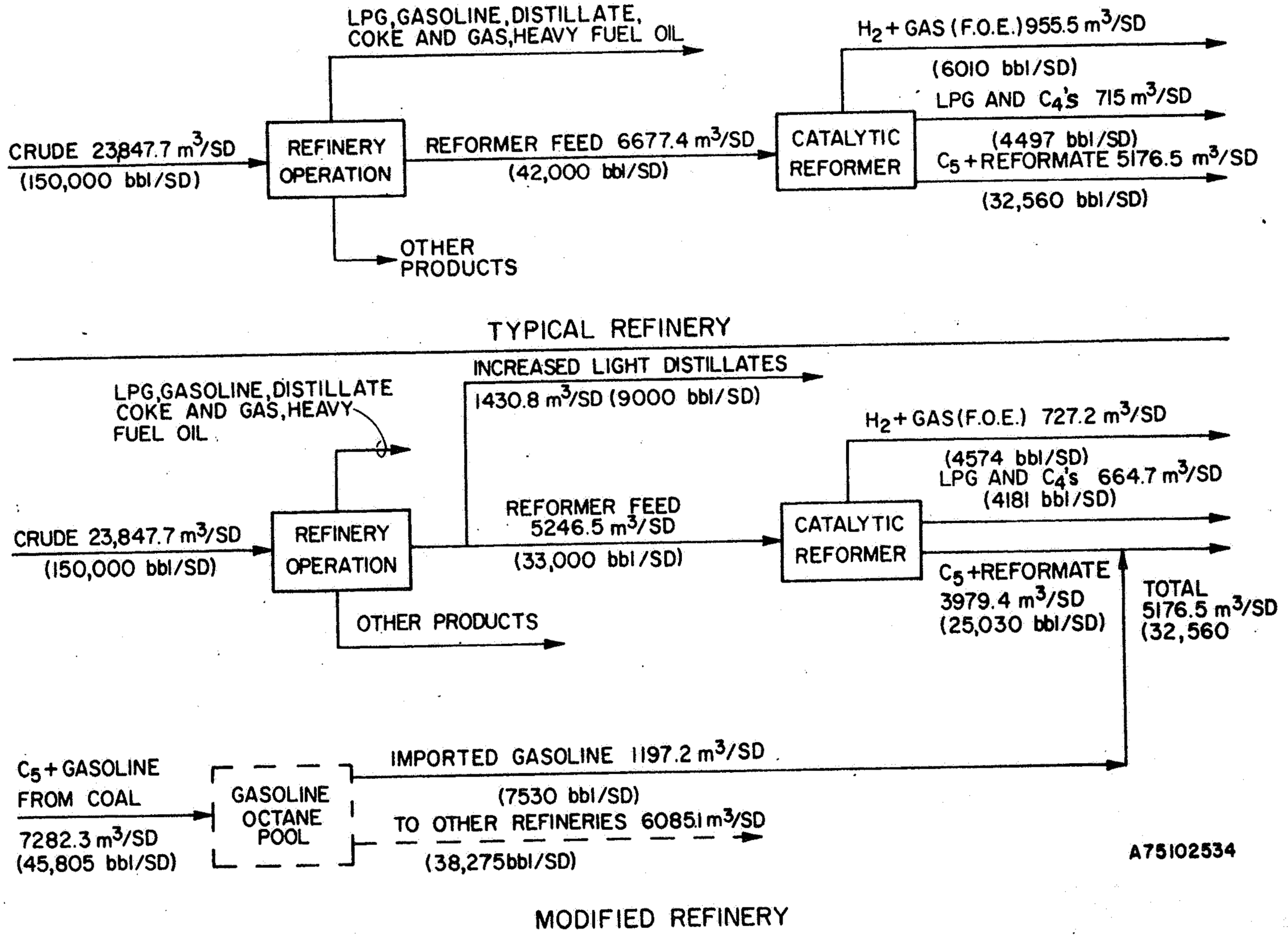

Fig. 9. MODIFICATION OF CONVENTIONAL REFINERY-IMPORTED GASOLINE ALLOWS REDUCTION OF CATALYTIC REFORMING AND INCREASED LIGHT DISTILLATE

(150,000 bbl/SD) refineries and release a total of 8702.9 m^3/SD (54,740 bbl/SD) of light distillate for jet fuel. The overall effect is shown in Figure 10.

The conventional refinery benefits from this modification from an energy standpoint, and there are also savings in capital and operating costs because of reduced reforming.

Table 48 summarizes the refinery products for typical refinery operations and for those importing aromatic gasoline from coal. The product quantities for the refinery given here are those derived by Mr. Oberle from generalized refinery concepts. The scope of this project did not permit a more detailed examination of refinery operations. However, the numbers were derived using conservative bases and show what appears to be a very reasonable way to use the highly aromatic coal products where they can best contribute to the supply of refinery products.

The fact that there are six refineries that import gasoline from one coal plant should not be a serious hindrance because the gasoline can be pipelined to refineries or to a blending location, depending on the logistics of supply. Using as a basis an average pipeline transmission distance of 1852 km (1000 mi) and an energy consumption of 4.1 kJ/m^3-km (50 Btu/bbl-mi), the total energy consumed is only about 0.6% of the total coal feed to the CSF plant.

As is indicated in Table 48, the modified refinery saves fuel costs and produces more light distillate than the conventional refinery while meeting 9% of its gasoline needs by imports. For the modified refinery we have assumed payment for the importation of coal-based aromatic gasoline at the same price as the product — 9.1¢/liter (34.5¢/gal). As a result, to keep the refinery revenues constant, a subsidy of 0.55¢/liter (2.1¢/gal) of imported gasoline must be applied and charged to the increased production of light distillate via the coal-based route to calculate the costs for this processing scheme.

Overall Efficiency

The refinery operations obviously make an essential contribution in increasing the supply of jet fuel component (light distillate) by the substitution of 9% of the refinery-derived reformate gasoline with coal-derived aromatic gasoline.

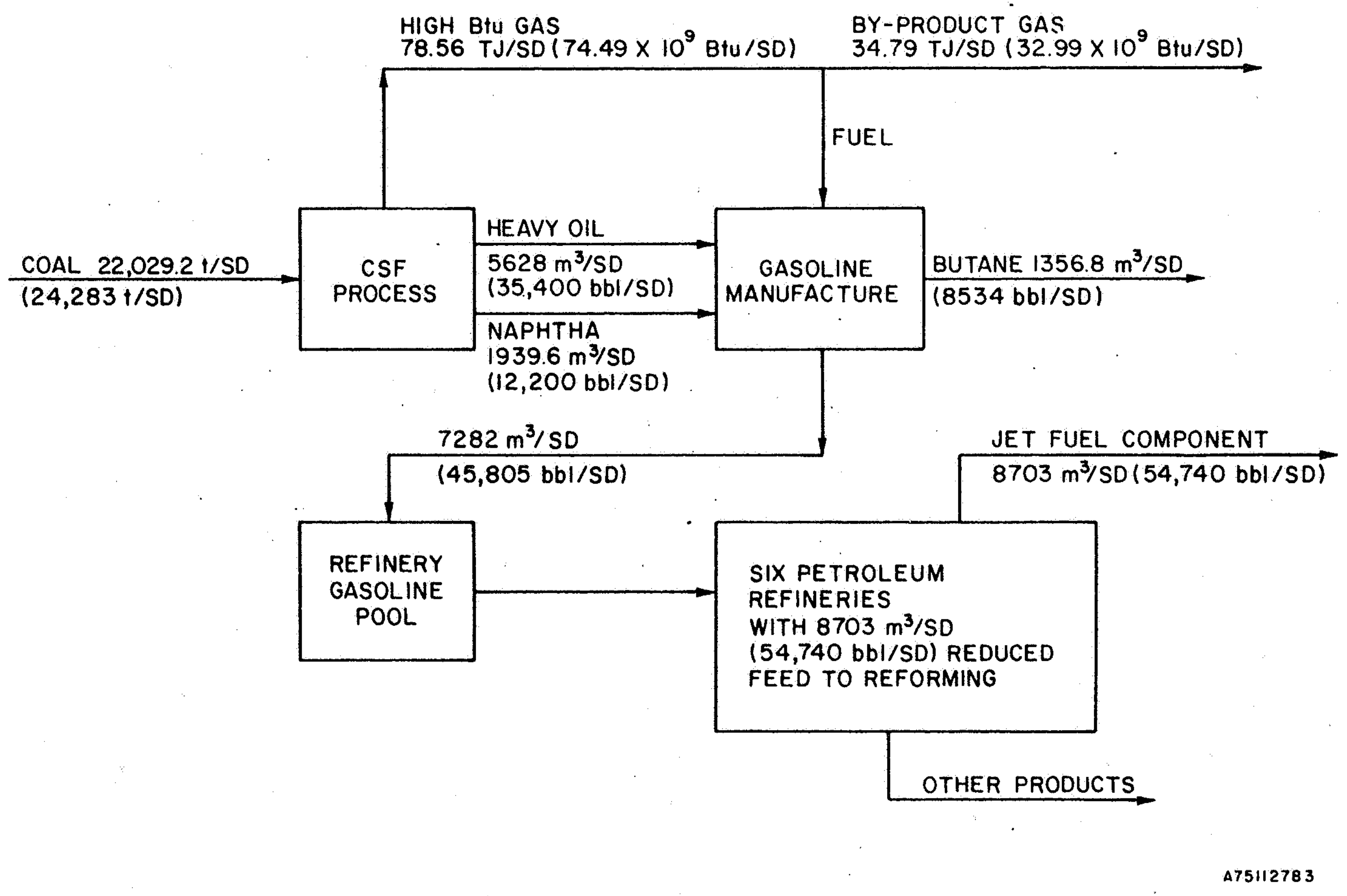

Fig. 10. OVERALL EFFECT OF IMPORTING COAL-DERIVED GASOLINE INTO REFINERIES TO REDUCE THE AMOUNT OF LIGHT DISTILLATE FEED TO CATALYTIC REFORMING TO INCREASE JET FUEL COMPONENT

Table 48. REFINERY OUTPUTS WITH IMPORTATION OF COAL-DERIVED AROMATIC GASOLINE TO PERMIT HIGHER OUTPUT OF DISTILLATE FOR JET FUEL COMPARED WITH TYPICAL U.S. REFINERY OPERATIONS FOR SAME GASOLINE OUTPUT

Aromatic Gasoline Import →
7282.3 m³/d
(45,805 bbl/d)

Crude Oil Import →
145,073.7 m³/d
(912,500 bbl/d)
(6 typical refineries)

Energy Consumed →
13,768.1 m³/d
(86,600 bbl/d)
(F.O.E. 4.89 GJ/m³
= 6.3 X 10⁶ Btu/bbl)

Several Refineries With Imports of Aromatic Gasoline From Coal and Reduced Reformer Runs to Raise Amount of Light Distillate

	Products m³/d	Products bbl/d	$/bbl	$/d
→ LPG	4,239.4	26,665	7.65	203,987
→ Gasoline	80,511.0	506,400	14.49	7,337,736
→ Distillates*	57,306.8	360,450	12.81	4,617,365
→ Coke and Gas, Fuel Oil Equivalent	7,178.9	45,154	12.60	568,940
→ Heavy Fuel Oil	2,909.5	18,300	12.60	230,580
		Total Products		12,958,608
Less Fuel	13,768.5	86,600	12.60	−1,091,160
Less Imported Gasoline	7,282.3	45,805	14.49†	−663,714
				11,203,734

Crude Oil Import →
145,073.7 m³/d
(912,500 bbl/d)

Energy Consumed →
14,362.7 m³/d
(90,340 bbl/d)

Several Typical U.S. Refineries Having Same Gasoline Output as Above

	Products m³/d	Products bbl/d	$/bbl	$/d
→ LPG	4,356.2	27,400	7.65	209,610
→ Gasoline	80,511.0	506,400	14.49	7,337,736
→ Distillates	48,602.3	305,700	12.81	3,916,017
→ Coke and Gas, Fuel Oil Equivalent	8,696.6	54,700	12.60	689,220
→ Heavy Fuel Oil	2,909.5	18,300	13.44	230,580
		Total Products		12,383,163
Less Fuel	--	90,340	12.60	−1,138,284
				11,244,879

* Importation of 7282 m³/d (45,805 bbl/d) coal-derived gasoline allows increased manufacture of 8704.5 m³/d (54,750 bbl/d) of light distillate for use in jet fuel.

† For refinery to break even with typical operation, imported gasoline must be available at $13.59/bbl or 2.1¢/gal less than product.

Note: No charge is shown for crude oil since it is the same for both cases and our purpose is to show differences.

Since our study is based on the conversion of coal to various fuels, the overall efficiency of the increased jet fuel production should be referred back to the coal conversion operation in order to be on the same basis as the other processes. Any net debits or credits in the overall refinery operations should be included in the costs.

Table 49 presents overall efficiencies for the three steps in increasing the supply of jet fuel by this method. All product outputs are expressed as percentages of the coal feed to the CSF plant. The overall efficiency is 70% for this process, including 1.73% conversion to ammonia and sulfur, which is carried as a constant for each of the three tabulations. In making the aromatic gasoline, 56 % of the high-Btu gas is used and 1356.8 m^3 (8534 bbl) of butane are made as by-products. As a result, the overall efficiency drops to 60.85%. However, when the aromatic gasoline is imported into the refinery system, resulting in an increased output of jet fuel, the amount of this product is greater than the imported gasoline because of the larger volume of jet fuels even though it sells for less; when referred to the coal, this represents an apparent increase in overall efficiency. The increased volume of refinery light distillate is accompanied by reductions in refining LPG, coke, and gas production. The modified refinery operation saves a little fuel when there is less reforming, so there is no efficiency loss. The differences in products resulting from the change in refinery operations are shown in Tables 48 and 49 and accounted for in the economics, so the unit cost of the increased light distillate can be referred to the full 8703 m^3 (54, 740 bbl).

Table 49. COMPARISON OF OVERALL EFFICIENCIES FOR CONVERSION OF COAL TO LIQUID PRODUCTS

Total Coal Into System —— 18,938.9 Dry Metric Tons/Day (20,876.6 Short Tons/Day ——
—— 554.21 TJ/Day (525.48 X 10^9 Btu/Day) ——

	Product Output Relative to Coal								
	Consol Synthetic Fuel Process (CSF)			CSF Plus Conversion to Aromatic Gasoline			Increase in Light Distillate From Petroleum Refinery Due to Imported Coal-Derived Gasoline		
	TJ/d	10^9Btu/d	%*	TJ/d	10^9Btu/d	%*	TJ/d	10^9Btu/d	%*
High-Btu Gas	78.56	74.49	14.18	34.79	32.99	6.28	34.79	32.99	6.28
Naphtha	65.86	62.45	11.88	--	--	--	--	--	--
Fuel Oil	233.97	221.85	42.22	--	--	--	--	--	--
Ammonia	2.62	2.48	0.47	2.62	2.48	0.47	2.62	2.48	0.47
Sulfur	7.00	6.64	1.26	7.00	6.64	1.26	7.00	6.64	1.26
Butane	--	--	--	39.23	37.20	7.08	39.23	37.20	7.08
C_5+ Gasoline	--	--	--	253.60	240.45	45.76	--	--	--
Increase in Light Distillate by Reduction in Reforming	--	--	--	--	--	--	310.37	294.28	56.00
Total	388.01	367.91	70.01	337.24	319.76	60.85	394.01	373.59	71.09

* Percent of input coal to CSF plant.

THE MANUFACTURE OF JET FUEL FROM THE PRODUCTS OF COAL LIQUEFACTION

The direct conversion of coal to jet fuel requires as the final step the upgrading of the products of the CSF Process to meet the jet fuel specifications. In order to meet specifications, the aromatic content must be reduced to 20% or less, and the boiling range and freezing point must be lowered (Table 40).

We were fortunate in obtaining the process requirements for upgrading the heavy fuel oil to jet fuel from the Process Division of Universal Oil Products Co. (UOP).

The process design[3] for the CSF Process did not include detailed analyses of the liquid products. Therefore, in order to give UOP sufficient information for their work, we supplied them with analytical data for synthetic crude fractions determined by Consolidation Coal Co. (Consol) in 1965 from a sample of synthetic crude obtained from laboratory studies of the Consol Synthetic Fuels Process. During the bench-scale research effort, Consol had worked with UOP in evaluating their synthetic crude in a pilot plant study of the UOP's Isomax Process. Therefore, it is logical to consider UOP a good source of information on this subject.

A two-step operation is specified by UOP for the manufacture of jet fuel from the heavy fuel oils. The first step is hydrocracking (HDC Unibon Process) followed by an aromatics hydrogenation step (AH Unibon). Figures 11 and 12 show typical flow diagrams for these processes. Overall process requirements and investment costs were estimated by UOP from process correlations based on a composite feed analysis derived from the analytical data mentioned above. Since a detailed process design was not prepared, these two flow diagrams represent schematically the flow sequences involved in manufacturing jet fuel.

Figure 13 is an overall block flow diagram, starting with the CSF coal conversion plant and its three products (heavy oil, naphtha, and high-Btu gas) and continuing with their subsequent utilization in making jet fuel. Figure 7, presented earlier in this report, shows the CSF Process. The processing scheme developed here does not utilize naphtha to make

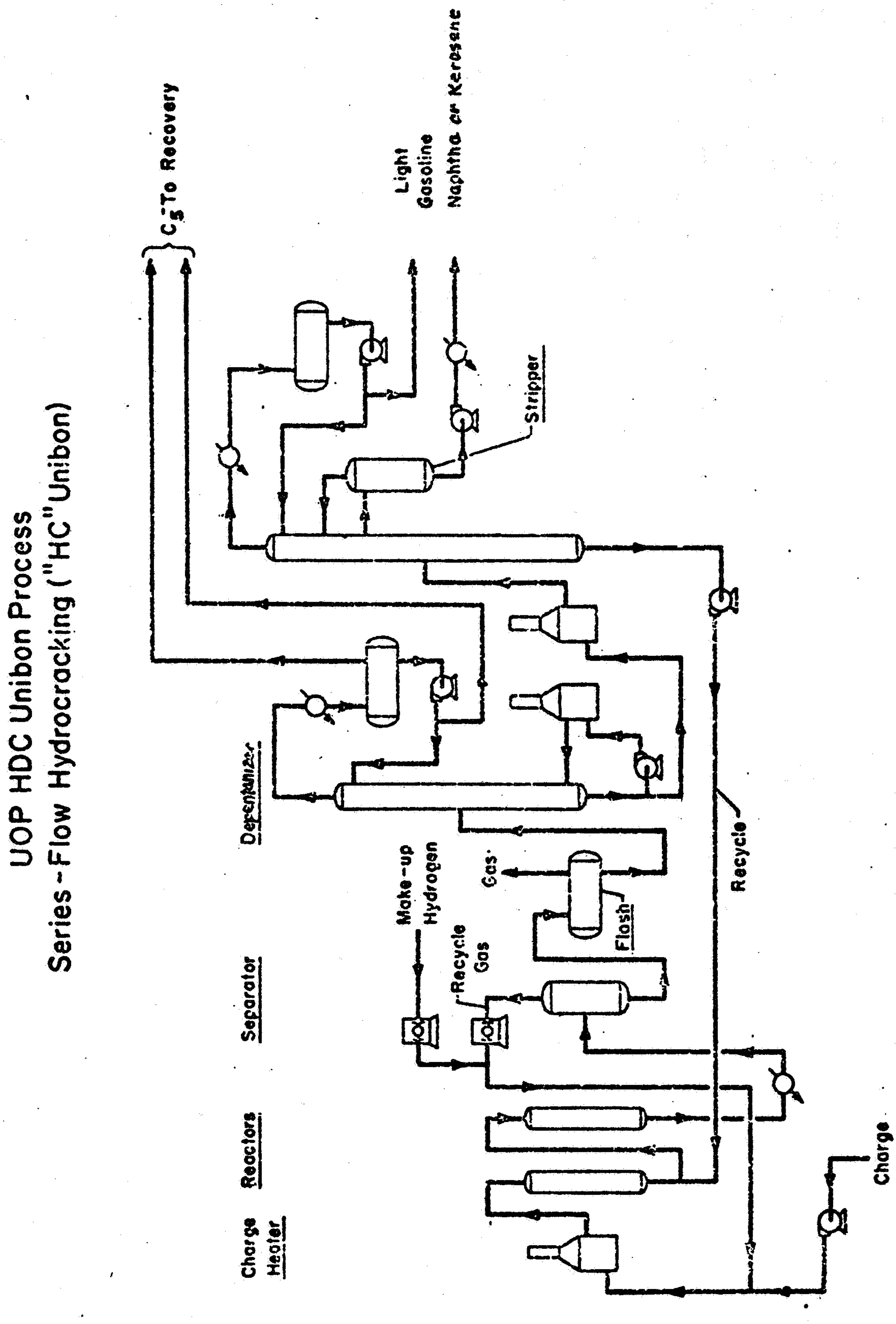

Figure 11. MODIFIED FLOW SCHEME FOR IGT STUDY

Note: The basic flow scheme for this project would be similar to the one above.

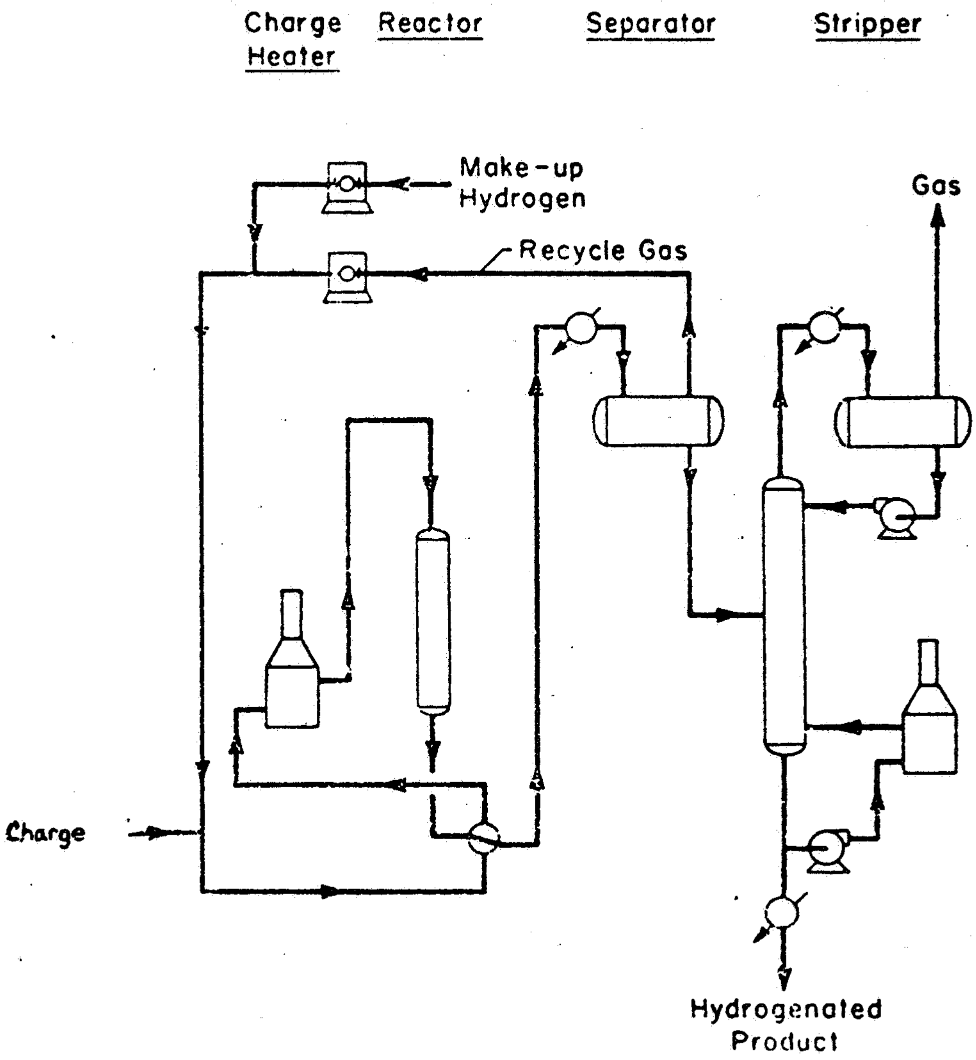

Figure 12. UOP AH UNIBON PROCESS

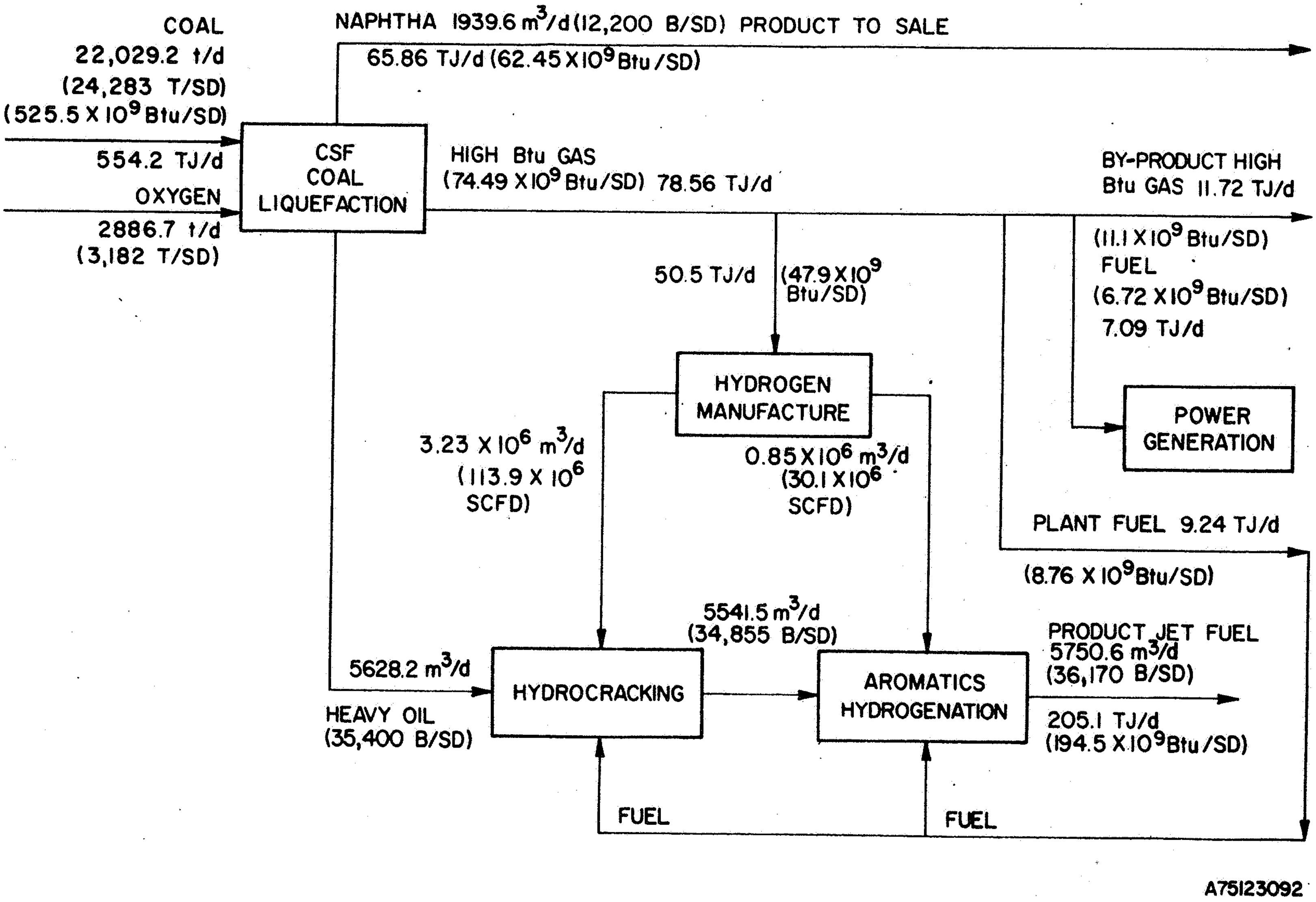

Figure 13. MANUFACTURE OF JET FUEL FROM COAL-DERIVED HEAVY FUEL OIL

jet fuel. Some of the naphtha could be used to increase the volume of jet fuel produced, but it would not raise the overall efficiency, as we have taken full credit for this stream as a by-product from the CSF Process. Table 50 summarizes the process flow quantities and analytical data from the information supplied by UOP.

Hydrocracking (HDC)

HDC Unibon is a highly versatile process for the conversion of a variety of petroleum fractions to yield more valuable, lower boiling products. Concomitant with the hydrocracking reaction is the almost complete conversion of sulfur-, nitrogen-, and oxygen-containing compounds. Thus, the process yields products that are essentially mixtures of only paraffins, naphthenes, and aromatics. Some applications are:

1. Conversion of the petroleum fractions ranging from kerosene to solvent-extracted oil to yield principally naphtha, either for motor fuel or for aromatics production; LPG is a by-product of such an operation.
2. Conversion of vacuum gas oil and/or solvent-extracted oil to maximize the yield of diesel/light oil and/or kerosene.
3. Production of conventional or high viscosity index (VI) lube oil base stocks from vacuum gas oil and/or solvent-extracted oil.

The process is carried out at elevated temperature and pressure over a fixed catalyst bed, where the fresh feed is cracked in a hydrogen atmosphere. The exact process conditions vary widely depending on the properties of the charge stock and the products desired.

Hydrogen is consumed in all hydrocracking reactions. The carbon-carbon bonds in the paraffins are broken, and the fragments are hydrogenated to form two or more lower molecular weight paraffins. Polycyclic aromatics are partially saturated and ring opening in one or more places follows, although at least one ring is usually retained.

A large variety of flow schemes have been employed in different installations; a two-stage, "series flow" system was chosen for this application. In this system there are two separate reaction zones, each employing a different type of catalyst, but essentially within a single reaction section. In other words, there is a single charge-pump, heater,

Table 50. CONVERSION OF HEAVY FUEL OIL FROM THE CSF PROCESS TO JET FUEL

Processing Step	HDC Unibon					AH Unibon	
	Feed		Product			Feed	Product
°API		15.7				41.2	45.1
IBP	208°C	(406° F)				75° C (168° F)	74° C (155° F)
10%	214	(418)				---	99 (210)
30%	223	(441)				--	127 (260)
50%	248	(479)				147 (296)	146 (295)
70%	267	(512)				--	171 (340)
90%	301	(573)				--	207 (405)
EP	359	(678)				232 (450)	232 (450)
M³/sd (bbl/sd)	5,628.1	(35,400)	7,057.1	(44,388)		5,541.5 (34,855)	5,751.0 (36,173)
98% H_2, 10^6 SCF/d)	3.22,	(113.9)*				0.852 (30.1)*	
H_2 Consumption, kg/hr)	10,604	(23,378)				2,835 (6,250)	
Component			kg/hr	lb/hr	wt %		
NH_3			518	1,142	0.22		
H_2S			991	2,184	0.42		
C_1			225	496	0.10		
C_2			428	943	0.18		
C_3			6,799	14,990	2.88		
C_4			18,506	40,799	7.85		
C_5			19,204	42,338	8.15		
C_6			20,262	44,671	8.60		
C_7 +			168,808	372,158	71.60		
Total	102,120	(496,343)	235,741	519,721	100.00	189,008 (416,692)	191,843 (422,942)
Vol. % Aromatics		83.5		38.7		38.7	17
Freeze Point, °C (°F)		--		--		-57 (-70)	-57 (-70)
Smoke Point, m.m.		--		--		~17	Smoke Volatility Index = 57
Sulfur, wt %		0.17					0.0004
Nitrogen, wt %		0.19					

* Includes Losses

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recycle compressor, etc. This system will normally utilize at least two reactors arranged for series flow, with the total effluent from the first reactor flowing directly to the second. The first state reactor decomposes nitrogen and sulfur compounds, saturates olefins, and partially saturates polycyclic aromatics. The second reactor carries out the hydrocracking in the presence of H_2S and NH_3.

In the hydrocracking operation, 5628 m^3/d (35,400 bbl/d) react with 3.23 X 10^6 m^3/d (113.9 X 10^6 SCFD) of process hydrogen (98% purity). Because of the reduction in density that accompanies the formation of lighter compounds, the volume of liquid products (C_4 and heavier) is increased to 7057.1 m^3/d (44,388 bbl/d). Out of a total effluent of 235,741 kg/h (519,721 lb/h), 80.2 weight percent is C_6 and heavier material. Another 16% is C_4 and C_5 material, and the remaining 3.8% is lighter hydrocarbons. Gases are flashed off for fuel and the liquid product is depentanized and fractionated to yield 5541.5 m^3/d (34,855 bbl/d) of C_6 and heavier material that is then fed to the second step in the overall conversion process, the aromatics hydrogenation step. The change in properties due to the hydrocracking step can be seen in Table 50. The aromatics content is reduced from 83.5% to 38.7%, the boiling range is reduced from 480 to 632 K (406° to 678°F) to 347 to 505 K (165° to 450°F), and the API gravity is raised from 15.7° to 41.2°.

Aromatics Hydrogenation (AH)

As it is typically applied to the production of jet fuel, the AH Process follows a Unibon operation.

The HDC Unibon product fractionator overhead is mixed with recycle and makeup hydrogen, heated, and charged to the aromatics saturation reactor, which contains a highly active hydrogenation catalyst that selectively saturates the aromatics contained in the feed. The reaction is carried out at an intermediate pressure and space velocity, with the specific conditions determined by the feed characteristics and the extent of aromatic saturation necessary to meet product smoke point requirements. The reactor effluent is cooled and then flows to a separator for the recovery of recycle hydrogen. Separator liquid is stripped for the removal of dissolved hydrogen and light ends that may have entered the unit with the makeup gas.

Operating conditions may be selected to yield a treated product almost entirely free of aromatics. Minor amounts of light ends are produced, essentially all of which are a result of fracturing the contaminant-carrying organic feed compounds.

For aromatics saturation, 5541.5 m^3/d (34,855 bbl/d) of hydrocracked product react with 0.85 X 10^6 m^3/d (30.1 X 10^6 SCFD) of process hydrogen. The increased product weight and the reduced density, due to hydrogen absorption, result in a product volume of 5751 m^3/d (36,173 bbl/d).

Hydrogen Supply and Offsites

In addition to the HDC and AH units, the other major unit in the jet fuel add-on is the hydrogen plant. UOP supplied information on the chemical consumption of hydrogen plus an estimate of the losses. The hydrogen plant capacity is 4.08 X 10^6 m^3/d (144 X 10^6 SCFD) of 98% pure hydrogen at 1480 kN/m^3 (200 psig). The hydrogen input and utility requirements are summarized below:

	Process Hydrogen Input		Power	Cooling Water		Fuel Input	
Unit	10^6 m^3/d	10^6 SCFD	kW	m^3/h	gpm	GJ/h	10^6 Btu/h
Hydrocracking	3.23	113.9	22,500	636	2800	284.8	270
Aromatics Hydrogenation	0.85	30.1	5,410	227	1000	100.2	95
Total	4.08	144.0	27,910	863	3800	385.0	365

The high-Btu by-product gas from the CSF Process is sufficient to meet these fuel requirements and to supply the reformer feed and fuel for the hydrogen plant. This gas has a higher heating value of 34.7 MJ/m^3 (932 Btu/SCF) and is composed of 7.1% CO, 30.9% H_2, 38.7% CH_4, 11.7% C_2H_6, 8.1% C_3H_8, and 3.4% N_2 (Table 42). The high-Btu gas is converted to hydrogen by the steam hydrocarbon process. Many commercial plants currently in operation are based on natural gas feedstock. The basis for the steam-hydrocarbon hydrogen manufacturing process is the fact that the above hydrocarbons (alkanes) react catalytically with steam as follows:

$$C_nH_{2n+2} + n\ H_2O \rightarrow nCO + (2n+1)\ H_2 \qquad (1)$$

Some CO_2 is produced by the conversion of CO:

$$CO + H_2O \rightarrow CO_2 + H_2 \qquad (2)$$

Figure 14 shows a typical hydrogen plant based on natural gas reforming and indicates the major steps in the process. Because this type of plant is so well established, we have treated this section of the coal conversion plant as a package item, as raw material requirements and capital costs are available on an overall basis.

The reformer effluent passes through the steps of waste heat recovery, CO conversion by Reaction 2, CO_2 removal, and "cleanup" methanation to remove residual carbon oxides. Because there is 38% H_2 + CO in the reformer feed gas, the estimates of fuel requirements based on natural gas feed to the hydrogen plants are therefore conservative. On this basis, there is more than enough gas to supply the 385 GJ/h (365 X 10^6 Btu/h) fuel input to the HDC and AH units, to supply the hydrogen plant, and to provide fuel for the power requirements. The distribution of this gas is as follows:

	GJ/h	10^6 Btu/h	TJ/d	10^6 Btu/d	lb-mol/d	kg-mol/d
Hydrogen Plant	2105	1996	50.51	47.90	804.5	364.9
Process Fuel	385	365	9.24	8.76	1048.8	475.7
Power Generation	295	280	7.09	6.72	5735.3	2601.5
Excess (by-product)	488	463	11.72	11.11	1330.4	603.5
Total	3273	3104	78.56	74.49	8919.0	4045.6

Table 51 gives the product summary for the conversion of coal to jet fuel (the major product) and the other by-products. The major by-product is the naphtha from the CSF Process. All by-products have been priced conservatively to represent the value at the conversion plant, which is located near the coal mine.

The overall conversion efficiency for coal is summarized in Table 52. The naphtha heating value is 19,253 GJ/m³ (121,900 Btu/gal) and the jet fuel heating value 20,217 GJ/m³ (128,000 Btu/gal).

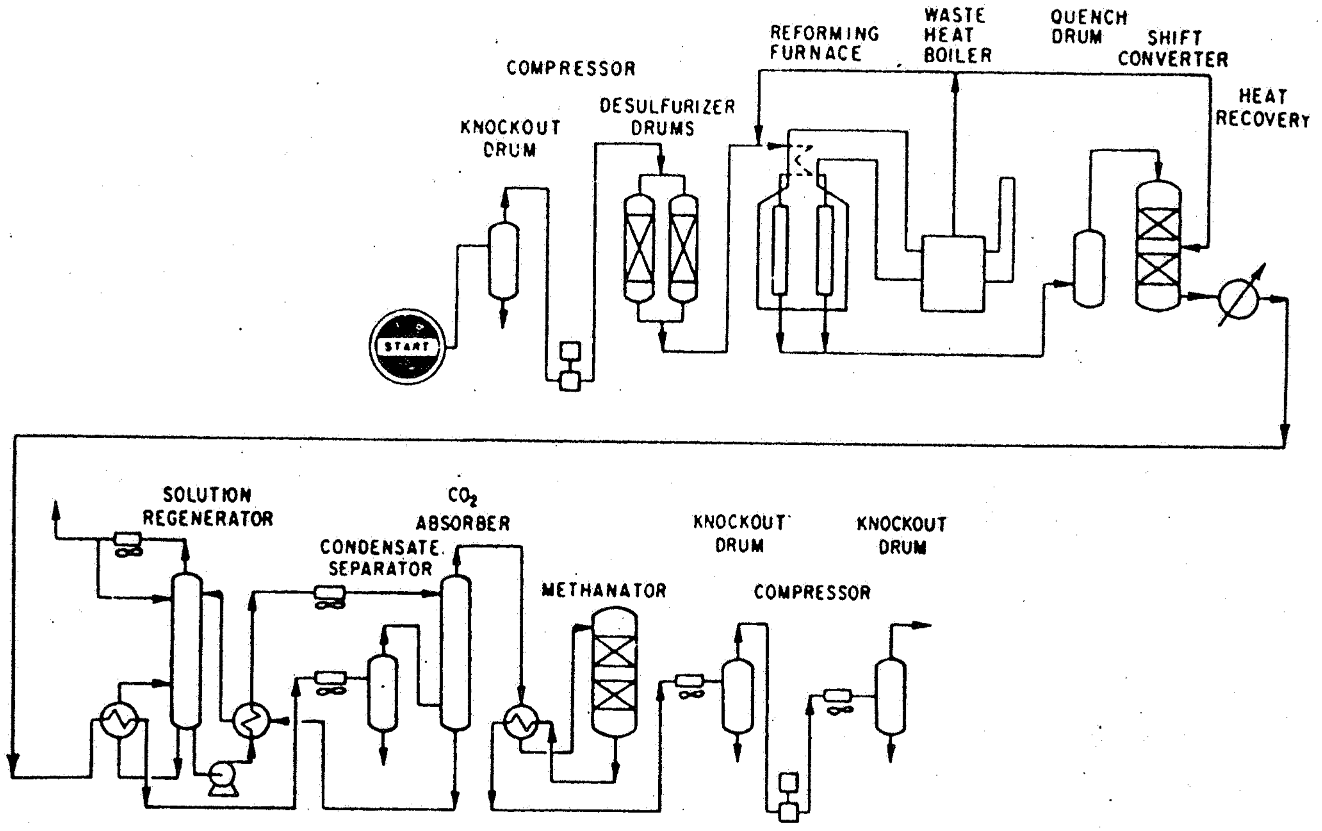

Figure 14. HYDROGEN MANUFACTURE BY THE STEAM REFORMING OF NATURAL GAS

Adapted From Lee, G.T., Leslie, J.D. and Rodekohr, H.M., "The Cost of Hydrogen Made From Natural Gas," Hydrocarbon Process 42, 126 (1963) September.

Table 51. PRODUCTS SUMMARY FOR CONVERSION OF COAL BY THE CONSOL SYNTHETIC FUEL PROCESS PLUS UPGRADING OF HEAVY FUEL OIL TO JET FUEL

Products	Amount/Stream-Day	Amount/Yr	Assumed Unit Value at Coal Conversion Plant	Annual Value, $
C_5–204°C. (400°F) Naphtha	1939.5 m^3 (12,200 bbls)	637,173.4 m^3 (4,007,700 bbls)	5.3¢/liter (20¢/gal)	33,664,680
High-Btu Gas	11.72 TJ (11.11 X 10^9 Btu)	3849.2 TJ (3,649.6 X 10^9 Btu)	1.61/GJ ($1.70/10^6 Btu)	6,204,320
Sulfur	754.9 metric tons (743 long tons)	247,992 metric tons (244,075 long tons)	9.84/metric ton ($10/long ton)	2,440,750
Ammonia	116.5 metric tons (128.4 tons)	38,265 metric tons (42,179 tons)	$38.6/metric ton ($35/ton)	11,476,270
Total By-Product Value				43,786,020
Jet Fuel Product	5751 m^3 (36,173 bbls)	1,889,215 m^3 (11,882,830 bbls)		

Table 52. OVERALL EFFICIENCY FOR CONVERSION OF COAL TO JET FUEL

Total Coal Into System — 554.21 TJ/d (525.48 X 10^9 Btu/d)

	Product Output Relative to Coal					
	CSF Process			Jet Fuel Manufacture		
	TJ/d	10^9 Btu/d	%*	TJ/d	10^9 Btu/d	%*
High-Btu Gas	78.56	74.49	14.18	11.72	11.11	2.11
Naphtha	65.86	62.45	11.88	65.86	62.45	11.88
Fuel Oil	233.97	221.85	42.22	--	--	--
Ammonia	2.62	2.48	0.47	2.62	2.48	0.47
Sulfur	7.00	6.64	1.26	7.00	6.64	1.26
Jet Fuel	--	--	--	205.10	194.47	37.01
Total	388.01	367.91	70.01	292.30	277.15	52.73

*Percent of Input Coal

Total liquefaction by the CSF Process shows an overall efficiency of 70.0%. In converting the heavy fuel oil to jet fuel, 13.3% of the former is lost. In making hydrogen and supplying fuel, 75% of the high-Btu gas is consumed. The overall efficiency is 52.7%. This efficiency is lower than for conversion of coal to aromatic gasoline, and much less than the alternative method for increasing the supply of jet fuel.

PROCESS EFFICIENCY ANALYSIS

Tables 53 and 54 present comparative summaries of the major material requirements, energy balances, overall power yields, and overall efficiencies for the processes analyzed in this study. Since coal is the base energy source, all product percentages are based on the total coal input. It is important to remember this, particularly in the case of liquids from coal, as the products from the CSF Process are further upgraded into gasoline or jet fuel.

Although some of the processes produce significant amounts of by-products, which are included in the overall efficiency, none are considered to be multi-product plants.

It should also be remembered that overall efficiency is not the only criterion; both the overall efficiency and the process economics should be considered in assessing a process.

Hydrogen Processes

The U-GAS Process shows the highest overall efficiency (66.4%), compared to the Steam-Iron Process (62.6%) and the Koppers-Totzek Process (57%). However, with steam-iron, 18.0% of the coal feed is converted to electric power, a high-grade product that normally is generated at a low fuel conversion efficiency. It is essential that some use be made of the large amount of heat energy remaining in the spent producer gas in order to raise the overall efficiency from a low value of 44.6% for hydrogen alone to the final value. Generation of electricity is one way of upgrading this heat to a universally valuable product. If the heat energy input to the power plant could be sold as a by-product, then the overall efficiency would be 81.5%.

Hydrogen by Koppers-Totzek gasification, although a commercially operating process, requires about 40% more oxygen than the U-GAS Process, primarily because of the much higher gasifier temperature. The lower oxygen requirement and higher pressure of the U-GAS gasifier save much energy and give a significantly higher efficiency (66.4%) than the 57% overall efficiency for the Koppers-Totzek Process.

Table 53. SUMMARY OF PROCESS MATERIAL REQUIREMENTS, PRODUCT YIELDS, AND OVERALL EFFICIENCIES

Main Product	Hydrogen						Methane				Jet Fuel Component Via CSF Coal Liquefaction	
Gasifier	Koppers-Totzek		U-Gas		Steam Iron		HYGAS*		CO_2 Acceptor		Increased Refinery Output by Importing Gasoline from Coal	Jet Fuel by Hydrocracking Synfuel from coal
	kg/h	lb/h	kg/h	lb/h	kg/h	lb/h	kg/h	lb/h	kg/h	lb/h	kg/h	lb/h
Reaction Coal (dry)	557,733	1,229,590	494,144	1,089,400	909,035	2,004,080	471,613	1,039,728	630,072	1,389,070	755,980	1,666,670
Fuel Coal (dry)	179,721	396,218	138,055	304,360	30,096	66,350	109,675	232,974	25,619	56,483	29,730	65,550‡
Total Coal (dry)	737,454	1,625,808	632,199	1,393,131	939,131	2,070,430	577,288	1,272,702	655,291	1,445,553	783,710	1,732,220
											H₂ Only	
Gasifier Steam	122,544	270,164	170,449	375,776	2,086,166	4,599,209	455,013	1,003,131	499,460	1,101,120	161,025	355,000
CO Conversion Steam	734,900	1,620,178	623,787	1,375,215	No CO Conversion		459,875*	1,013,650*	No CO Conversion		--	--

	Koppers-Totzek		U-Gas		Steam Iron		HYGAS		CO_2 Acceptor		Jet Fuel Component Via CSF Coal Liquefaction	
	Metric tons/d	Short tons/d	Metric tons/d	Short tons/d	Metric tons/d	Short tons/d	Metric tons/d	Short tons/d	Metric tons/d	Short tons/d	Metric tons/d	Short tons/d
Total Coal (as received)†	22,690	25,012	19,453	21,443	28,897	31,853	17,763	19,580	20,175	22,239	22,029	24,283
Oxygen	10,969	12,092	7,830	8,631	15,756	17,368	2,720	2,999	8,769	9,666	2,887	3,182
	(98% Purity)		(98% Purity)		(as Air)		(98% Purity)		(as Air)		99.5%	

	Koppers-Totzek			U-Gas			Steam Iron			HYGAS			CO_2 Acceptor			Increased Refinery Output by Importing Gasoline from Coal			Jet Fuel by Hydrocracking Synfuel from coal		
	TJ/d	10^9 Btu/d	%	TJ/d	10^9 Btu/d	%	TJ/d	10^9 Btu/d	%	TJ/d	10^9 Btu/d	%	TJ/d	10^9 Btu/d	%	TJ/d	10^9 Btu/d	%	TJ/d	10^9 Btu/d	%
H. H. V. of Coal	464.6	440.5		398.3	377.7		591.7	561.0		363.7	344.8		413.1	391.7		554.2	525.5		554.2	525.5	
H. H. V. Main Product	264.0	250.3	56.8	263.9	250.2	66.2	263.9	250.2	44.6	254.7	241.5	70.0	264.4	250.7	64.0	310.4	294.3	45.8††	205.1	194.5	37.0
H. H. V. Naphtha	--	--	--	--	--	--	--	--	--	--	--	--	--	--	--	--	--	--	65.9	62.5	11.9
H. H. V. C_4	--	--	--	--	--	--	--	--	--	--	--	--	--	--	--	39.2	37.2	7.1	--	--	--
H. H. V. Power	--	--	--	--	--	--	106.1	100.7	18.0	--	--	--	10.0	9.5	2.4	--	--	--	--	--	--
H. H. V. Other Products	** 0.8	0.8	0.2	0.7	0.7	0.2	--	--	--	14.3	13.6	4.0	2.9	2.7	0.7	44.4	42.1	8.0	21.3	20.2	3.8
H. H. V. of Other Coal to Products and Plant Efficiency	264.8	251.1	57.0	264.6	250.9	66.4	370.0	350.9	62.6	269.0	255.1	74.0	277.3	262.9	67.1	394.0	373.6	60.9	292.3	277.2	52.7

* Obtained from vaporized slurry water used in feeding coal to the gasifier.

† For hydrogen and methane, Montana subbituminous coal at 22% moisture; for jet fuel, Pittsburgh seam bituminous coal at 14.4% moisture.

‡ Fuel indicated is for CSF Process design only. Fuel for gasoline and jet fuel manufacture is obtained from high-Btu gas made in CSF Process.

** Ammonia and sulfur; high-Btu gas from CSF Process.

†† C_4+ gasoline. Increased refinery light distillate after importation of this gasoline is equivalent to 56% of coal HHV.

Table 54. OVERALL ENERGY BALANCES AND PROCESS EFFICIENCIES FOR VARIOUS PROCESS DESIGNS FOR COAL CONVERSION

	H_2 Product						Methane (Pipeline Gas)				Coal Liquefaction	
Gasifier Type	Koppers-Totzek		U-GAS™		Steam-Iron		HYGAS®		CO_2 Acceptor		By CSF Process†	
	TJ/d	10^9 Btu/d	TJ/d	10^9 Btu/d	TJ/d	10^9 Btu/d	TJ/d	10^9 Btu/d	TJ/d	10^9 Btu/d	TJ/d	10^9 Btu/d
High Heating Value, Total Coal Input	464.6	440.5	398.3	377.7	591.7	551.0	363.7	344.9	413.1	391.7	554.2	525.5
Product Gas and Liquids	264.0	250.3	263.9	250.2	263.8	250.2	254.7	241.5	264.4	250.7	378.4	358.8
Output, % of Coal High Heating Value												
Product Gas	55.8		66.2		44.6		70.0		64.0		14.18	
Product Liquids	--		--		--		--		--		54.10 ‡	
By-Product Power	--		--		18.0		--		2.4		---	
By-Product Chemicals	0.2		0.2		--		4.0		0.7		1.73	
Process Vent Gases	3.2		2.6		2.3		3.0		3.7			
Stack Gases	3.2		2.9		13.9		2.1		2.3		29.99	
Heat Dissipated to Cooling Water and Air	33.6		25.1		14.5		17.2		20.9			
Gasifier Residue	1.4		1.4		5.2		1.5		0.8			
Assumed Losses	1.1		1.3		1.3		1.0		3.0			
Unaccounted Losses	0.5		0.3		0.2		1.2		2.2			
Total	100.0		100.0		100.0		100.0		100.0		100.0	

* Contains 94.7% CH_4; 4.4% H_2.

† This process design is the basis for manufacture of aromatic gasoline from coal for refinery import or direct conversion of heavy oil to jet fuel.

‡ 42.22% heavy fuel oil plus 11.88% naphtha.

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Methane Processes

By the use of a water slurry feed for Montana coal, we have achieved the highest design efficiency for the HYGAS Process reported to date, 74%. Of this figure, 4% is due to by-products, primarily B-T-X generated in the HYGAS reactor. These compounds are relatively easy to extract from the raw gas. The water used to slurry the coal is vaporized in the slurry dryer to provide steam for the CO shift reaction, thus eliminating the losses that would result from condensation and subsequent revaporization in a boiler. Although methanation, an exothermic reaction, is required as a final step in the process, it does not detract much from the overall efficiency because most of the heat is recovered for use in the process. The reduced oxygen consumption, process steam, and process horsepower requirements are the major reasons for the higher efficiency for the conversion of coal to methane relative to hydrogen.

The alternative process, the CO_2 Acceptor Process, converts 64% of feed coal to product, 6% less than the HYGAS Process. The major reason for this is probably the larger amount of power needed to run the CO_2 Acceptor Process compared to HYGAS, 33,500 versus 102,000 kW. The former figure includes 169,000 kW to drive the regenerator air compressor. While power for the latter is recovered from the regenerator off-gas, there are power losses due to inefficiency in the compressors and expanders. We have used a combined-cycle power recovery system, as in the Steam-Iron Process, which generates 115 MW of by-product electric power and 2.4% of the total coal input heating value.

Jet Fuel Component

The overall process of using coal to manufacture gasoline that is then imported into the refinery gasoline pool, which allows more refinery output of light distillate, is more efficient than the direct conversion of coal to jet fuel. The overall efficiency for the CSF Process is 70%. This drops to 61% because of the energy required to upgrade the CSF products to gasoline. The importation of 7282 m^3/SD (45,805 bbl/SD) of coal-derived gasoline results in an increase in the jet fuel component of 8703 m^3/d (54,740 bbl/d). When calculated as a percentage of the original coal, this results in an apparent increase in the overall efficiency (to 71%), because of the larger volume of jet fuel. However, the increased volume of this material is accompanied by reductions in refinery LPG, coke, and gas because of reduced catalytic reforming. The refinery suffers no loss in efficiency; in fact, there is a slight improvement. The changes in products and refinery economics have been accounted for in the economics of coal conversion, so that the unit cost of the increased light distillate can be referred to the full 8703 m^3 (54,740 bbl) jet fuel component.

Direct conversion of CSF heavy oil to jet fuel by hydrocracking and aromatics hydrogenation shows an overall efficiency of only 53% because of the larger hydrogen and fuel requirements. In both cases, the efficiency for the conversion to liquid products leaving the coal plant is less than for the two methane-producing processes.

Liquid Fuels

Although not a required part of this study, because of our basic interest in aircraft fuels, we have estimated overall efficiencies for the conversion of coal to liquid hydrogen and to methane for the three hydrogen processes and for the HYGAS Process. The results are given in Table 55. The requirements for hydrogen liquefaction are based on work done by the Linde Division, Union Carbide Co., for another NASA project.* The energy requirement for methane liquefaction is typical of LNG plants.

The additional energy load for liquefaction requires additional coal for hydrogen production by Koppers-Totzek and U-GAS Processes, while the Steam-Iron Process generates more than the needed electric energy as a by-product. On the basis of coal required for the generation of hydrogen and electricity for liquefaction, and including by-products, the Steam-Iron Process shows the highest overall efficiency.

The power requirement for methane liquefaction is just a fraction of that for hydrogen liquefaction. The overall efficiency from coal to liquid is thus much higher for methane than for hydrogen, 66.6% versus 34% to 48%, depending on the hydrogen process. While this efficiency is lower than the overall efficiency for the CSF Process, it is higher than the efficiency for upgrading the oils to gasoline or jet fuel.

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* Linde Division, Union Carbide Co., "Survey Study of the Efficiency and Economics of Hydrogen Liquefaction," n.d.

Tablé 55. LIQUEFACTION OF HYDROGEN AND METHANE FROM 250 BILLION Btu/DAY PRODUCT GAS DERIVED FROM COAL GASIFICATION

	Hydrogen			Methane
Process	Koppers-Totzek	U-GAS	Steam-Iron	HYGAS® [d]
Feed to Liquefaction Plant, mol/hr				
H_2	71,443	73,072	80,767	1,250
CH_4	4,252	3,722	1,245	26,776
CO	77	76	84	29
CO_2	4	8	84	29
N_2 + Ar	977	646	2,220	192
H_2O	10	8	11	4
Total	76,763	75,524	84,411	28,280
Product Liquefied, mol/hr [a]	62,656	64,084	70,833	24,362
Product Fuel Value, 10^6 Btu/hr	7,707.9	7,883.6	8,713.9	9,334.5
Power for Liquefaction, kW	682,096	697,642	771,115	102,800
Fuel for Liquefaction, 35% Efficiency, 10^6 Btu/hr	6,651.4	6,803.0	--	1,002.45
Excess Power, kW	--	--	458,385 [b]	--
Less Fuel Gas Recovered, 10^6 Btu/hr	−2,361.7	−2,068.3	1,054.7	1,002.45
Additional Coal Required, 10^6 Btu/hr	4,289.7	4,734.7	--	--
Gasification Plant Coal, 10^6 Btu/hr	18,355.4	15,735.6	23,374.6	14,876
Total Coal, 10^6 Btu/hr	22,645.1	20,470.3	23,374.6	14,876
Overall Efficiency, Coal to Liquid H_2, or CH_4, %	34.0 [e]	38.5 [f]	48.5 [c]	62.7 (plus 4.0% by-product)

[a] 87.7% of input hydrogen, 91% of input methane.

[b] Steam-Iron Process yields 1,229,500 kW of by-product power, more than enough for liquefaction.

[c] Includes liquid hydrogen, excess electric power, and recovered fuel gas – 11,331.1 X 10^6 Btu/h.

[d] Adjusted for 250 billion Btu/day.

[e] With 48% efficiency for conversion of fuel to electricity, overall efficiency is 37%.

[f.] With 48% fuel to electricity conversion efficiency, overall efficiency is 42.3%.

ECONOMICS OF COAL CONVERSION TO FLUID FUELS

We decided in the initial stages of this study to cost one process in each of the categories of hydrogen, methane, and kerosene production. We chose the U-GAS Process as the hydrogen process and the HYGAS Process as the methane process because both are new processes, both are processes in which oxygen gasification is employed, and both are IGT processes, which gives a good basis for comparability. As both are IGT processes, we felt that we had the best information available, and would not be biasing someone else's process. After the process designs for the three hydrogen processes were completed, the very favorable outlook for the Steam-Iron Process prompted the sponsor to request that we add the cost estimation for the steam-iron hydrogen process to the project tasks.

The CSF Process for coal liquefaction was chosen as the basis for making jet fuel component from coal. Economics for the manufacture of aromatic gasoline from coal and its use as an import to increase the petroleum refinery supply of light distillate for jet fuel are also presented.

The project plan calls for mid-1974 costs and the use of private-investor financing method of calculating the return on the investment. We have used the discounted cash flow method of the Supply — Technical Advisory Task Force — Synthetic Gas-Coal of the FPC. The basic features of this method are summarized below:

Project life	25 years
Depreciation	16-year sum of the digits on total plant investment
Capital	100% equity
DCF return rate	12%
Federal income tax (FIT)	48%
Return on investment during construction	DCF return rate X 1.875* years X total plant investment

Other factors we used in the cost estimates are:

Plant stream factor	90%
Contingencies	15% of installed plant cost
Contractor's overhead and profit	15% of total plant cost

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* 10% for 3 years, 90% for 1.75 years.

Start-up cost	5% of total plant investment
Working capital	a) Coal inventory (60-day feed at full rate) b) Material and supplies (0.9% of total plant investment) c) Net receivables (1/24 X annual revenue received)

Results are presented as a 25-year average revenue requirement and product price. The following equation is used to calculate the quantities.

$$P = \frac{N + 0.2353\ I + 0.1275\ S + 0.2308\ W}{G} = \frac{\text{Revenue}}{G}$$

where —

P = 25-year average product price

N = Annual operating costs

I = Total plant investment

S = Start-up cost

W = Working capital

G = Annual product output

The Annual Revenue (REV) required is the sum of the following components:

a) Net operating cost (gross operating cost less by-product credit)
b) Annual depreciation
c) 25-year average net income after FIT
d) FIT

To calculate the 25-year average value for capital charges, the 25-year average net income is calculated by the following formula:

$$X = \frac{[13(\text{REV} - N) - 0.52\ (I + S)]}{25}$$

where —

X = Net income

REV = 25-year average annual revenue = (P)(G)

FIT = $(X)\frac{(48)}{(52)}$

A 25-year average equivalent depreciation is given by I/25. The difference equals REV − N − X − (FIT) − Depreciation, or 1/25 of the initial start-up cost, distributed over the 25-year plant life.

As discussed in the section on process designs, we have used two coals in this study, Montana subbituminous coal for the manufacture of hydrogen and methane and Pittsburgh seam bituminous coal for the manufacture of jet fuel component. A study of coal prices is beyond the scope of this study. Prices have increased very rapidly since 1970-72. The cost of coal from a mine integrated with the ownership of the gasification plant could easily be less than the prices quoted by individual mine operators.

We believe that a mid-1974 base cost of 28.4¢/GJ (30¢/10^6 Btu) for Montana coal is reasonably representative. Following the example of the Synthetic-Gas Coal Task Force in regard to coal pricing, we have doubled this cost for Eastern coal. As we have presented graphs for each process showing the effect of coal cost in product price, any variations from or disagreements with these prices can easily be adjusted for.

Cost estimates for hydrogen and methane manufacture, based on Montana coal, do not include costs for stack gas cleanup, as the sulfur content is low enough to meet state and Federal standards. If more stringent pollution standards must be met, additional investment will be required.

For the CSF Process, based on high-sulfur Eastern coal, costs for SO_2 removal are included.

ECONOMICS OF THE U-GAS PROCESS

Table 56 summarizes investment costs for hydrogen by the U-GAS Process. The total installed plant cost (ex-contingency) is $310 million, which, after the addition of the various factors shown in the table, results in a total capital requirement of $540 million. The largest cost items are oxygen supply, acid-gas removal, steam generation, and electric power supply; together, they amount to about 63% of the total plant cost. Of the total cost, coal preparation is 5%, gasification plus oxygen supply is 30.9%, purification and upgrading are 31.9%, and utilities and offsites are 32.3%. The reactor is a relatively inexpensive vessel, compared to the more complex HYGAS reactor. However, the large investment necessary to supply 7829 t/SD (8630 short tons/SD) of high-purity oxygen at 2585 kN/m^2 (375 psia) costs $73 million. The large amount of CO_2 generated by the reaction of this oxygen with coal and the conversion of CO must be scrubbed out of the hydrogen stream; this results in a large investment for acid-gas removal. The large amounts of oxygen and CO_2 are reflected in high utility costs.

Annual operating costs and revenue requirements are summarized in Table 57. With coal at 28.4¢/GJ (30¢/10^6 Btu), the 25-year average price of gas is $2.06/GJ ($2.17/10^6 Btu). Coal at this base price represents about half of gross operating costs but only about 20% of the total revenue because of the high capital charges. Figure 15 shows the effect of varying coal cost on the price of gas, a 16¢ change in gas price for a 10¢ change in coal cost.

The effect of variations in plant investment is essentially the same as that shown in Figure 16, which shows relationships derived earlier in this study for the HYGAS Process.

Table 56. CAPITAL INVESTMENT SUMMARY FOR 263.9 TJ/d (250.2 X 10^9 Btu/d) HYDROGEN FROM MONTANA SUBBITUMINOUS COAL BY THE U-GAS PROCESS

(Timing: Mid-1974 Costs)

Section		Installed Cost, $\$10^6$
Coal Storage — Reclaiming		3.31
Coal Grinding and Drying		11.76
Reactors, Dust Removal, Feed and Discharge Systems		22.92
Gas Cooling and Final Dust Removal		7.91
Syngas Compressors		6.69
CO_2 and H_2S Removal		65.15
CO Shift		8.55
Product Gas Compressors		5.35
Methanation and Drying		5.52
Process Waste-Heat Recovery for Process Steam		14.05
Oxygen Plant		72.68
Process and Turbine Steam Generation (Boiler and BFW pumps)		29.32
Turbogenerator Electric Power Distribution	Electric Power Supply	12.52
Cooling and Plant Makeup Water		7.11
Sulfur Recovery — Stretford		3.19
Wastewater Treatment		5.03
Particulate-Emission Control		3.44
General Facilities		25.00
Installed Plant Cost, excluding contingencies		309.50
Contingencies at 15%		46.43
Total Bare Cost		355.93
Contractors' Overhead and Profits (15%)		53.39
Total Plant Investment (I)		409.32
Return on Investment During Construction (12% X 1.875 years X I)		92.10
Start-Up Cost (5% of Total Plant Investment)		20.47
Working Capital	60 days' coal at full rate 0.9% of Total Plant Investment 1/24 X Annual Revenue Required	17.90
Total Capital Required		539.79

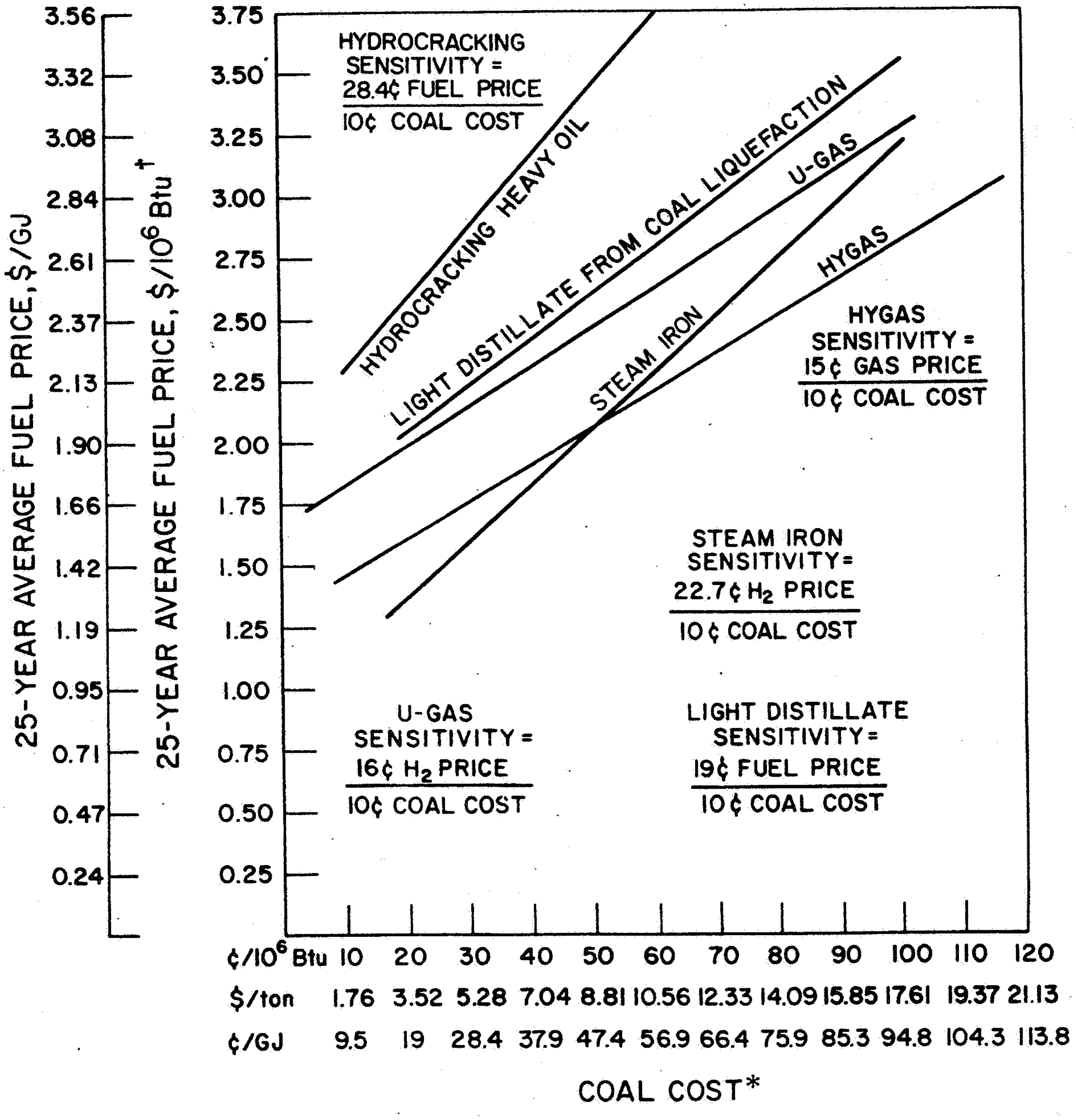
Note: No stack gas cleanup required because of low sulfur coal-SO_2 in stack gas <1.2 lb $SO_2/10^6$ Btu of Feed Coal.

Table 5.7. ANNUAL OPERATING COSTS AND REVENUE REQUIRED FOR 263.9 TJ/d (250.2 X 10^9 Btu/d) HYDROGEN FROM MONTANA SUBBITUMINOUS COAL BY THE U-GAS PROCESS

(90% Plant Service Factor, Mid-1974 Costs, 25-Year Project Life)

Component	Annual Cost, $1000
Purchased Materials	
Coal Feed at 28.44¢/GJ (30¢/10^6 Btu)	37,218
Catalysts, Chemicals, and Other Direct Materials	1,636
Raw Water at 3.96¢/m^3 (15¢/1000 gal)	593
Labor	
Process Operating Labor (61 men/shift at $5.50/h and 8760 man-hrs/yr)	2,939
Maintenance Labor (1.5% of Total Plant Investment)	6,140
Extra Maintenance Labor for Lock Hopper System	376
Supervision (15% of Operating and Maintenance Labor)	1,418
Administration and General Overhead (60% of Total Labor, including Supervision)	6,524
Supplies	
Operating (30% of Process Operating Labor)	882
Maintenance (1.5% of Total Plant Investment)	6,140
Extra Maintenance for Lock Hopper System	376
Local Taxes and Insurance (2.7% of Total Plant Investment)	11,051
Total Gross Operating Cost	75,293
By-Product Credits, Sulfur at $10/long ton	−250
Net Operating Cost	75,043
1/25 Start-Up Cost*	820
Depreciation, 25-yr average equivalent	16,372
Net Income	44,648
Federal Income Tax	41,213
Total Annual Revenue	178,096
Annual Gas Production, TJ (10^9 Btu)	86,691 (82,191)
25-yr Average Hydrogen Price, $/GJ ($/10^6 Btu)	$2.06 ($2.17)

* Spread over 25-year period as addition to annual operating cost.



* 26,251 kJ/kg (11,290 Btu/lb) DRY H.V., 22% MOISTURE AS RECEIVED.

† Major product.

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Figure 15. EFFECT OF COAL COST ON FUEL PRICE BY VARIOUS PROCESSES

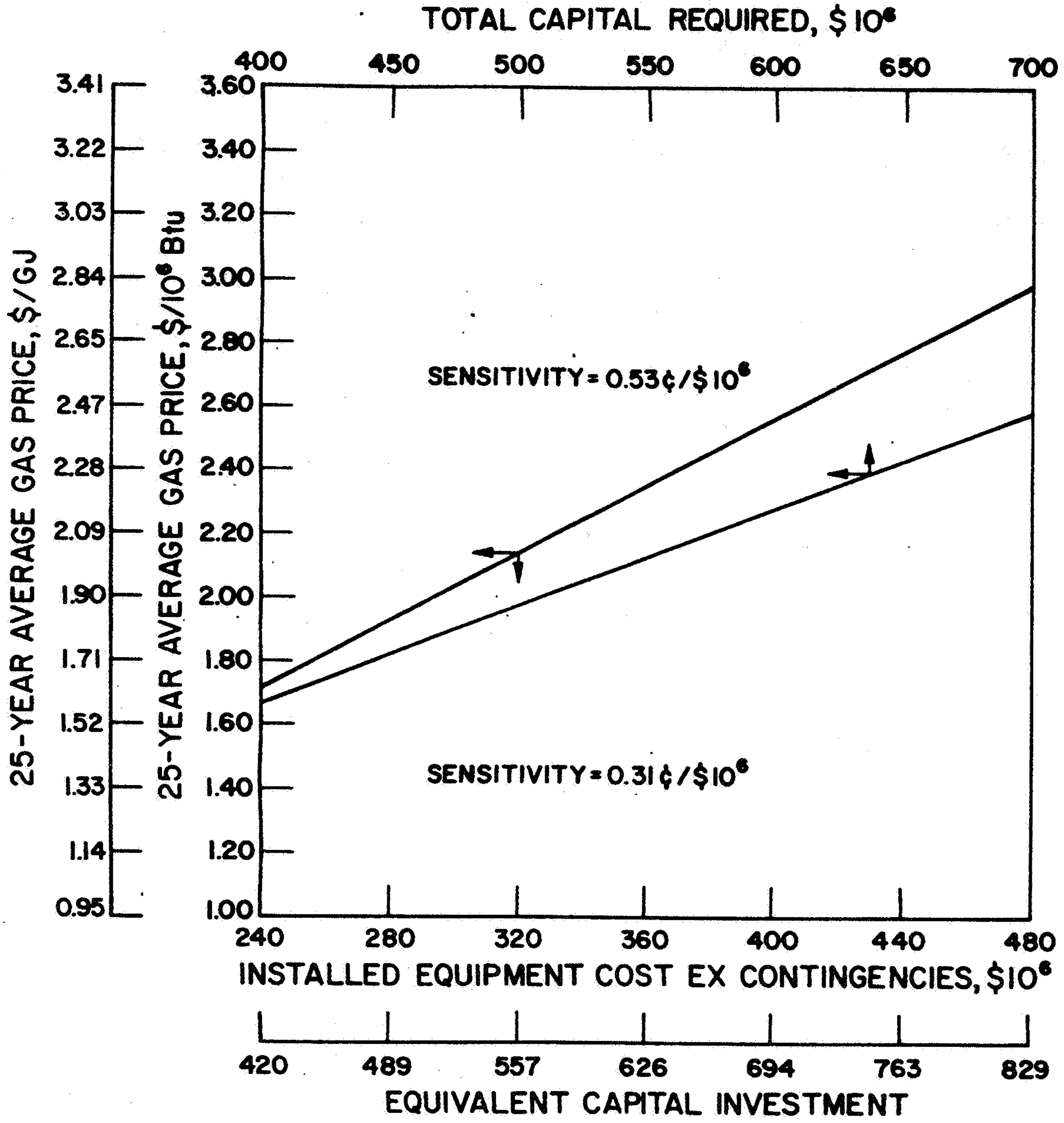

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Fig. 16. EFFECT OF PLANT COST ON GAS PRICE, FROM MONTANA SUBBITUMINOUS COAL BY THE HYGAS STEAM-OXYGEN PROCESS

ECONOMICS OF THE HYGAS PROCESS

Table 58 summarizes investment costs for methane produced by the HYGAS Process. Although the plant produces 254.7 TJ/d (241.6 billion Btu/day) of pipeline gas, it also makes an additional 12.4 TJ/d (11.77 billion Btu/d) of B-T-X liquids which have fuel value, but are more valuable as chemicals. The total direct installed cost (ex-contingency) is approximately $250 billion; after addition of the various factors shown in Table 58, this gives a total capital requirement of $436 million. The larger items, which represent 60% of the total plant cost, are hydrogasification, acid gas removal, oxygen supply, and steam and electric power supplies. Of the total installed plant cost, the coal feed preparation cost is 10%, the hydrogasification plus oxygen supply cost is 27.9%, purification and upgrading is 22.5% and utilities and offsites is 39.6%. Compared with the U-GAS Process,(the oxygen-based hydrogen process), the hydrogasification reactor system costs more because of its greater size, complexity, and much higher operating pressure. However, the differences are overbalanced by the much higher costs for oxygen supply and purification required for the hydrogen plant. Lower oxygen and acid gas removal costs mean lower utility costs also. The total capital investment for methane by the HYGAS Process is $104 million less than for hydrogen by the U-GAS Process.

Annual operating costs are summarized in Table 59. With coal at 28.4¢/GJ (30¢/10^6 Btu), the 25-year average price of gas is $1.68/GJ ($1.77/10^6 Btu), which is 38¢/GJ (40¢/10^6 Btu) less than the similar cost for hydrogen. Coal at 28.4¢/GJ (30¢/10^6 Btu) represents about half of the gross operating costs but only about one-fourth of the total revenue due to the high capital charges. The lower plant investment and the higher conversion efficiency give the economic advantage to HYGAS rather than to U-GAS. Figure 15 shows the effect of varying coal costs on the manufacture of both hydrogen and methane from coal. The sensitivity for HYGAS is 15¢ change in gas price for each 10¢ change in coal cost, compared to a sensitivity of 16¢ for U-GAS. Table 60 gives the products summary.

In Figure 16 are shown sensitivities of gas price to changes in capital cost. Because the term capital cost can mean different things, we show the

Table 58. CAPITAL INVESTMENT SUMMARY FOR 254.7 TJ/d (241.5 X 10^9 Btu/d) PIPELINE GAS FROM MONTANA SUBBITUMINOUS COAL BY THE HYGAS STEAM-OXYGEN PROCESS (Timing: Mid-1974 Costs)

Section		Installed Cost, $10^6	
Coal Storage – Reclaiming		3.00	
Coal Grinding and Drying		11.63	
Coal-Water Slurry Feed System		8.94	
Slurry Feed Preheat (fired Heater)		1.42	
Slurry Water Vaporizers Hydrogasifiers Steam-Oxygen Gasifiers	Integral Reaction Vessels	36.40	
Char Residue and Plant Ash Disposal		2.00	
Gasifier Effluent Dust Removal System		2.60	
Carbon Monoxide Conversion		9.50	
Benzene Recovery		4.70	
Prepurification (Hot K_2CO_3 Bulk, Activated Carbon, Zinc Oxide)		28.90	
Methanation and Drying		12.87	
Process Waste-Heat Recovery		16.26	
High-Pressure Oxygen Supply		29.00	
Process and Turbine Steam Generation		22.11	
Turbogenerator		6.50	
Electric Power Distribution		6.50	
Cooling and Plant Makeup Water		4.20	
Sulfur Recovery – Stretford		3.80	
Waste-Water Treatment		11.10	
Particulate-Emission Control		3.17	
General Facilities		25.00	
Installed Plant Cost, Excluding Contingencies		249.60	
Contingencies at 15%		37.44	
Total Bare Cost		287.04	
Contractors' Overhead and Profits (15%)		43.06	
Total Plant Investment (I)		330.10	
Return on Investment During Construction (12% X 1.875 years X I)		74.27	
Start-Up Cost (5% of Total Plant Investment)		16.51	
Working Capital	60 days' coal at full rate	6.21	15.04
	0.9% of Total Plant Investment	2.97	
	1/24 X Annual Revenue Required	5.86	
Total Capital Required		435.92	

Note: No stack-gas cleanup required because of low-sulfur coal – SO_2 in stack gas <1.2 lb $SO_2/10^6$ Btu of feed coal

Table 59. ANNUAL OPERATING COSTS AND REVENUE REQUIRED FOR 254.7 TJ/d (241.5 X 10^9 Btu/d) PIPELINE GAS FROM MONTANA SUBBITUMINOUS COAL BY THE HYGAS STEAM-OXYGEN PROCESS
(90% Plant Service Factor — Timing: Mid-1974 Costs, 25-Year Project Life)

Component		Annual Cost, $1000
Purchased Materials		
Coal at 28.44¢/GJ (30¢/10^6 Btu)		33,984
Catalysts, Chemicals, and Other Direct Materials		2,075
Raw Water at 3.96¢/m^3 (15¢/1000 gal)		303
Labor		
Process Operating Labor (58 men/shift at $5.50/h and 8760 man-hrs/yr)		2,794
Maintenance Labor (1.5% of Total Plant Investment)		4,952
Supervision (15% of Operating and Maintenance Labor)		1,162
Administration and General Overhead (60% of Total Labor, including Supervision)		5,344
Supplies		
Operating (30% of Process Operating Labor)		838
Maintenance (1.5% of Total Plant Investment)		4,952
Local Taxes and Insurance (2.7% of Total Plant Investment)		8,913
Total Gross Operating Cost		65,317
By-Product Credits (Table 60), $		
Sulfur	214,500	
Ammonia	796,800	
Light Oil (B-T-X)	6,910,300	
Subtotal	7,921,600	7,922
Net Operating Cost		57,395
1/25 Start-Up Cost*		660
Depreciation (25-year average equivalent)		13,204
Net Income		36,087
Federal Income Tax		33,311
Annual Revenue Required		140,657
Annual Gas Production, TJ (10^9 Btu)		83,669 (79,333)
25-Year Average Gas Price, $/GJ ($/10^6 Btu)		$1.68 ($1.77)

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*Spread over 25-year period as addition to annual operating cost.

Table 60. PRODUCTS SUMMARY FOR 254.7 TJ/d (241.5 X 10^9 Btu/d) PIPELINE GAS FROM MONTANA SUBBITUMINOUS COAL BY THE HYGAS STEAM-OXYGEN PROCESS

Products	Amount/Stream-day	Amount/Yr 90% Stream Factor	Unit Value	Annual Value, $
Pipeline Gas	254.7 TJ (241.5 X 10^9 Btu)	83,669.0 TJ (79,333.0 X 10^9 Btu)	--	--
Light Oil (B-T-X)	318,519 liters (84,144 gal)	104,633.4 X 10^3 liters (27,641.3 X 10^3 gal)	6.6¢/liter (25¢/gal)	6,910,300
Sulfur	66.3 metric tons (65.3 long tons)	21,795.3 metric tons (21,451.0 long tons)	$9.84/metric ton ($10/long ton)	214,500
Ammonia	62.9 metric tons (69.3 tons)	20,658 metric tons (22,765 tons)	$38.58/metric ton ($35/ton)	796,800
Total By-Product Credit				7,921,600

sensitivity of gas price as functions of both the installed equipment cost and total capital required. The abscissa shows two scales that relate the installed equipment cost to the capital investment. The installed plant cost is the basic value; the ratio between the two is 1.74. A $10 million change in installed plant cost changes the gas price by 5.3¢, whereas the same change in total capital requirement changes the gas price by 3.1¢. For the U-GAS Process, sensitivity of gas price to changes in the two costs are essentially the same as for HYGAS.

ECONOMICS FOR THE STEAM-IRON PROCESS

Capital investment for hydrogen by the Steam-Iron Process is summarized in Table 61A. The total direct cost for the combined-cycle power recovery section, $142.7 million, is taken directly from the UTRC Report in the appendix. The capital cost summary from that report is presented as Table 61B. Scheme II was used for our study. Its cost is 39% of the total plant cost. The producer gas generator is about 10% and the steam-iron reactor 20% of the total plant cost. Hydrogen upgrading, utilities, and offsites make up the remaining 31% of the total plant cost. For consistency with the hydrogen section and all the other cases, our allowances for contingency, contractor charges, and other factors for the power recovery section are incorporated into the general factors for the entire steam-iron plant. Since these factors add a total of approximately 72% of the total direct or installed plant cost, the equivalent figure for the power recovery system would be $245 million, a little higher than the UTRC figure, $223 million, but still giving the low unit investment cost of $185/kW. The total installed plant cost is $363.4 million. The total capital investment cost of $623 million is higher than U-GAS or HYGAS, but not high when one considers the value of the products produced. Annual operating costs and revenue requirements are presented in Table 62. Operating costs for the hydrogen plant are calculated on the same basis used for the other processes in this study. For the power recovery plant, we have used the UTRC costs for operation and maintenance. In Table 62, we show a unit by-product credit value of 1¢/kWhr, not a particularly high price in the current period of high fuel costs and power plant investment costs. The income from this power will pay for 90% of the gross operating costs.

To allow for residual amounts of fine particles in the reductor off-gas, even after cleaning, and for unexpected upsets, we have calculated the effect of shortened turbine blade life, in the range of 1 to 5 years, compared to the 25-year life for the rest of the plant. Based on the costs in the UTRC study, we have calculated replacement costs for the blades and vanes for all the drive turbines, the turbines powering the two air compressors, and the turbine that drives the power generator. The base case calculation given in Table 62 includes replacement every year — a very conservative assumption.

Table 61A. CAPITAL INVESTMENT SUMMARY FOR 263.9 TJ/d (250.2×10^9 Btu/d) HYDROGEN FROM MONTANA SUBBITUMINOUS COAL BY THE STEAM-IRON PROCESS

(Timing: Mid-1974 Costs)

Section		Cost, $\$10^6$	
Hydrogen Plant			
Coal Storage and Reclaiming		4.90	
Coal Grinding and Drying		18.70	
Lock Hopper Coal Feed System		6.03	
Producer Gas Generator, Ash Disposal		31.53	
Steam-Iron Reactor, Including Dust Removal		71.97	
Methanation and Drying		17.14	
Hydrogen Compression		5.64	
Process Waste-Heat Recovery		14.34	
Electric Power Distribution		13.43	
Cooling + Makeup Water + Boiler Feedwater Pumps		3.08	
Power Plant: Direct Cost for Power Recovery Section (From UTRC Table 59A)		142.70	
Waste-Water Treatment and Particulate Emission Control		8.95	
General Facilities		25.00	
Installed Plant Cost Excluding Contingencies		363.41	
Contingencies at 15%		54.51	
Total Bare Cost		417.92	
Contractors' Overhead and Profits (15%)		62.69	
Total Plant Investment (I)		480.61	
Return on Investment During Construction (12% X 1.875 years X I)		108.14	
Start-Up Cost (5% of Total Plant Investment)		24.03	
Working Capital	60 days Coal at Full Rate	10.10	19.90
	0.9% of Total Plant Investment	4.33	
	1/24 X Annual H_2 Revenue Required	5.47	
Total Capital Required		632.68	

Note: No stack-gas cleanup required, because of low-sulfur coal - SO_2 in stack gas $<$ 1.2 lb. $SO_2/10^6$ Btu of Feed Coal.

Table 61B. (TABLE VIII FROM UTRC REPORT) CAPITAL COST SUMMARY FOR COMBINED-CYCLE POWER GENERATING SYSTEMS

System Steam Cycle		Scheme I Reheat	Scheme II Reheat	Scheme III Nonreheat
Gas Turbine Output (mw)		865.3	1,083.8	1,083.8
Steam Turbine Output (mw)		420.1	241.6	237.0
Total System Output (mw)		1,285.4	1,325.4	1,320.8
Federal Power Commission Account Number		Capital Cost (Thousands of 1974 Dollars)		
Boiler Plant Eq.	312	54,179	43,502	42,561
Steam Turbine Gen.	314	27,335	18,904	20,079
Struct. & Improv'ts.	341	12,598	11,450	11,236
Prime Mover (Gas Turb.)	343	42,212	37,409	37,409
Generator (Gas Turb.)	344	9,887	12,129	12,129
Acc. Elect. Eq.	345	15,388	15,825	15,767
Misc. Power Plant Eq.	346	670	680	679
Station Eq.	353	In Acct 345	In Acct 345	In Acct 345
Other Expenses		3,245	2,798	2,797
Total Direct Cost		165,514	142,697*	142,657
Engineering and Supervision		24,825	21,405	21,399
Contingency		13,240	11,416	11,413
Escalation		9,203	7,742	7,678
Interest During Construction		43,785	37,713	37,690
Total Installed Power Station Cost		256,567	220,972	220,837
Capital Cost per kw ($)		199.6	166.7	167.2

Table 62. ANNUAL OPERATING COSTS AND REVENUE REQUIRED FOR 263.9 TJ/d (250.2 X 10^9 Btu/d) HYDROGEN FROM MONTANA SUBBITUMINOUS COAL BY THE STEAM-IRON PROCESS (90% Plant Service Factor -Timing: Mid-1974, 25-Yr Project Life)

Component	Annual Cost, $1000
Purchased Materials	
Coal at 28.44¢/GJ (30¢/10^6 Btu)	55,286
Catalyst and Chemicals	2,006
Replacement Cost for Blades and Vanes for Power Recovery Turbines (1-year life)	10,053
Raw Water at 3.96¢/m^3 (15¢/1000 gal)	570
H_2 Plant	
Labor	
Process Operating Labor (40 men/shift at $5.50/h and 8760 man-hrs/yr)	1,927
Maintenance Labor (1.5% of Total Plant Investment for Hydrogen Plant)	4,378
Supervision (15% of Operating and Maintenance Labor)	946
Administrative and General Overhead (60% of Process Operating Labor + Maintenance Labor + Supervision)	4,351
Supplies	
Operating (30% of Process Operating Labor)	578
Maintenance (1.5% of Total Plant Investment for Hydrogen Plant)	4,378
Power Plant	
From UTRC Annual Cost of Operation Supplies and Maintenance at 0.95 mills/kWh	9,927
Local Taxes and Insurance (2.7% of Total Plant Investment)	12,976
Total Gross Operating Cost	107,376
By-Product Power at 1¢/kWh (1,229,473 kW)	96,932
Net Operating Cost	10,444
1/25 Start-Up Cost*	961
Depreciation (25-year average equivalent)	19,224
Net Income	52,291
Federal Income Tax	48,269
Annual Revenue Required	131,189
Annual Hydrogen Production, TJ (10^9 Btu)	86,691 (82,191
25-Yr Average Hydrogen Price, $/GJ ($/10^6 Btu)	$1.51 ($1.60

* Spread over 25-year period as addition to annual operating cost.

We expect longer service than a 1-year life, but have used this figure only to show that blade and vane replacement is not a major item, but well under 10% of the annual costs. Replacement every year adds about 11.4¢/GJ (12¢/ 10^6 Btu) to the product price. Figure 17 shows that, if blade and vane life extends to 5 years, the effect on hydrogen price is reduced by 9.5¢, to only a 2.5¢ penalty.

This curve does not take into consideration the effect of reduced power recovery due to blade and vane deterioration during the period between blade replacement, because the amount of power lost is currently unknown. The main problem appears to be getting the gas clean enough to allow a reasonable work life of the turbine blade and vanes before deterioration reduces the power recovery to an unacceptable level. There is a three-way relationship among the costs for dust removal, the cost of replacing turbine components, and the frequency of replacement, which is influenced by the loss of power due to erosion of the blades and vanes. We do not have sufficient information at this time to calculate any trade-off; this is a development problem.

Figures 18 and 19 show the tremendous leverage exerted by the by-product power. Because of the predominant effect of the large amount of by-product power, we have calculated the effect of a fourfold variation in its unit cost at the plant gate, from 0.5 to 2¢/kWhr. The potential economic importance of this by-product power emphasizes the necessity of further work to more firmly establish the characteristics of the Steam-Iron Process.

Currently a large pilot plant to study the Steam-Iron Process is being built in Chicago; it is expected to be in operation in 1976. Substantial data to support the feasibility of the process have already been obtained in a small pilot unit. Operation of the large pilot plant, followed by design, construction, and operation of a larger demonstration plant, will probably require 6 to 8 years.

As shown in Figure 18, a variation in the by-product value of 0.5¢/kWhr changes the hydrogen price by 57¢/GJ (60¢/million Btu). Figure 19 shows the sensitivity of the hydrogen price to the coal cost at various by-product power values. Because of the much larger amount of coal feed to this

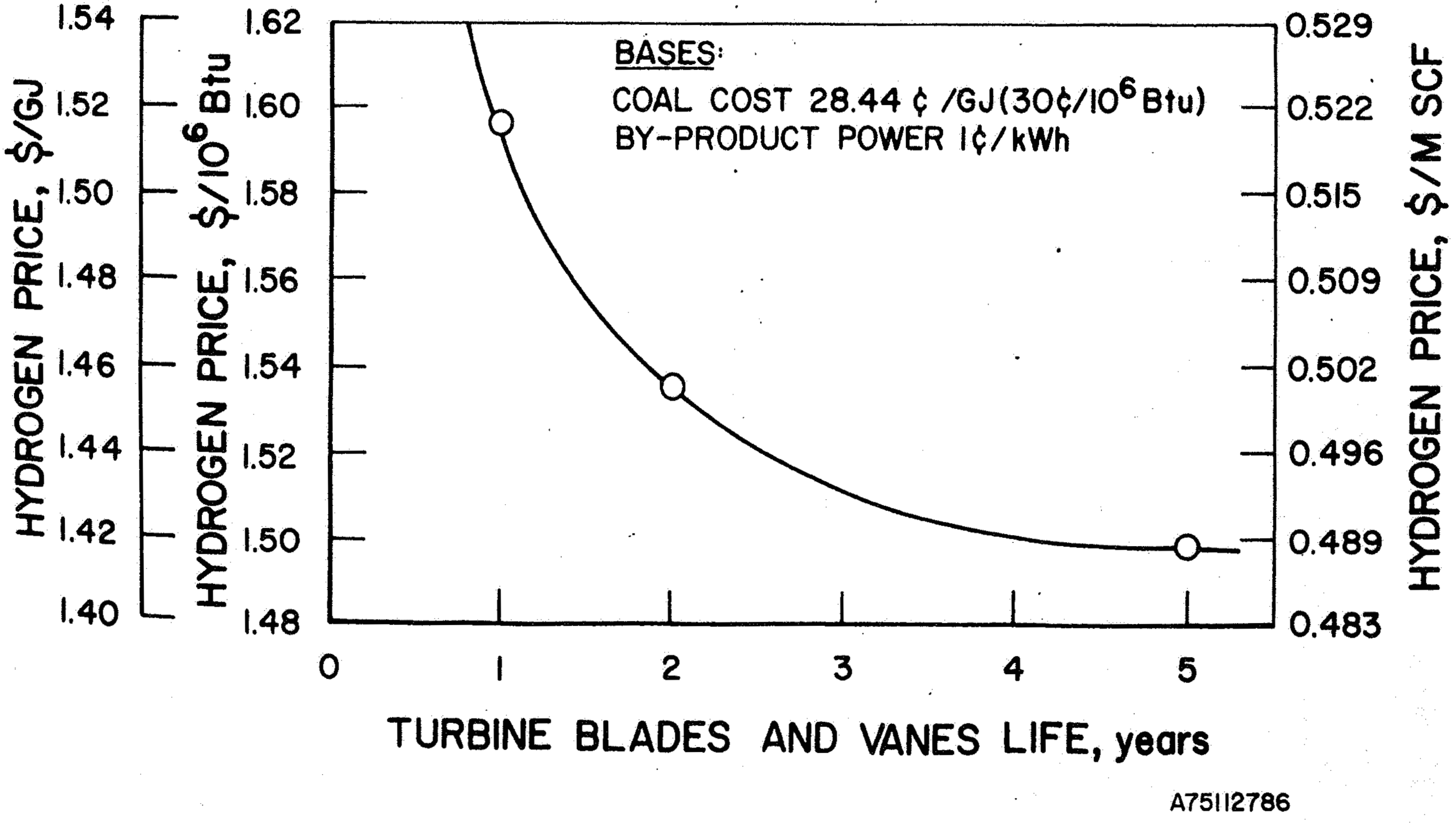


Fig.17. EFFECT OF TURBINE BLADES AND VANE LIFE ON HYDROGEN PRICE, FROM MONTANA SUBBITUMINOUS COAL BY THE STEAM-IRON PROCESS

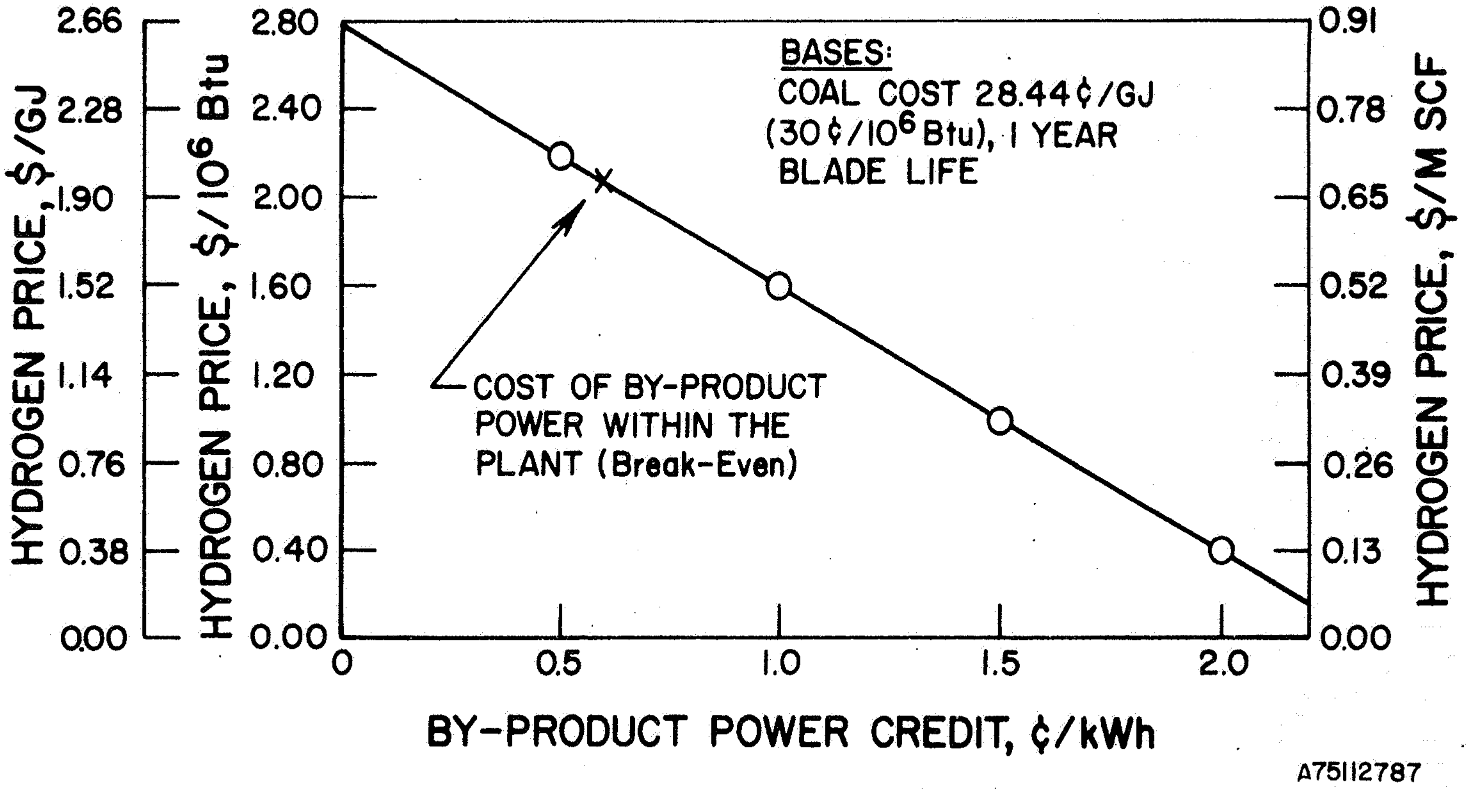
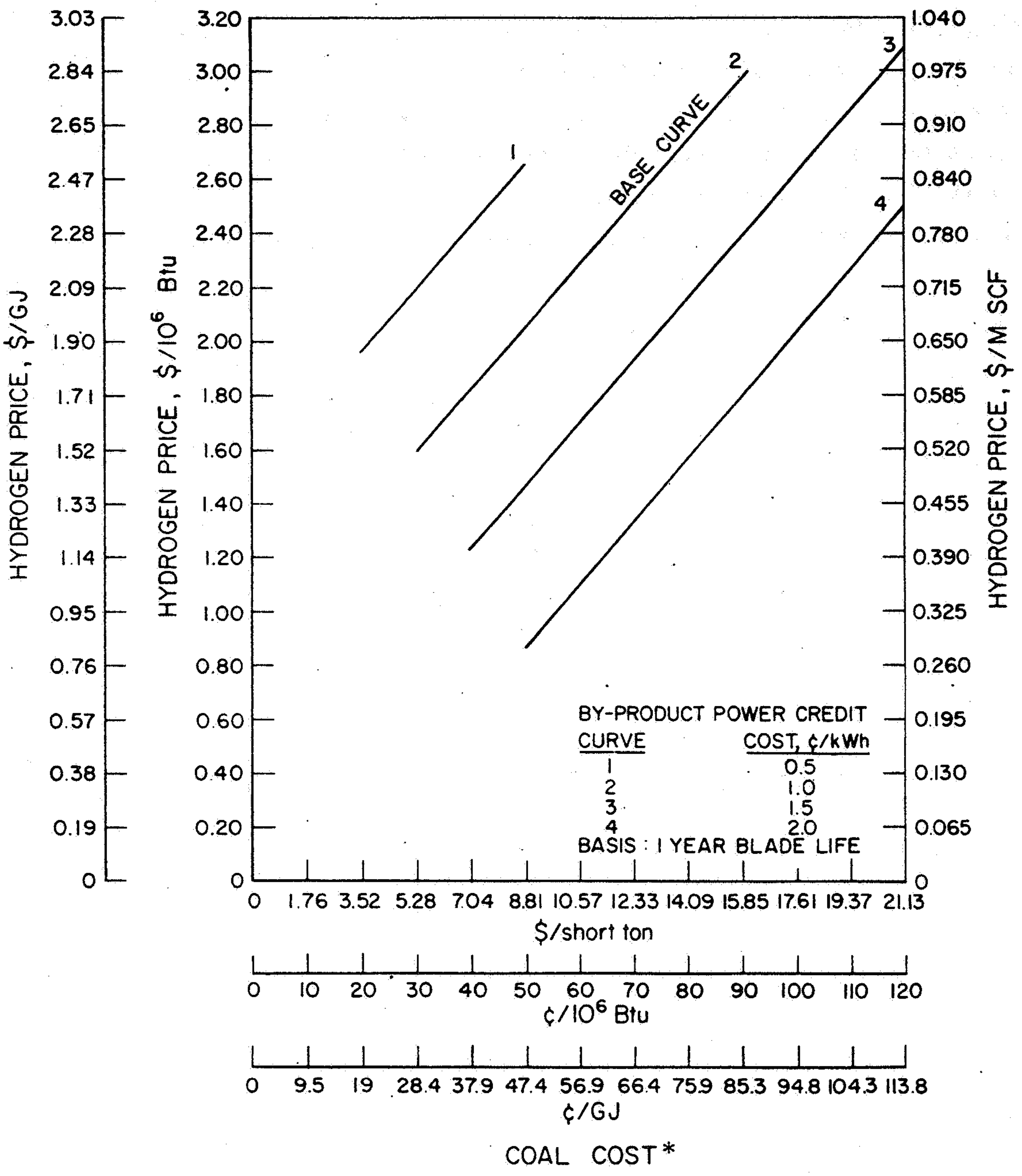


Fig.18. EFFECT OF BY-PRODUCT POWER CREDIT ON HYDROGEN PRICE, FROM MONTANA SUBBITUMINOUS COAL BY THE STEAM-IRON PROCESS



*26,251 kJ/kg (11,290 Btu/lb) DRY HHV, 22% MOISTURE AS RECEIVED.

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Fig. 19. EFFECT OF COAL COST ON HYDROGEN PRICE, FROM MONTANA SUBBITUMINOUS COAL BY THE STEAM-IRON PROCESS

hydrogen plant compared to the others, the sensitivity to coal cost is much greater than for the other processes —22.7¢ per 10¢ change in coal cost. The upper limits of the 0.5¢ and 1.0¢ curves end before the curves extend to regions where the coal cost would be all, or nearly all, of the power cost. Similarly, the lower ends of the curves end to avoid coal costs that are disproportionately low for a particular power cost. Figure 20 shows the effect of variation in the capital investment on the gas price.

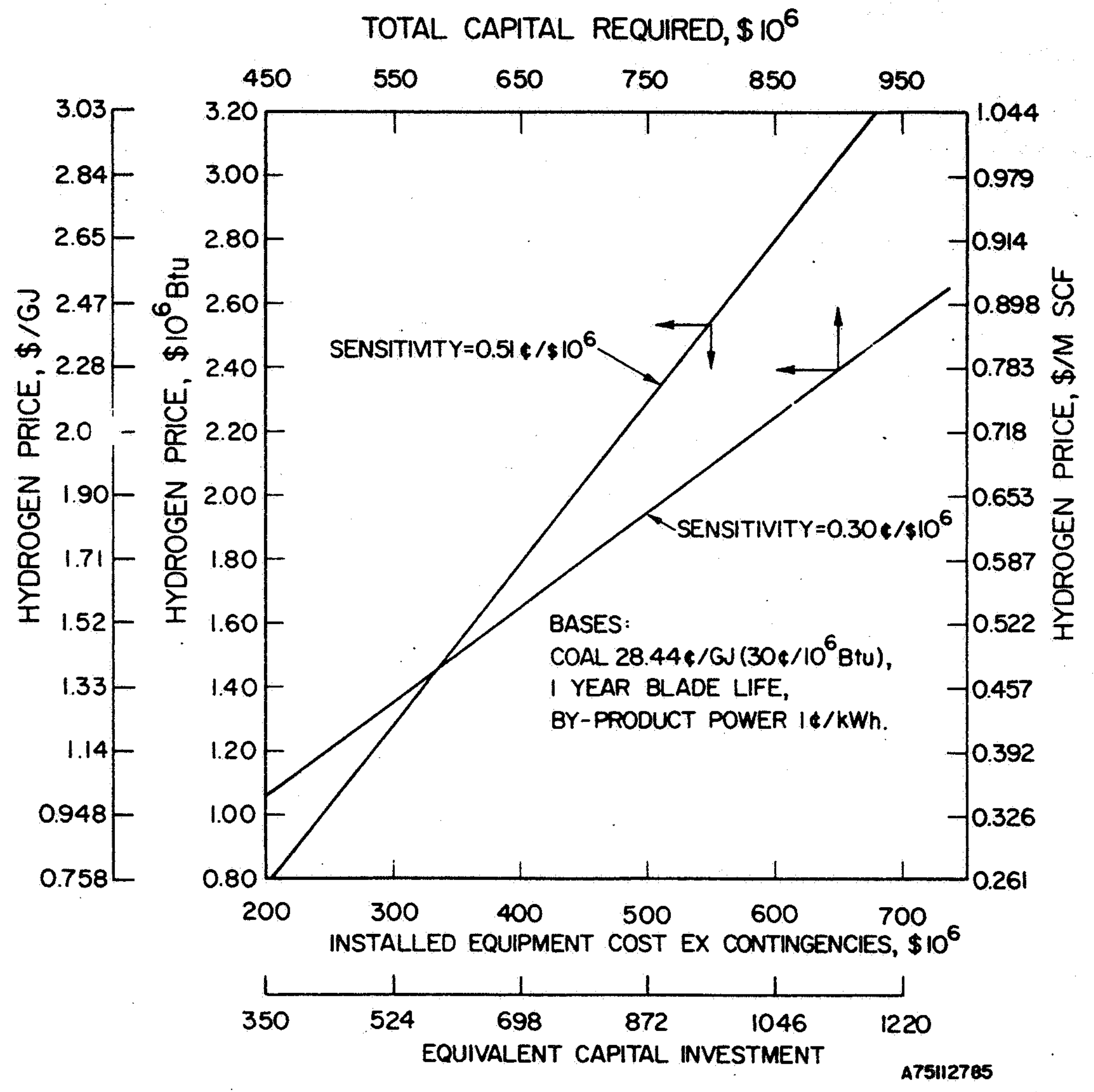

Fig. 20. EFFECT OF PLANT COST ON HYDROGEN PRICE FROM MONTANA SUBBITUMINOUS COAL BY THE STEAM-IRON PROCESS

ECONOMICS OF LIQUID FUELS FROM COAL

Increased Petroleum Refinery Light Distillate Through Importation of Coal-Derived Gasoline

Table 63 summarize capital investment for the CSF coal liquefaction process plus the add-on gasoline plant and the effect of importing the gasoline into petroleum refineries to allow increased production of light distillate. Most of the costs presented here are derived from the design of the CSF Process.[2] The costs have been updated to mid-1974 from the third quarter of 1971, for which the original costs are based. For those criteria which were escalated directly, a factor of 1.26 was used.

To be consistent with the other processes in this study (hydrogen and methane processes), the costs of items in the hydrogen section such as oxygen supply, CO_2 removal, CO conversion, and hydrogen compression, were adjusted to make them more comparable with our hydrogen estimates. Costs for onsite power generation — boiler, SO_2 removal, turbine generator, and power distribution costs — have been added.

Two cases are presented: 1) the manufacture of aromatic gasoline, and 2) a calculation that includes all the costs for gasoline but allows a reduction in the plant investment and operating costs due to saving at the refinery as a result of importing coal-derived gasoline. Table 63 presents investment costs and Table 64 presents operating costs and revenue requirements for these two cases.

Major items in the CSF Process are the hydrogen section (including the oxygen supply at $30.2 million), hydroconversion, hydrogen letdown and absorption, and power generation. The refinery add-on, including reformer, hydrocracker, distillation, auxiliaries, and offsites, amounts to a total of $125 million, an addition of 38.5%. The addition of 15% contingency, interest during construction, start-up costs, and working capital brings the total capital requirement to $694 million.

In calculating the capital cost chargeable to the increased production of light distillate at the petroleum refinery, we have taken credit for the

Table 63. CAPITAL INVESTMENT SUMMARY FOR THE LIQUEFACTION OF COAL BY THE CSF PROCESS AND SUBSEQUENT MANUFACTURE OF AROMATIC GASOLINE AND LIGHT DISTILLATE FOR JET FUEL *

(Mid-1974 Cost)

Section	Gasoline		Light Distillate by Gasoline Import
		$10^6	
Coal Preparation		14.49	
Extraction		13.05	
Separation and Solvent Recovery		13.91	
Low-Temperature Carbonization and Tar Distillation		17.43	
Extract Hydroconversion		43.28	
Hydro Letdown, Absorption, and Distillation		22.99	
Gas Plant		11.09	
Hydrogen Manufacture and Compression		93.23	
Boiler		28.20	
Turbine Generation		6.77	
Power Distribution		8.55	
Boiler Feed and Waste-water Treatment		5.24	
Cooling Water		7.23	
Sulfur Recovery Plant		8.82	
Liquid Product Storage		10.00	
Other Offsites and General Facilities		20.00	
Total Installed Plant Cost for Coal Conversion, Including Contractor Charge		324.29	
Refinery Add-on to Coal Plant		125.00	
Subtotal	449.29		449.29
Net Savings at Conventional Petroleum Refinery by Importation of Gasoline From Coal	- --		–52.00
Total Installed Plant With Contractor Charges	449.29		397.29
Contingency at 15%	67.39		59.59
Total Plant Investment (T.P.I.)	516.68		456.88
Interest During Construction (22.5% of T.P.I.)	116.25		102.80
Start-Up Cost (5% of T.P.I.)	25.83		22.84
Working Capital			
Coal, 60-Day Supply	18.92		18.92
0.009 X T.P.I.	4.65		4.11
1/24 Annual Revenue	11.00		11.00
Total Capital Requirement	694.33		616.55

* Ultimate product is increased jet fuel component (light distillate) 8703 m^3/d (54,740 bbl/d) equivalent to 310.37 TJ/d (394.3 X 10^9 Btu/d).

reduction in reformer capital (amounting to $52 million in installed plant cost) as a result of importing the gasoline into the refineries.

In assessing the economics of this increased light distillate (jet fuel component) with reference to coal liquefaction, we are faced with a somewhat unusual situation. Gasoline is the product from coal, and its importation into the refinery products system releases the additional jet fuel component at the refinery, making gasoline an <u>indirect</u> product of coal liquefaction. All products are to be sold at refinery prices. However, a price must be calculated for light distillate, referred to the coal plant, that includes any benefits in refinery operation plus any subsidy charges necessary to keep the refinery revenue balance intact.

As shown in Table 64, the revenue requirement for gasoline from coal results in a price of 10.6¢/liter (40.2¢/gal), which is 1.5¢/liter (5.7¢/gal) higher than the refinery price. Note in Table 48 that, with the unit costs used, the conventional refinery shows a little more revenue than the modified refinery. Even though the modified refinery makes more light distillate, no profit is made on that portion of the gasoline output supplied by coal-derived imported gasoline. The revenue difference, when divided by 7282 m^3/d (45,805 bbl/d) of import amounts to 0.6¢/liter (2.1¢ gal). This refinery subsidy <u>plus</u> the price differential, 10.6 minus 9.1¢/liter (40.2 minus 34.5¢/gal), amount to a total subsidy charge to the coal liquefaction plant of $49 million/yr, which is added to the other operating costs. This subsidy charge is equivalent to 1.7¢/liter (6.6¢/gal) light distillate. The subsidy considerably overbalances the savings in refinery capital and operating costs credited to the light distillate case, so that the required revenue for the increased light distillate is higher than for gasoline. However, because of the larger amount of light distillate, the unit price is lower, 9.4¢/liter (35.5¢/gal). This is actually a <u>reference price</u>, to be used only as a factor for comparing this particular case to hydrogen, methane, and other fuels derived directly from coal. The refinery selling price of light distillate is 8.1¢/liter (30.5¢/gal).

The reference price of the light distillate is equivalent to $2.64/GJ ($2.78/10^6 Btu) with coal at 56.4¢/GJ (60¢/10^6 Btu). This price is comparable to $2.51/GJ ($2.65/10^6 Btu) for U-GAS and $2.09/GJ ($2.20/10^6 Btu) for HYGAS at the same coal cost.

Table 64. ANNUAL OPERATING COSTS AND REVENUE REQUIREMENT FOR THE LIQUEFACTION OF COAL BY THE CSF PROCESS AND SUBSEQUENT MANUFACTURE OF AROMATIC GASOLINE TO YIELD 310.37 TJ/d (294.3×10^9 Btu/d) OF LIGHT DISTILLATE FOR JET FUEL

(90% Stream Factor, Mid-1974 Costs, 25-Yr Project Life)

<table>
<tr><th rowspan="2">Component</th><th rowspan="2"></th><th colspan="2">Gasoline From Coal</th><th colspan="2">Equivalent Cost of Light Distillate Considered as a Product of Coal Liquefaction</th></tr>
<tr><th colspan="4">Annual Cost, $1000</th></tr>
<tr><td><u>Purchased Materials</u></td><td></td><td colspan="2"></td><td colspan="2"></td></tr>
<tr><td>Coal:554,213 GJ/d X 56.8¢/GJ
(525,480 X 10⁶ Btu/d X 60¢/10⁶ Btu)</td><td></td><td colspan="4">103,572</td></tr>
<tr><td>Catalyst and Chemicals</td><td></td><td colspan="4">7,120</td></tr>
<tr><td>Raw Water: 54,550.12 m³/d (14,411,500 gal/d) at 3.96¢/m³ (15¢/1000 gal)</td><td></td><td colspan="4">710</td></tr>
<tr><td><u>Labor</u></td><td></td><td colspan="2"></td><td colspan="2"></td></tr>
<tr><td>Operating Labor (102 men*/Shift at $5.50/h)</td><td></td><td colspan="4">4,914</td></tr>
<tr><td>Maintenance Labor (1.5% of Total Plant Investment)</td><td></td><td colspan="2">7,750</td><td colspan="2">6,853</td></tr>
<tr><td>Supervision (15% of Operating and Maintenance Labor)</td><td></td><td colspan="2">1,900</td><td colspan="2">1,765</td></tr>
<tr><td>Administration and General Overhead (60% of Total Labor, Including Supervision)</td><td></td><td colspan="2">8,738</td><td colspan="2">8,119</td></tr>
<tr><td><u>Supplies</u></td><td></td><td colspan="2"></td><td colspan="2"></td></tr>
<tr><td>Operating (30% of Operating Labor)</td><td></td><td colspan="2">1,474</td><td colspan="2">1,474</td></tr>
<tr><td>Maintenance (1.5% of Total Plant Investment)</td><td></td><td colspan="2">7,750</td><td colspan="2">6,853</td></tr>
<tr><td><u>Local Taxes and Insurance</u> (2.7% of Total Plant Investment)</td><td></td><td colspan="2">13,950</td><td colspan="2">12,336</td></tr>
<tr><td>Total Gross Operating Cost</td><td></td><td colspan="2">157,878</td><td colspan="2">153,716</td></tr>
<tr><td><u>By-Product Credits</u></td><td><u>$1000</u></td><td colspan="2"></td><td colspan="2"></td></tr>
<tr><td>High-Btu Gas</td><td>18,422.9</td><td colspan="2"></td><td colspan="2"></td></tr>
<tr><td>Butane</td><td>14,129.2</td><td colspan="2"></td><td colspan="2"></td></tr>
<tr><td>Sulfur</td><td>2,440.8</td><td colspan="2"></td><td colspan="2"></td></tr>
<tr><td>Ammonia</td><td>1,476.3</td><td colspan="2"></td><td colspan="2"></td></tr>
<tr><td></td><td>36,469.2</td><td colspan="2">- 36,469</td><td colspan="2">-36,469</td></tr>
<tr><td>Savings Due to Reduced Catalytic Reforming in a Refinery That Imports Gasoline From Coal</td><td></td><td colspan="2"></td><td colspan="2">-17,000</td></tr>
<tr><td>Subsidy to Make up for Price Differential Between Gasoline Prices and to Make up for Lost Refinery Revenue</td><td></td><td colspan="2"></td><td colspan="2">49,799</td></tr>
<tr><td>Total Net Annual Operating Cost</td><td></td><td colspan="2">121,409</td><td colspan="2">150,046</td></tr>
<tr><td>1/25 Start-Up Cost†</td><td></td><td colspan="2">1,034</td><td colspan="2">914</td></tr>
<tr><td><u>Capital Charges</u></td><td></td><td colspan="2"></td><td colspan="2"></td></tr>
<tr><td>Equivalent Annual Depreciation</td><td></td><td colspan="2">20,667</td><td colspan="2">18,275</td></tr>
<tr><td>Net Income</td><td></td><td colspan="2">57,789</td><td colspan="2">51,520</td></tr>
<tr><td>Federal Income Tax</td><td></td><td colspan="2">53,343</td><td colspan="2">47,557</td></tr>
<tr><td>Annual Revenue</td><td></td><td colspan="2">254,242</td><td colspan="2">268,312</td></tr>
<tr><td><u>Annual Production</u></td><td></td><td></td><td></td><td></td><td></td></tr>
<tr><td>Light Distillate, m³ (bbl)</td><td></td><td></td><td></td><td>2,858,920</td><td>(17,982,090)</td></tr>
<tr><td>Gasoline, m³ (bbl)</td><td></td><td>2,392,270</td><td>(15,046,940)</td><td></td><td></td></tr>
<tr><td>TJ (10⁹ Btu)</td><td></td><td>83,307</td><td>(78,988)</td><td>101,964</td><td>(96,671)</td></tr>
<tr><td><u>25-Year Average Product Price</u></td><td></td><td></td><td></td><td></td><td></td></tr>
<tr><td>$/m³ ($/bbl)</td><td></td><td>106.3</td><td>(16.9)</td><td>93.8</td><td>(14.92)</td></tr>
<tr><td>¢/liter (¢/gal)</td><td></td><td>10.6</td><td>(40.2)</td><td>9.38</td><td>(35.52)</td></tr>
<tr><td>$/GJ ($/10⁶ Btu)</td><td></td><td>3.05</td><td>(3.22)</td><td>2.64</td><td>(2.78)</td></tr>
</table>

* 86 men for coal liquefaction, 16 men for gasoline plant.

† Spread over 25-year period as addition to annual operating cost.

If the prices of refinery products rise due to the tighter supply relative to products from coal, then the need for a subsidy will decrease. If the refinery gasoline price were 10.6¢/liter (40¢/gallon), then to keep the refinery revenue for typical and modified refineries equal, the light distillate would sell for 9.7¢/liter (36.9¢/gal) at the refinery. The subsidy would then be eliminated because the coal-derived and refinery gasoline prices would be the same, 10.6¢/liter (40¢/gal), and the 2.1¢ differential between modified and typical refineries would also be eliminated.

In Figure 21 we show the sensitivity of gasoline and light distillate prices to variations in coal cost. When referred to a basis of cents per unit heating value, these prices are: for gasoline, 23¢/10¢ in change in coal cost and for light distillate, 18.8¢/10¢ change in coal cost.

Jet Fuel by Direct Conversion of Products of CSF Process

Table 65 summarizes capital investment for the CSF coal liquefaction process plus the addition for jet fuel manufacture. The basic cost for the CSF plant, including contractor charges, is $324.29 million. UOP estimated battery limits costs for the hydrocracking and aromatics hydrogenating units plus the cost for the initial catalyst charges. We have added costs for the hydrogen unit and for utilities and general facilities that are estimated as incremental costs with respect to similar items in the CSF plant. The total add-on for jet-fuel manufacture is $99.5 million, which, with the 15% contingency, gives a total plant investment of $487.35 million. The addition of interest during construction, start-up, and working capital cost brings total capital requirement to $655 million. This number is between the capital costs for the two cases given in Table 63. However, the differences in the total capital requirement for jet fuel and the other two cases is less than 10%.

Table 66 presents annual operating costs and revenue requirements. The required annual revenue is lower for the jet fuel cost than for the others. However, because the output is much less, the product cost is significantly higher. In fact, the output of product, a function of overall efficiency,

Table 65. CAPITAL INVESTMENT SUMMARY FOR THE LIQUEFACTION OF COAL BY THE CSF PROCESS AND THE SUBSEQUENT UPGRADING OF HEAVY FUEL OIL TO JET FUEL

(Mid-1974 Costs)

Section	$\$10^6$	
CSF Plant, including contractor charges from Table 60		324.29
Added Plant for Jet Fuel Manufacture		
Hydrocracking Unit (HDC Unibon)	28.20	
Aromatics Hydrogenation (AH Unibon)	5.50	
Hydrogen Unit	23.13	
Additional Utilities and General Facilities	25.75	
Subtotal	82.58	
Contractor Charges at 15%	12.38	
Initial Catalyst	4.50	
Total Add-On to CSF Plant	99.46	99.46
Total Installed Plant with Contractor Charge		423.75
Contingency at 15%		63.60
Total Plant Investment (T.P.I.)		487.35
Interest During Construction, 22.5% of T.P.I.		109.65
Startup Cost, 5% T. P. I.		24.37
Working Capital		
Coal — 60-day supply		18.92
0.009 Avg. X T.P.I.		4.38
1/24 Annual Revenue		9.90
Total Capital Requirement		654.57

Table 66. ANNUAL OPERATING COSTS AND REVENUE REQUIREMENTS FOR THE LIQUEFACTION OF COAL BY THE CSF PROCESS AND SUBSEQUENT MANUFACTURE OF 5751 m^3 (36,173 bbl) PER DAY OF JET FUEL (205.1 TJ/d; 194.5 X 10^9 Btu/d)

(90% Stream Factor, Mid-1974 Cost, 25-Yr Project Life)

Purchased Materials		Annual Cost, $1000	
Coal: 554,213 GJ/d X 56.8¢/GJ (525,480 X 10^6 Btu/d X 60¢/10^6 Btu)		103,572	
Catalysts and Chemicals		7,120	
Raw Water: 54,550.1 m^3/d (14,411,500 gal/d)		710	
Labor			
Operating Labor (102 men/shift at $5.50/hr)		4,914	
Maintenance Labor (1.5% of Total Plant Investment)		7,310	
Supervision (15% of Operating and Maintenance Labor)		1,834	
Administration and General Overhead (60% of Total Labor, including Supervision)		8,435	
Supplies			
Operating (30% of Operating Labor)		1,474	
Maintenance (1.5% of Total Plant Investment)		7,310	
Local Taxes and Insurance (2.7% of Total Plant Investment)		13,158	
Total Gross Operating Costs		155,837	
By-Product Credits, $1000			
Naphtha	33,664.7		
High-Btu Gas	6,204.3		
Sulfur	2,440.8		
Ammonia	1,476.3		
	43,786.1	–43,786	
Total Net Operating Cost		112,051	
1/25 Start-Up Cost*		973	
Capital Charges			
Equivalent Annual Depreciation		19,494	
Net Income		54,599	
Federal Income Tax		50,400	
Annual Revenue		237,517	
Annual Production, m^3 (bbl)		1,889,215	(11,882,030)
TJ (10^9 Btu)		67.38	(63.88)
25-Yr Average Product Price			
$/$m^3$ ($/bbl)		125.73	(19.99)
¢/liter (¢/gal)		12.57	(47.6)
$/GJ ($/10^6 Btu)		3.53	(3.72)

---

* Spread over 25-year period as addition to annual operating cost.

seems to be the major factor in determining product prices. The resulting product price is \$125.7/m^3(\$20/bbl). This is equivalent to \$3.53/GJ (\$3.72/10^6 Btu), making this jet fuel the most expensive of all the products from coal studied in this project.

The effect of coal cost on the price of the different liquid products from coal is shown in Figure 21. Because all three products are based on the same amount of coal feed to the base CSF liquefaction process, the sensitivity of the product price to variations in coal cost depends on the product-to-coal ratio. For jet fuel the sensitivity is 27.4¢/10¢ charge in coal cost; the ratio is 1.9 for light distillate.

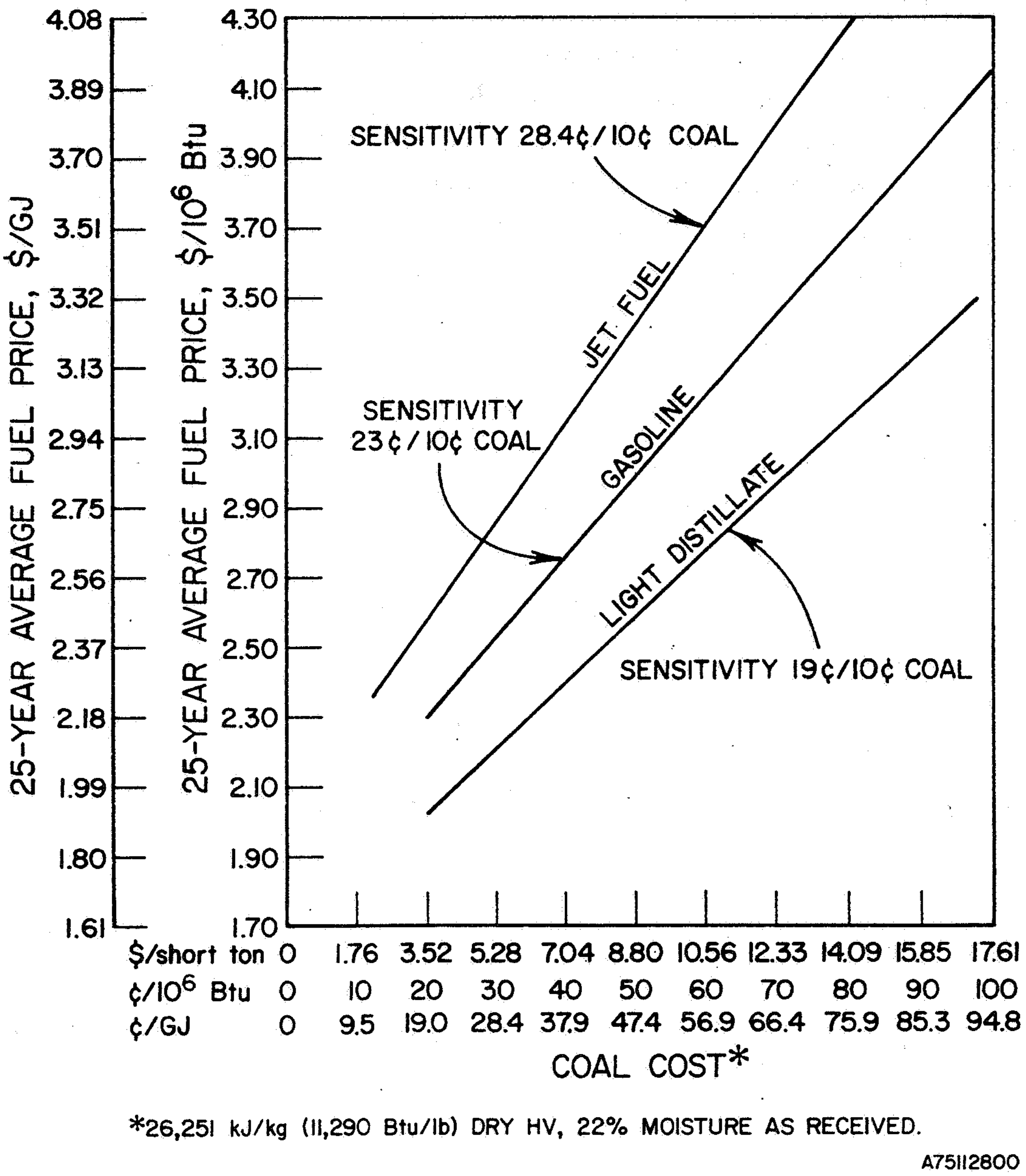

Figure 21. EFFECT OF COAL COST ON THE PRICE OF LIQUID PRODUCTS

ECONOMIC COMPARISON OF COAL CONVERSION PROCESSES

Table 67 and Figure 15 present comparative economics for the processes for which investment and operating costs were determined during this study. The economics have been presented in more detail in the discussions of the individual processes.

The hydrogen and methane process designs were sized to make approximately 263.7 TJ/d (250 billion Btu/d) of product gas. The Steam-Iron Process produces a large amount of by-product electric power and the HYGAS Process produces a small amount of B-T-X; both are included with the indicated total product in order to put the unit capital investment on a more comparable basis. The different products are explained in the footnotes of Table 67. The capacity of the liquids from coal process is basically a result of the capacity of the CSF coal conversion process. We used the published process design and cost estimate[3] as a basis for this part of the work. This energy output is in line with the general level of the other processes, so the economics should be typical for a plant of this capacity. Unit capital costs, based on product output, range from 1604 to 2316 \$/GJ (1691 to 2442 \$/10^6 Btu). The highest unit cost is for jet fuel by direct conversion, and the lowest is for light distillate obtained from increased refinery output due to the importation of gasoline derived from coal. The unit capital investment for the HYGAS Process is only slightly higher than for light distillate.

With coal at 28.4¢/GJ (30¢/10^6 Btu), hydrogen by the Steam-Iron Process shows the lowest price and jet fuel the highest. The effect of coal cost on product price is shown graphically in Figure 15. As the cost of coal rises above the 50¢ level, the HYGAS Process shows the lowest cost of all the processes and jet fuel the highest. With coal costing below 50¢, the Steam-Iron Process shows the lowest product price. At higher coal costs, it shows the second lowest product price. The figures for the Steam-Iron Process are based on by-product electric power at 1¢/kWhr. Recall from the economics for steam-iron that an increase in by-product power of 0.5¢/kWhr will reduce the hydrogen price by 56.8¢/GJ (60¢/10^6 Btu).

Table 67. SUMMARY OF ECONOMICS OF COAL CONVERSION TO FLUID FUELS; 25-YEAR LIFE – MID-1974 COSTS

Product	Hydrogen		Methane		
Process	U-Gas	Steam-Iron [b]	Hygas [c]	Lt. Distillate for Jet Fuel; CSF Coal Liquefaction to Gasoline For Refinery Import	Jet Fuel by Conversion of Heavy Oil from the CSF Process
Daily Fuel Production, TJ (10^9 Btu)	263.9 (250.2)	370.1 (350.9)	267.2 (253.3)	384.4 (364.47) [d]	282.69 (268.03) [e]
Total Plant Investment, $\$10^6$ [a]	409.32	480.51	330.10	455.88	487.35
Total Capital Requirement, $\$10^6$	539.79	532.58	435.92	616.55	654.57
Unit Capital, \$/Daily GJ (10^6 Btu)	2045 (2157)	1709 (1803)	1631 (1721)	1604 (1591)	2316 (2442)
Annual Operating Costs and $\$10^6$					
Coal, (¢/GJ) (¢/10^6 Btu)	37.218 (28.44) (30)	55.286 (28.44) (30)	33.984 (28.44) (30)	103.572 (56.8) (60)	103.572 (56.8) (60)
Other operating Costs	38.075	52.090	31.333	50.144	52.265
Gross Operating Costs	75.293	107.375	65.317	153.715	155.837
Byproduct Credits	-0.250	-95.932	-7.922	-36.459	-43.786
Net Subsidy to Oil Refineries	-	-	-	+32,799	--
Net Operating Cost	75.043	10.444	57.395	150.046	112.051
Capital Changes	103.050	120.745	83.252	118.265	125.466
Annual Revenue Required	178.095	131.189	140.557	268.312	237.517
25-Yr. Avg. Major Product Price					
Coal at 28.4¢/GJ	\$2.05/GJ	\$1.51/GJ	\$1.68/GJ	\$2.10/GJ	\$2.72/GJ
(30¢/10^6 Btu)	(\$2.17/10^6 Btu)	(\$1.60/10^6 Btu)	(\$1.77/10^6 Btu)	(\$2.21/10^6 Btu)	(\$2.87/10^6 Btu)
" " 56.8¢/GJ	\$2.52/GJ	\$2.15/GJ	\$2.11/GJ	\$2.54/GJ	3.53/GJ
60¢/10^6 Btu)	(\$2.65/10^6 Btu)	(\$2.28/10^6 Btu)	(\$2.22/10^6 Btu)	(\$2.78/10^6 Btu)	(\$3.72/10^6 Btu)

a. - Includes contingencies at 15% and contractor's charges at 15%.

b. - Includes 250.2 X 10^9 Btu/d hydrogen and 100.7 X 10^9 Btu/d by-product electric power.

c. - Includes 241.5 X 10^9 Btu/d pipeline gas and 11.77 X 10^9 Btu/d B-T-X liquid.

d. - Includes 294.3 X 10^9 Btu/d light distillate, 37.2 X 10^9 Btu/d Butane, and 33.0X10^9 Btu/d high Btu gas.

e. - Includes [illegible].47 X 10^9 Btu/d Jet Fuel, 62.[illegible]5 X 10^9 Btu/d Naphtha, and 11.11 X 10^9 Btu/d high Btu gas.

In all cases the sensitivity of product price to changes in coal cost depends on the ratio of the product to total coal input, both expressed in the same energy units. A process may produce a substantial amount of by-products, which raise the overall efficiency, but the slope of the line is a function of the product ratio.

CONCLUSIONS

1. Coal can be converted to gaseous hydrogen and methane and to normally liquid hydrocarbon fuels without a serious energy loss. Overall coal conversion efficiencies ranged from 57% for the least efficient hydrogen process to 74% for methane production.

2. Based on the processes analyzed in this study, the conversion of coal to methane is more efficient and less expensive than its conversion to hydrogen. This is not true where all of the methane is generated by the methanation of synthesis gas (CO + H_2). However, if the major part of the methane is made during the gasification step, then the amount of oxygen or other heat source used can be reduced to a fraction of that needed to make hydrogen only.

3. The results of this study for methane gas and hydrocarbon liquids manufacture do not support a generalized conclusion as to which product can be made more efficiently from coal; this depends on the liquid product. For the more refined liquid products (such as gasoline), methane produced from Western coal by the HYGAS Process shows an advantage.

4. When all three products are produced as liquids, then the order of decreasing overall efficiency is: liquids (at normal ambient conditions), methane, and hydrogen. This statement also depends on the liquid product, as noted in Item 3. The greatest loss of efficiency occurs in the liquefaction of hydrogen, which, for the same heating value of gas liquefied, requires about seven times more energy to liquefy than does methane.

5. The major items of cost in the conversion of coal are capital investment and coal. Although Montana coal, at a base cost of 28.4¢/GJ (30¢/million Btu) is used for hydrogen and methane, and Pittsburgh seam coal, at 56.8¢/GJ (60¢/million Btu) is used for liquids, when product prices are compared at equal coal costs, HYGAS and steam-iron show the lowest costs. At the lower coal cost the range of major product prices is \$1.57/GJ (\$1.60/million Btu) for steam-iron hydrogen to \$2.72/GJ (\$2.87/million Btu) for jet fuel. At the higher coal cost the range is \$2.11/GJ (\$2.22/million Btu) for HYGAS to \$3.53/GJ (\$3.72/million Btu) for jet fuel. Methane and hydrogen are priced as gas at 6996 kN/m^2 (1000 psig).

6. The Steam-Iron Process rejects a large amount of heat in the form of hot, spent producer gas. This heat is advantageously used to generate a large amount of by-product electricity. Because of its high market value, this electricity exerts a strong leverage on the hydrogen price. At bus-bar by-product credits of 1.5¢ to 2.0¢/kWhr, the Steam-Iron Process shows the lowest product price of any of these processes. However, further work should be done to determine the economics for the other processes under conditions of joint power and major product production, at total coal inputs the same as for the Steam-Iron Process, in order to avoid a biased conclusion.

7. The advantages of the Steam-Iron Process can be used to supply hydrogen for the HYGAS Process. For the same output of product gas heating value as methane, only about 40% as much hydrogen is needed as when hydrogen is the product, so the effect of by-product power will be less. Further work is needed on this process.

8. Although methane shows cost and efficiency advantages over hydrogen when the two are made from coal, the results do not preclude the use of hydrogen as fuel. Hydrogen can also be made from water using other energy sources, such as nuclear or solar energy, whereas methane requires a source of carbon. As coal becomes more costly, or its use reserved for chemicals, hydrogen from nonfossil sources should become more important.

9. Although jet fuel can be made from coal, it requires that the high aromatic content of the liquefaction products be reduced at a large loss in overall efficiency. It appears preferable to use the aromatic materials as the gasoline component imported into the petroleum products system. This, in turn, will permit the diversion of catalytic reformer feedstock into the supply of jet fuel. Optimization of this method should be investigated.

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Appendix. ANALYSES OF ELECTRIC POWER SYSTEMS FOR THE IGT STEAM-IRON GASIFICATION PROCESS

APPENDIX. ANALYSES OF ELECTRIC POWER SYSTEMS FOR THE IGT STEAM-IRON GASIFICATION PROCESS

In the Steam-Iron Process for the manufacture of hydrogen, a large amount of energy remains in the spent producer gas as heating value and sensible heat. This energy is recovered as shaft horsepower and electric power. We obtained the services of United Technologies Research Center for estimating the amount and cost of this power recovery; the results of this work have been incorporated into our study. The results of Scheme II were used in the IGT design. UTRC's full report comprises this Appendix.

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ANALYSES OF ELECTRIC POWER SYSTEMS FOR THE IGT STEAM-IRON COAL GASIFICATION PROCESS

Report No.
R75-952234-1

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September 22, 1975

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ANALYSES OF ELECTRIC POWER SYSTEMS FOR THE IGT STEAM-IRON COAL GASIFICATION PROCESS

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Analyses of Electric Power Systems For The IGT Steam-Iron Coal Gasification Process

SUMMARY

An analysis is described in which technical and economic characteristics of electric power producing systems using spent producer gas and process waste heat from a steam-iron coal gasification system were examined. Three separate combined-cycle systems incorporating differing turbomachinery and steam turbine components intended for initial operation in the early 1980's were considered. The performance results are presented in terms of system flow rates, state-point conditions, installed machinery power, and net production of electricity. Economic results are presented in terms of detailed turbomachinery manufacturing costs and selling prices, power station component costs by FPC Account Number, and net costs of busbar power.

This project was undertaken by the United Technologies Research Center under IGT Purchase Order No. 70554 as part of their prime contract with NASA Langley Research Center.

CONCLUSIONS

1. A combined-cycle, electric-power-producing system using waste energy from an ICT-designed, steam-iron coal gasification system intended to produce hydrogen as an aircraft fuel, is technically feasible and economically attractive. There appear to be no major restrictions which would preclude the combined-cycle system from becoming operational by the mid-1980's.

2. Although the economic analyses of the combined-cycle systems were based on the production of at least fifty of each of the turbomachinery components in the system, estimates based on manufacturing costs of significantly fewer units would not generally alter the economic attractiveness of the system.

3. If relatively few gasification-power producing systems are envisioned, the turbomachinery development costs most likely could not be recovered completely from gross engine sales receipts, and therefore, some Government subsidy may be required before prototype designs can be transformed into production units.

RESULTS

1. For the initial design conditions supplied by IGT, it appears that gas turbine, steam turbine, and heat exchanger components can be combined in such a manner that up to 1325 Mw(e) can be produced from the combined-cycle system.

2. For the combined-cycle electric-power-producing systems examined, the installed capital costs, including allowances for escalation and interest expenses, are below $200 per kw.

3. The estimated busbar power costs, including capital, operating, and maintenance charges, for the electric-power-producing systems was significantly less than one cent per kwhr, a result due primarily to the fact that fuel costs are charged to the gasification system and not to the power-producing system.

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INTRODUCTION

The scarcity of readily-available, inexpensive energy has made its impact upon many sectors of the economy, and today, several approaches including conservation, greater exploration, and synthetic fuel production are being investigated in efforts to alleviate the projected scarcities expected in the mid- and late-1980's. Although air travel presently affects a smaller portion of this nation's economy than such other energy-consuming sectors as auto transportation and space heating, it is none-the-less an important sector which is projected to grow at a faster rate than any other segment within the transportation sector of the U.S. economy (Ref. 1). It is for this reason that the U.S. Government, through NASA, is presently concerned whether future energy supplies will be adequate to meet the anticipated demands of the air travel industry. There may be a need for drastic action which relies on advanced technology not now considered state-of-the-art, but which would be the logical result of technical developments now under investigation in laboratories throughout the nation.

One approach being investigated jointly by NASA-Langley and the Institute of Gas Technology (IGT) in Chicago is the production of hydrogen from the gasification of coal by means of the steam-iron process. The gaseous hydrogen could be transported in pipelines to airports where subsequent liquefaction and storage could result in an attractive, pollution-free aircraft fuel. Investigations conducted by IGT indicate that basically, the steam-iron process for the production of hydrogen fuel is technically feasible and economically attractive. Furthermore, these investigations reveal that the system tail gases contain substantial amounts of energy in a chemical form (heat of combustion in the carryover of small amounts of producer gas), in potential energy form (the high pressure of this tail gas), and in thermal form (waste heat from the steam-iron component processes). IGT investigators believe that if this energy could be utilized in a combined-cycle system to produce electrical power, the overall attractiveness of the gasification system would be enhanced.

The object of the investigation program reported herein was to review system performance estimates provided by IGT and to develop capital and operating cost estimates for combined-cycle, electric power generating systems which could operate in association with an IGT-developed, steam-iron, coal gasification system for the production of hydrogen.

SYSTEM DESCRIPTIONS

The IGT investigators supplied the basic system boundary conditions such as producer gas flow rate, temperature, and pressure, parasitic air compressor power and flow rate, and heat-to-process energy requirements (Table I). Technical discussions between IGT and UTRC representatives defined three systems which incorporated turbomachinery components, combustors, heat exchangers and steam turbine components for the production of electric power.

Scheme I - Expansion Prior to Combustion

The first combined-cycle, power-producing system, defined as Scheme I, is shown schematically in Fig. 1. In this system, spent producer gas from the steam-iron process is expanded in Turbine A from 365 psia to 193 psia. The net power output of Turbine A is used entirely to drive Compressor C which supplies air at 118 psia to a combustor where the air and fuel are combined and burned. A 75 psi pressure drop is assumed through the gaseous fuel control valves, manifolding nozzles, etc. prior to combustion, and an additional 5 percent pressure drop (6 psia) is assumed during combustion. (The pressure losses are consistent with experimental values determined by the Power Systems Division of UTC for low-Btu fuels). The heat content of the spent producer gas is estimated to be no greater than 85 Btu/scf.

The combustion exhaust is divided into two streams with a portion being directed to Turbine D which drives Compressor B supplying process air at 400 psia to the steam-iron process. The second exhaust stream from the combustor is directed through Power Turbine E which is directly coupled to a synchronous-speed electric generator. The exhaust streams from Turbines D and E are recombined and then passed through two heat exchangers which remove additional thermal energy. A total of 2170 x 10^6 Btu/hr is transferred back to the steam-iron process, while more than 3520.7 x 10^6 Btu/hr of additional exhaust heat is transferred into a reheat steam cycle operating at 2400 psi with primary and reheat temperatures at 1000 F. Heat from the methanization effluent (356.7 x 10^6 Btu/hr) in the fuel-gasification system is used to heat the feed water in the steam system. Industrial practice has shown that an exhaust gas stack temperature of approximately 300 F, the minimum level at which potentially harmful condensates will not be produced at stack exit. Heat content of the exhaust at 300 F is considered sufficiently small as not to have a significant effect on the overall system output and efficiency.

Except for Power Turbine E and the steam turbine, all available shaft energy is reused within the overall system itself. For purposes of this analysis, a dual process train system was assumed, and all components shown schematically in Fig. 1

are assumed to be incorporated in each of two identical systems, rated at one-half the total for each component. This practical approach allows greater combined-cycle system control and flexibility of operation, particularly during shutdowns for overhaul and repair, and during part-load, steam-iron process operation.

Scheme II - Combustion Prior to Expansion

In Scheme II, the turbine and compressor components are combined as shown schematically in Fig. 2. In this system combustion air is compressed in Compressor A, introduced into the Combustor where it combines with the spent producer gas and burns at such a rate that the exhaust temperature of 2400 F is attained. The combustion pressure of 290 psia results from an allowance of a 75 psi pressure drop in the fuel meter, manifold, nozzle, and related components in a manner similar to that assumed for Scheme I. A 15 psi pressure drop was also assumed during the combustion process.

The exhaust gas flows are divided into three streams at the burner exit, each of which is proportional to the work required in the turbine expanders. In Turbine C, the net output power is used only to drive Compressor A, and the unit is completely self-contained. Turbine D which drives Compressor B, the source of pressurized air for the steam-iron process, is similarly self-contained, and Power Turbine E is the only turbine component intended to drive an electric generator. All turbine components expand to a common discharge pressure level of 16.5 psia which is sufficient to cover losses in subsequent heat exchanger units. The common exhaust gas stream is then divided into two streams, one of which is used to transfer process heat to water in a heat-to-process heat exchanger; the second exhaust stream heats water in the steam turbine system. Not shown in Fig. 2 is a second heat exchanger in the heat-to-process stream which recovers heat from a portion of the methanization effluent heat exchanger. Additional heat from the methanization effluent stream is used to preheat the water from the condenser discharge in the steam turbine system. Flow rate restrictions and thermodynamic temperature limits in the steam cycle dictate that the discharge temperature from the water preheater should not exceed 250 F.

In Scheme II, two steam cycles were considered. The first, a steam reheat cycle, had thermodynamic characteristics identical to those in Scheme I. A second, nonreheat steam system was also investigated. The steam cycle efficiency of the nonreheat cycle is slightly less than that of the reheat cycle, but is is also less complex than the reheat cycle. Depending on the operating conditions, a slight efficiency loss may be compensated by lower installed costs and carrying charges.

METHOD OF APPROACH

The methods used to investigate the systems discussed in this report incorporated combustion, thermodynamic, stress, sizing, and cost analysis programs. Although several of the analysis programs are proprietary to UTRC, each has had extensive prior use on Government contracts as well as Corporate study programs. (c.f., Refs. 2 and 3).

Scheme I

In order to meet the work requirements of the Compressor C-Turbine A unit in Scheme I, analysis of the combustion of the producer gas and inlet air at the state-point conditions indicated in Fig. 1 resulted in an air-to fuel flow ratio of 0.575 at a pressure of 193 psi. Adiabatic component efficiencies of 85 percent and 88 percent were assumed for the compressor and turbine respectively. (Such high efficiencies are expected to be well within the state-of-the-art by 1980 for rotating machines of the size considered for this system.) Because of the necessity to match pressures, flow rates, and work requirements, a combustor discharge temperature of 2406 F was required. This value was compatible with the original IGT goal of 2400 F. All combustion calculations were performed by means of a digital computer program developed in the Chemical Sciences Department of UTRC.

The performance of Compressor B was determined on a trial-and-error basis by using the thermodynamic properties of air from Ref. 4. This approach was made necessary because of the requirement to match a discharge pressure of 400 psia with the work split of a two-stage turbine. Although an intercooler between the two spools of the compressor would have allowed the compressor discharge temperature to exactly match a process requirement of 1000 F, the added cost of incorporating this unit into the system was not believed worth the slight improvement in system output which might have resulted. Furthermore this assumption adds conservatism to the results. The thermodynamic properties of the combustor exhaust gases were based on the mole-percent, weighted averages of the constituent gases assuming complete combustion, and these properties, including those for water vapor, were also based on data from Ref. 4. The work output from and the flow rates through Turbine D were determined by matching exactly the requirements of Compressor B so there would be no excess power from this unit. The exhaust flow not required in Turbine D was directed through other ducts and expanded through Power Turbine E which was assumed to be connected to a 98 percent efficient electric generator. An adiabatic efficiency of 88 percent also was assumed for the power turbine which is the only unit in the exhaust gas stream from which net power is extracted.

The mixed turbine exhaust flow at a common pressure of 16.5 psia was subsequently divided into two different streams which passed through heat exhangers in the system. The first of these heat exchangers extracted exhaust gas energy to heat process water from 235 F to saturated steam conditions at 400 psia. Because of favorable temperature difference between the exhaust gas and water streams, it was possible to reduce the exhaust gas temperature to 300 F. The remaining portion of the turbine exhaust gas flow was passed through a steam boiler/superheater/reheater, and because of the characteristics of the steam cycle selected, the temperature of the exhaust gases at the discharge of this unit could also be reduced to 300 F. All intermediate-state-point conditions shown in Fig. 1 and discussed subsequently in this report were based on the enthalpy differences in the exhaust gas mixture. All. heat exchanger performance calculations followed conventional log-mean-temperature difference relationships.

The steam cycle performance was based on the heat inputs and the selected steam operating conditions of 2400 psia/1000 F/1000F. All steam cycle calculations were results of systems studies being conducted for a concurrent EPA contract (Ref. 3) at UTRC.

Scheme II

In Scheme II, an approach similar to that just described was undertaken to accommodate the thermodynamic performance analyses of the components shown in Fig. 2. A separate set of combusion calculations conducted with the aid of the UTRC digital computer program resulted in an air-to-fuel flow ratio match point of 0.845 at 290 psia and a discharge temperature of 2400 F. These overall thermodynamic calculations were less complicated relative to those in Scheme I, since it was not simutaneously necessary to match the work output characteristics of turbomachinery components in the Scheme I system.

The flow from the combustor was divided into three streams, the first of which was directed to Turbine C whose output power was exactly matched to the requirements of Compressor A based on conventional thermodynamic analysis techniques. (The discharge pressure common to all turbine components in Scheme II was assumed to be 16.5 psia, a value identical to that assumed in Scheme I.) The second exhaust gas stream was directed to Turbine D whose output power was matched exactly with the requirements of Compressor B. The remaining flow stream from the combustor was directed to Power Turbine E, the output power of which was used to drive the only electric generator in the exhaust gas stream of Scheme II.

As a rule of thumb, turbine design experience has shown that above a pressure ratio of approximately 5.8, three stages of turbine expander are required, and beyond a pressure ratio of approximately 12.3, four expander stages are necessary. Although the system designs selected in this analysis may not be at exactly optimum conditions, the use of this and other empirical guidelines assured that the turbomachinery design would be practical and viable.

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Component Sizing and Matching

Whenever possible during turbine expansion calculations in both Schemes I and II, it was assumed that the same amount of work would be extracted from each turbine stage on a common shaft (i.e., that of the high, the low or the power turbine.) For each turbomachinery component, the work per turbine stage was used to determine the rotational speed, blade and vane discharge angles, and blade rotational stresses from standard sets of turbomachinery equations. The stress level and average blade metal temperature then correlate directly with the selection of the blade material from creep-stress vs temperature relationships, a typical set of curves for which is shown in Fig. 3. In order to design for long system life and to assure maximum gas turbine rotating component life, an averaged metal temperature of 1500 F was assumed to be the maximum allowable in all system designs considered during this present analysis.

Stress and rotational speed are based on mean flow-path diameters, and once determined, the disk (hub) diameters and blade lengths can be calculated directly from the stress-related hub-tip ratios. With rotational speeds established, a trial-and-error process is used to select a corrected airflow per unit area and corrected compressor tip speed which result in an integral number of compressor stages whose performance matches that of the turbine. Similar sets of stress-speed matching analyses were undertaken for other compressor-turbine combinations in the system. In the power turbine, the requirement of equal work per stage allowed a design to be determined directly since the output rotational speed of 3600 rpm was specified by generator requirements. In all systems, it was assumed that metallurgical developments would have advanced sufficiently that specialty or exotic materials of today would be state-of-art materials by the time a systems such as those considered herein would be ready for operation. Blade cooling techniques considered common today were expected to be sufficient to accommodate the requirements of the turbomachinery systems as contemplated.

System Costing

Each individual system component was costed in as great a detail as was possible. In most cases, cost models developed at UTRC were used to estimate manufacturing costs, and then mark-up ratios, known to be typical of those in the industry, were applied to these costs to arrive at the selling price (i.e., the cost to the final customer). A gas turbine manufacturing cost digital computer program, which was developed on Corporate funds, was used to estimate the cost of all rotating machinery. This program, which currently is being exercised on a NASA-related (Ref. 5) contract by UTRC, was developed in cooperation with a large number of vendors who service United Technologies Corporation Divisions. The program is capable of predicting, with great accuracy, the total manufacturing costs (including

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indirect items) for components, such as blades, vanes, disks, shafts, bearings, casings, burners, and accessories, which account for over 80 percent of the engine components, or essentially all the major cost items in an engine. Within the program, allowances are made for blade and vane cooling, when necessary, material changes, burner can shrouds, and even changes in the base year of cost estimates desired. All correlations are in equation form, and input data requirements are primarily based on engine component dimensions, unit production rates, and materials selection.

Heat exchanger costs are based on overall heat transfer coefficients and materials selection cost allowances at $6 per sq. ft. This estimate, the choice of appropriate heat transfer coefficients, and an addition allowance of 35 percent for fabrication and erection are based on information obtained from vendor contacts made during the fulfillment of technical studies for the EPA contract of Ref. 3. The costs of steam turbines and electric generators were taken directly from manufacturers published price data (e.g., Ref. 6).

The estimating procedure which allows the cost of an entire power system to be made was developed at UTRC with the assistance of Burns and Roe, Inc., a large East-coast architectural and engineering firm, during the fulfillment of a NAPCA contract (Ref. 2) in 1970. Subsequent contacts with Burns and Roe personnel, in addition to using the updating correlations from industry-accepted sources such as Ref. 7, allows this system cost estimating procedure to be kept up to date. Although such power station estimates must be considered as budgetary values, they correlate well with more detailed estimates made when actual bids are later established. At the request of NASA and IGT, all values presented in this report are typical of those representing a mid-1974 time period. As noted, subsequent use of industry indexes, such as those in Ref. 7, can be used to update the results presented herein, if so desired.

DISCUSSION OF RESULTS

The results of Scheme I and those of two versions of Scheme II are discussed separately in the following sections. Although certain similarities are common to all of the systems, each section contains a complete discussion of performance and detailed cost estimates.

Scheme I

Scheme I encompasses a set of turbomachinery components arranged in such a manner that the high (365 psia) pressure of spent producer gas from the steam-iron gasification process is partially expanded prior to its combustion. (This arrangement differs from that of conventional gas turbine components which is typified more by the arrangement defined as Scheme II.) Because of the relatively low temperature (1520 F) of the spent producer gas, less work can be extracted from Turbine A than if this gas were expanded from a temperature of 1800-2000 F which is more typical of that used in gas turbines of today. However, the attractiveness of the approach used in Scheme I is that low-priced materials can be used in the turbine section, and blade cooling is not required. Also, for a fixed combustion temperature, more heat energy remains in the exhaust gases to be captured in subsequent heat exchangers.

The net output power from the exhaust gas Power Turbines E is calculated to be 865.3 MW after allowing for a 2 percent loss in the electric generators. The system was assumed capable of accommodating the total flow of 4421 lb/sec in two identical streams, thereby providing greater operating flexibility and continued, partial operation in case of a failure in one stream. Although the 2400 F temperature at the inlet to Turbine D and to the Power Turbine E is considered advanced by standards of today, normal progress within the gas turbine industry will make temperatures of this magnitude commonplace in eight-to-ten years. Where possible, conservative design practices which did not compromise costs, were selected for the systems analyzed.

The Compressor B - Turbine D units would resemble conventional gas turbine engines without a burner. However, the flow rate, temperature, and speed requirements of the power turbine dictate that these latter units must be double-ended, that is, a "mirror" set of rotating components on a single shaft driving a single electric generator. Because of the large fuel gas and air flow requirements, a large, single-chamber combustor design, nearly 15 ft in diameter and 45 ft long, is indicated. Ducts would lead in a scroll-like fashion from the burner exit to the inlets of Turbine D and of Power Turbine E located close by to assure compactness and a high overall efficiency. The high volume flows of the air and fuel and the low energy content of the fuel necessitate a large pressure drop through the fuel controls and fuel manifolds prior to combustion. A loss of 75 psi in the expanded producer gas stream is considered average for this type of gas based on recent

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United Technologies Corporation Power Systems Division test data. Such a loss is obviously detrimental to overall system performance, but it would be unrealistic and unfair to assume more optimistic conditions for the analysis.

The performance of the reheat-type steam turbine system is based on the temperature-heat energy diagram of Fig. 5. It can be seen that throughout the entire heat energy range, sufficient temperature differences exist so the minimum difference can be established at the exhaust gas heat exchanger exit, thereby achieving a 300 F stack temperature. The heat energy added to this cycle in the boiler/superheater/reheater represented by that of Fig. 4, is approximately 89.3 percent of the total; the remainder (equal to 356.7 x 10^6 Btu/hr as noted in Table II) is obtained from the methane effluent stream (see Fig. 1). After a total output of 210,050 kw per stream is produced, the remaining 2414.4 x 10^6 Btu/hr are rejected in cooling tower heat exchangers.

The overall manufacturing costs of the major components shown by major categories are presented for the three separate exhaust gas turbomachinery units in Tables III, IV, and V, respectively. The major categories are self-explanitory with possible exception of miscellaneous parts, a general category which incorporates such items as assembly and testing labor, the starter, the lubrication system, a major X-ray of the basic unit, and internal miscellaneous parts such as seals, spacers, clutch assemblies, and fasteners. Further details of some major design features material selections, and cooling choices for the three respective units are presented in Table VI. The estimated manufacturing costs which range from approximately $0.75 million per unit for the Compressor C - Turbine A unit to nearly $2.0 million for each end of the power turbine are believed typical of those for similarly-sized large units today. Prime reasons for the high costs of these units are their large sizes and the requirement for relatively expensive turbine components, many of which must be cooled. The estimates shown for these units correspond to that of the 50th unit of a long-run production setup where system design is fully established. Values typical of the second (i.e., each of the first two produced since a two-stream system was selected) and the sixth units would be approximately 122 and 69 percent greater respectively, than those presented in Tables III, IV, and V.

The estimated selling price of over $15 million for the fiftieth unit (and those beyond since a "learned-out" cost most likely would be reached by this point) including combustor and fuel control is presented in Table VII. These estimates include a mark-up of approximately 100 percent over total manufacturing cost to cover general and administrative expenses, amortized engineering expenses, selling costs, warrantee costs, and gross profit. A markup of similar percentage magnitude can not be expected at the low unit production rates of two or six units unless the development expenses, which could amount to several hundred millions of dollars, are absorbed by an agency other than the manufacturer of these units. Production of such large units represents a risky undertaking for one commercial organization, and unless a guaranteed market exists, or the development expense is underwritten, it is unlikely that a private organization will undertake this development.

Production of heat exchanger equipment is based on an entirely different philosophy, however. Since most heat exchangers in electric power generating stations today are custom designed and built, their costs can be estimated with a greater degree of confidence. A typical layout of the heat-to-process heat exchanger and the steam cycle boiler/superheater/reheater are shown diagramatically in Fig. 4 for Scheme-I. For the systems shown, the selling prices of the fabricated units, the heat transfer areas, and the heat capacities of the units are presented as part of Table II. Heat exchangers represent a large investment, and often this is a substantial portion of a typical electric steam station cost. It is partly because of the necessity to incorporate these large heat exchanger units that the capital costs of steam power stations will be higher than an all-gas turbine system which, of course, requires no such equipment. The combined-cycle systems, typified by Schemes I and II, represent a compromise between increased costs and increased overall system output and efficiency.

When the overall set of system components are incorporated in a station capable of producing large amounts of electrical power based on its thermodynamic characteristics, an estimated overall system installed investment of nearly $260 million, including escalation and interest, would be required as shown in the first column of costs in Table VIII. Based on the net output of 865.3 MW from the exhaust gas portion of the system and 420.0 MW from a 2400 psia/1000F/1000F steam reheat cycle operating at 37.72 percent efficiency, this capital cost is equivalent to $199.60 per net kilowatt output. Details of this composite estimate are presented in first columns of cost data in Tables IX through XII. This total system cost estimate is quite near that of other combined-cycle system estimates of $200 per kilowatt published recently (c.f., Ref. 8), thereby adding to the confidence in the estimates of this analysis.

The details presented in the individual tables are shown by overall FPC Account Numbers; only those account number totals shown are applicable to a combined-cycle electric power generating station. For brevity, some estimates within accounts have been combined, but from an overall standpoint, the results represent the costs of all required individual items. Several points should be mentioned in connection with these tabularized results. In Account 341, Table XI, the item entitled Tank Farm is included to insure an adequate fuel supply should an abnormal interruption in the primary fuel system develop. Sufficient fuel energy would still be available to meet contractural arrangements to sell electrical power. Elimination of this item would reduce the Total Installed Power Station Cost (Table VIII) by only approximately $1.14 million.

The installed price for Gas Turbines (Table XII) corresponds to that for the 50th unit and includes a 5 percent additional allowance for installation. The cost of the overall system with the first two units or the sixth and seventh units can be estimated using cost factors discussed in an earlier section of this report. Whereas steam turbine and generator prices were taken directly from manufacturers' catalogues, appropriate discounts were applied to these published prices to represent typical industry practice. Discount factors were obtained through proprietary UTRC contacts and therefore are not published here in order to maintain the confidential nature of both the values and the contacts. Items included in the general category entitled Escalation (Table VIII) include the Waste Heat Boiler (FPC Account 312-01), the Process Steam (Heat-to-Process) Heat Exchanger (FPC Account 312-07), and the Cooling Towers (FPC Account 314-06), each of which were assumed to be subject to a three-year escalation period, and the Steam Turbine-Generator (FPC Account 314-01) and the Generator-Gas Turbine (FPC Account 344), each of which had a two-year period of fixed price followed by a two-year escalation period. All escalations were calculated using an 8 percent per year rate; if another value is thought more typical of future rates, it is a simple matter to calculate the net effect of such a change on the overall capital cost. Similarly, the interest rate of 9 percent per annum and a four-year applicable construction period were selected for all systems, but this also can be changed if its effect on the overall installed capital cost must be examined.

Scheme II

The system defined schematically in Fig. 2 combines, as noted previously, the turbomachinery components in a conventional arrangement. However, because the exhaust gases in this configuration expand directly from a pressure greater than that in Scheme I, their temperature, when recombined after expansion, is lower in Scheme II than in Scheme I. This temperature, calculated to be only slightly below 1200 F, was adequate to be used in a reheat steam cycle similar to that selected for Scheme I. However, because of the relative steam-side and gas-side temperatures (Fig. 6) the exhaust gas temperature could not be reduced below 459 F. Therefore, it was thought that in an attempt to recover a greater amount of heat energy from the exhaust gases, the selection of a simple nonreheat steam cycle (see the temperature-heat diagram, Fig. 7) which would be accompanied by an exhaust gas temperature below that at the exit of the reheat cycle may be a viably attractive alternative system to examine. Therefore, the following discussion of Scheme II systems incorporates the results of analyses of two steam systems, noted Scheme II with Reheat-Steam Cycle and Scheme II with nonreheat Steam Cycle, each of which incorporates the same base set of exhaust-gas turbomachinery components.

Turbomachinery System

Since the turbine inlet temperatures of all three expanders is the same (2400 F), the assumption was made that the work per stage (on a Btu/lb basis) on the respective common shafts of Turbines C and D would be equal. This assumption then made it necessary only to find the work splits across the low and high sections of Compressor A and Compressor B, respectively, which corresponded to the airflow of these respective components. The flow of 864 lb/sec/unit through Compressor B is specified as a gasification process requirement, whereas, the flow of 1169.1 lb/sec/unit in Compressor A is dictated by the fuel-air ratio in the combustion necessary to achieve a combustion temperature of 2400°F. Both turbine and compressor units are assumed to be self-contained, and there is no net production of power from either. Again, two streams were selected for redundancy, ease of maintenance, and flexibility of operation.

The third exhaust stream is ducted to Power Turbine E which incorporates four stages because of the expansion ratio of 16.7-to-1.0 across this entire unit. The net output of these units, after allowing for a 2 percent loss in the electric generator, is estimated to be 1083.8 MW. When the output power from these exhaust gas turbines is compared with that from the Scheme-I units, it can be seen that there is a nearly 220 MW power increase. This is due to two factors. First, it was cited in an earlier discussion that expansion from a low temperature results in less output power than a similar expansion from a higher temperature. Although Turbine A in Scheme I is operating between different pressures than the turbines in the Scheme II system, the general concept of lower work from a lower initial temperature still applies. Second, because of the work and flow rate matching between the combustion air and fuel in Scheme I, a lower airflow per pound of fuel passes through its combustor, than in Scheme II. This greater total mass flow of exhaust gas in the Scheme II system has a significant influence on the increase of net output power.

As noted, the greater expansion ratio across the turbine components of Scheme II results in a lower temperature of the exhaust gas mixture entering the heat-to-process and the boiler/superheater heat exchangers. A complicated phenomenon exists with the arrangement of components shown in Fig. 2. Because of the need to extract 2170 x 10^6 Btu/hr from the exhaust stream(s) to satisfy process steam requirements in both Scheme I and in Scheme II, a greater mass of exhaust flow must pass through the heat-to-process heat exchanger in Scheme II than in Scheme I. Despite the fact that the flows through the boiler/super-heater units are nearly identical in both schemes, the incoming heat content of the flow in Scheme II is lower than that of Scheme I, and therefore, the steam which can be raised in Scheme II is less than in the Scheme I system. Since the waste heat available from the methanization effluent heat exchanger is fixed, the lower steam cycle flow means the heat in the methanization effluent must be divided between the water in the steam cycle and that in the heat-to-process heat exchanger. Characteristics of the heat exchangers incorporated in the Scheme II systems are presented in Table XIII.

The total estimated installed horsepower of the turbomachinery of Scheme II is 2,741,930 hp. This is divided as follows: 672,086 hp in Turbines C; 587,982 hp in Turbines D; and 1,481,862 hp from the Power Turbines E. This compares directly with a total of 2,049,330 hp installed in the Scheme I system. This latter total is divided among Turbines A: 278,846 hp; Turbines D: 586,434 hp; and the Power Turbines E: 1,184,050 hp. Of course, it should be realized that in the case of the turbine and compressor components, the compressor power absorbed is the same as that developed by the driving turbine.

Reheat Steam Cycle

A steam reheat cycle operating at conditions of 2400 psia/1000F/1000F could be installed in Scheme II. The output power of each of two identical systems would be 120,800 kw, and the steam flow rate would be slightly in excess of 1.412 million pounds per hour. Overall steam cycle efficiency was estimated at 37.72 percent based on extensive studies conducted at UTRC (Ref. 3). Because of the 1200 F inlet temperature, the minimum allowable steam-to-gas temperature difference of 50F occurs at the inlet to the boiler component (see Fig. 6). A total of 2230.1 million Btu/hr are transferred into the steam cycle of Scheme II, and of this total, slightly more than 2053 million Btu/hr are transferred in the boiler/superheater. The remaining load is transferred from the methanization effluent heat exchanger. It can be seen from Fig. 6 that because of the temperature-heat requirements of the steam cycle boiler/superheater, an exhaust temperature considerably higher than 300°F results. Whereas there still exists a large amount of thermal energy in the exhaust gases at the exit of the boiler/superheater, it was assumed that this heat would be exhausted directly to the atmosphere. Actually, the cost of recovering this heat may not be practical, because its low temperature would require a large and expensive heat exchanger. Furthermore, only the process water could act as a heat sink, and its requirements can be accommodated with the thermal energy from the methanization effluent heat exchanger.

Nonreheat Steam Cycle

Because the heat energy in the exhaust at the exit of the boiler/superheater of the Scheme II reheat cycle system is discharged to atmosphere, it was decided to examine a nonreheat steam system with a peak operating pressure of 1600 psi and a peak temperature of 1000°F (see Fig. 7). In this system, as in that of the reheat system, the minimum allowable temperature difference between exhaust gas and steam working fluid occurs at the inlet of the boiler section. However, because of the steam temperature profile, it is possible to reduce the exhaust gas temperature to 388°F, approximately 55 degrees below that of the reheat steam cycle just discussed. Note that on both Scheme II systems, the water from the condenser is preheated to 250°F in the methanization effluent heat exchanger. The thermal efficiency of the simple-cycle system is estimated at 33.32 percent, and a total net output, after allowing for 2 percent losses in the electric generator, is estimated to be 237 MW for the two identical steam units. The total steam flow is estimated at 1.775 million lb/hr or approximately 363 thousand lb/hr greater than that in the reheat steam system. Of the 2477 million Btu/hr transferred to the steam cycle working fluid, an excess of 2243 million Btu/hr are extracted from the exhaust gases; the remainder is transferred from the methanization effluent heat exchanger As in previous steam system analyses, basic performance analyses data were taken from Ref. 3.

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System Costs

The estimated turbomachinery manufacturing cost on a major component basis for Compressor A - Turbine C, Compressor B - Turbine D, and Power Turbine E components are shown in Tables XIV, XV and XVI respectively. Although the Compressor B components of Scheme II and that of Scheme I handle the same flows and operate between the same pressure ratio, it must be realized that compressor section costs cannot be considered in isolation. The difference in performances of Turbine D in System II and Turbine D in System I, is due to the differing match points in both work/lb/stage and the rotational speeds. Therefore, the $266,000 (Table XV) estimate for the inlet, disk, and airfoils of the Scheme II Compressor B can only casually be compared with the approximately $340,000 (Table IV) manufacturing cost for the same components of Compressor B in Scheme I. It should be noted that neither time nor funds of this subcontract allowed a true optimum system to be designed. However, when considered in relation to the overall system cost in which the gas turbine components are installed, the differences noted are of minor consequence; their values still are reasonable indications of the costs expected to be encountered. In Tables XIV, and XV and XVI, all allowances have been made for the proper material selections and airfoil cooling schemes necessary to achieve the performance levels indicated. A general, overall review of these major selections is presented in Table XVII.

In Table XVIII an estimate of the selling prices (the cost to the ultimate customer) are shown for all exhaust gas turbomachinery components plus the combustor and fuel control systems. The total price includes an estimated industry markup factor of 100 percent, and is the sum for all of the units installed in the Scheme-II system based on the production of at least 50 units. As is noted for the Scheme-I turbomachinery selections, the estimated costs (and prices) for the first two and the sixth and seventh units would be approximately 122 percent and 69 percent higher than the values indicated in Table XVIII. However, the fact that development costs write-offs would play an important part in establishing the level of price for total production runs of less than 50 units may lessen the reliability associated with the lower-production-rate price estimates as discussed earlier in this report.

The total installed cost of the reheat steam cycle system rated at 1325.4 MW is estimated to be in excess of $220 million including escalation and interest as shown in the center column of results in Table VIII. In the third column of results the total costs of the nonreheat steam system, including the same component selections is estimated to cost slightly in excess of $220 million for a net power output of 1320.8 MW. Tables IX through XII present the detailed cost estimates by component category for the Scheme II reheat cycle and nonreheat steam systems. Note that FPC Account Number 343, Table XII, relates strictly to the gas turbine system components, and as such columns two and three are applicable to both the reheat cycle steam system and the nonreheat steam systems of Scheme II. Changes in total system cost due to alterations to the escalation rate, interest rate, or to the production rate of gas turbines can be undertaken in a manner similar to that discussed earlier under Scheme I results.

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A seemingly contradictory set of economic data appear to be present when a comparison between the total prices of the gas turbine units in Schemes I and II is made (c.f., Tables V, XII and XVIII). The Scheme-II systems rated at a total of 1083.8 MW are estimated to sell for $26,026,684 (plus 5 percent installation), whereas the Scheme-I turbomachinery price estimate is $30,868,568 (plus 5 percent installation) for systems rated at a total of 865.3 MW. The reason for this difference is that gas turbines are volumetric flow machines, and since the largest part of the total installed power of the Scheme-I turbomachinery handles flow at a lower pressure than that of the Scheme-II turbomachinery (the total flow of both systems being nearly the same), the former units are physically larger and run somewhat slower than do those units in the Scheme-II systems. Price has been found to relate to machinery dimensions and to material selection so it can be seen that the larger rotating devices (as well as the larger, low-density burner) in the Scheme-I systems have higher associated costs/prices relative to those in the smaller machines of Scheme II.

A comparison of the estimated capital cost results in Table VIII indicates that the Scheme-II system incorporating the reheat steam system is only slightly more costly than is that system incorporating the nonreheat steam equipment. However, comparison of the specific costs reveals the advantage to be in favor of the system incorporating the reheat steam equipment. The higher specific cost of the nonreheat system can be attributed primarily to the steam turbine and generator equipment which are larger than the corresponding equipment in the reheat system with its thermal higher efficiency and lower steam flow per net kilowatt output. In direct contrast to this, the increased exhaust gas exit temperatures associated with the reheat cycle directly influence the heat which can be transferred to the steam cycle from the methanization heat exchanger, and therefore, the net remaining heat which can be delivered to the heat-to-process heat exchanger. Because of the higher cycle temperatures, a higher heat-to-process heat exchanger cost estimate is therefore associated with the reheat cycle. However, for purposes as a budgetary estimate such as that presented herein, the total installed package price of $166.7 per kw and $167.2 per kw for the Scheme-II systems can be assumed equal, and the ultimate choice of system likely would be a matter of other considerations taken into account during the overall system design.

A comparison of the capital and operating charges for all three systems considered in this analysis is presented in Table XIX for systems located at a Middletown USA site (an A.E.C. definition) and explained in more detail in Ref. 3. A summary of the annual owning and operating costs for these systems is shown at the bottom of this table. The capital charges assume an owning cost of 17 percent of the capital investment and a load factor of 70 percent values which are quite typical of those selected in present power plant analyses. Based on data from Ref. 3 the annual operation, supplies, and maintenance charges were assumed to be 3.5 percent of the capital cost of the plant, a conservative estimate, yet one which is nearly double that presently assumed for modern steam plant designs. Finally, the charge for fuel to the power station was assumed to be zero since the installation of a power system to the basic steam-iron coal gasification process is an added benefit whose real value comes in the sale of power at a price exceeding the total busbar power cost. The charge for the fuel is made only once, and that occurs in the basic conversion process preceding the power systems considered in this analysis. The total estimated cost of busbar power costs of less than 7 mills/kw hr for the three systems are quite attractive, particularly when considered relative to the estimated power generating costs being made today for similar advanced systems which are between two and three times greater, even for the most attractive of advanced systems (Ref. 3). Of course, the great factor contributing to the low cost of busbar power is the lack of a fuel charge. The economic results shown in this table should not be considered in isolation since they only explain a portion of the entire story. Only when these charges, or more specifically, the profits estimated to accrue from the sale of energy generated at these costs, plus the return from salable products from the basic coal gasification process comprising the primary portion of this system, are examined and compared with competing systems will the ultimate economic advantages be identified.

CONCLUDING REMARKS

Three separate power systems are shown which have characteristics and operating parameters which are extremely attractive relative even tc the best of modern systems being considered today. Although the results presented may not be representative of the optimum combination of machinery or operating conditions, they are, however, sufficiently representative of viable systems such that improvements should result in even more attractive designs. Whereas the results are presented in terms of mid-1974 dollar values, analyses in terms of current dollars can be made by applying appropriate inflator factors from reliable, published industry sources. No recommendations as to the type of system which is appropriate since it is believed to be a matter of individual choice depending possibly on the availability of critical components at the time of the actual system design. Proof of the entire concept, of course, will come at the time when the results presented herein are combined with technical and economic data for the gasification system, and providing these appear attractive, the ultimate construction of demonstration systems in the field. It can be concluded, however, that by incorporating a system whose primary purpose is to generate a salable gaseous fuel with an electric power generating system, an important benefit of meeting a portion of this nation's future electric power needs at costs which are reasonable and within reach of many electric power systems should result.

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TABLE I

BASIC INPUT DATA SUPPLIED BY IGT FOR COMBINED-CYCLE PERFORMANCE AND COST ANALYSES

Producer Gas Availability

Flow Rate - 354,989.4 moles/hr
Temperature - 1520 F
Pressure - 365 psia
Composition (Mole Percent)
CO - 8.76
CO_2 - 20.71
H_2 - 6.21
H_2O - 16.00
CH_4 - 0.38
H_2S - 0.12
N_2 - 47.82

Process Air Requirement

Flow Rate - 217280 moles/hr
Temperature - 1000F
Pressure - 400 psia

Process Heat Requirement (Steam)

Flow Rate - 2,162,440 lb/hr
Temperature - 445F
Pressure - 400 psia
Energy Content - 2170 x 10^6 Btu/hr

Waste Heat Availability (Methane Effluent Heat)

Temperature - Approximately 340F
Energy Content - 356.7 x 10^6 Btu/hr

TABLE II

CHARACTERISTICS OF HEAT EXCHANGERS IN SCHEME I

Heat-To-Process Heat Exchanger

Unit	Heat Exchanged (millions of Btu/Hr)	Surface Area	Total Price
Boiler	1694.5 (Q_B)*	676,500	$5,412,300
Economizer	475.5 (Q_E)*	450,100	3,646,000
Total	2170.0	1,126,600	$9,058,300

Steam Cycle Heat Exchanger

Unit	Heat Exchanger (millions of Btu/Hr)	Surface Area	Total Price
Methane Effluent	356.7	369,900	$2,996,200
Economizer	1,281.5	2,969,600	24,053,800
Boiler	813.3 } (Q_S)*	559,500	4,532,200
Superheater	929.5	469,400	3,802,300
Reheater	496.4 (Q_R)*	144,400	1,169,800
Total	3,877.4	4,512,800	$36,554,300

*Refers to Fig. 4

TABLE III

MANUFACTURING COSTS OF COMPRESSOR C - TURBINE A COMPONENTS

Scheme I
Costs Shown are for One Unit Only

Inlet Section	$ 12,202
Compressor (11 Stages)	
Blades	24,650
Vanes	39,698
Disks and Front Hub	89,562
Turbine (1 Stage)	
Blades	11,608
Vanes	6,353
Disks and Hubs	106,830
Casing	243,283
Bearings	14,335
Miscellaneous Parts	206,095
Total Assembled Engine Manufacturing Cost	$754,616

Table IV

MANUFACTURING COSTS OF COMPRESSOR B - TURBINE D COMPONENTS

Scheme I
Costs Shown are for One Unit Only

Inlet Section	$ 10,639
Low Compressor (9 Stages)	
Blades	30,165
Vanes	59,211
Disks and Front Hub	58,677
High Compressor (11 Stages)	
Blades	47,524
Vanes	77,107
Disks and Rear Hub	57,453
High Turbine (1 Stage)	
Blades	27,694
Vanes	18,120
Disks and Hubs	81,661
Low Turbine (2 Stages)	
Blades	495,380
Vanes	126,065
Disks and Hubs	126,950
Casing	237,547
Shafts	16,762
Bearings	19,988
Miscellaneous Parts	403,240
Total Assembled Engine Manufacturing Cost	$1,894,183

TABLE V

MANUFACTURING COST OF POWER TURBINE E COST COMPONENTS

Scheme I
Costs Shown are for One Unit Only

Turbine (3 Stages)	
Blades	$487,824
Vanes	595,992
Disks, Hubs, and Tie Rods	175,832
Casings	237,036
Bearings	14,760
Miscellaneous Parts	397,846
Total Assembled Engine Manufacturing Cost	$1,909,290

TABLE VI

MAJOR FEATURES OF TURBOMACHINERY UNITS OF SCHEME I

Compressor C - Turbine A

Compressor Section

Stage	1	2	3	4	5	6
No. of Blades	30	35	41	51	70	112
No. of Vanes	32	37	44	54	74	118

Blade and Vane Material - AMS 5616
Disk Material - Cameron Z 448

Turbine Section

No. of Blades - 148 (uncooled)
No. of Vanes - 81 (uncooled)

Blade and Vane Material - IN 713
Disk Material - Cameron Z 448

Compressor B - Turbine D

Compressor Section

Stage	1	2	3	4	5	6	7	8	9	10
No. of Blades	33	35	39	43	48	55	64	77	99	75
No. of Vanes	35	37	41	46	51	58	68	81	104	79

Stage	11	12	13	14	15	16	17	18	19	20
No. of Blades	76	78	81	83	86	90	95	100	107	117
No. of Vanes	80	82	86	88	91	95	100	105	113	123

Blade and Vane Materials - AMS 5616 and IN718
Disk Materials - Cameron Z 448 and AMS 5616

TABLE VI (continued)

High Turbine Section

No. of Blades - 63 (cooled)
No. of Vanes - 29 (cooled)

Blade and Vane Material - B1900
Disk Material - AMS 5719

Low Turbine Section

Stage	1	2	
No. of Blades	110	106	(First Stage Cooled)
No. of Vanes	79	53	(First Stage Cooled)

Blade and Vane Material - Stellite 31, B1900
Disk Material - AMS 5616

Power Turbine

Stage	1	2	3	
No. of Blades	156	110	74	(First 2 stages cooled)
No. of Vanes	124	87	59	(First 2 stages cooled)

Blade and Vane material - Stellite 31
Disk Materials - AMS 5719, AMS 5616

TABLE VII

SELLING PRICES OF TURBOMACHINERY UNITS IN SCHEME I

Compressor C - Turbine A (2 units)	$3,018,470
Compressor B - Turbine D (2 units)	7,576,720
Power Turbine (2 double-ended units)	15,274,300
Combustor (2 units)	3,999,080
Fuel Control (2 units; estimated price)	1,000,000
Total Assembled Selling Price	$30,868,570

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TABLE VIII

CAPITAL COST SUMMARY FOR COMBINED CYCLE POWER GENERATING SYSTEMS

System		Scheme I	Scheme II	Scheme III
Steam Cycle		Reheat	Reheat	Nonreheat
Gas Turbine Output (mw)		865.3	1,083.8	1,083.8
Steam Turbine Output (mw)		420.1	241.6	237.0
Total System Output (mw)		1,285.4	1,325.4	1,320.8
Federal Power Commission Account Number			Capital Cost (Thousands of 1974 Dollars)	
Boiler Plant Eq.	312	54,179	43,502	42,561
Steam Turbine Gen.	314	27,335	18,904	20,079
Struct. & Improv'ts.	341	12,598	11,450	11,236
Prime Mover (Gas Turb.)	343	42,212	37,409	37,409
Generator (Gas Turb.)	344	9,887	12,129	12,129
Acc. Elect. Eq.	345	15,388	15,825	15,767
Misc. Power Plant Eq.	346	670	680	679
Station Eq.	353	In Acct 345	In Acct 345	In Acct 345
Other Expenses		3,245	2,798	2,797
Total Direct Cost		165,514	142,697	142,657
Engineering and Supervision		24,825	21,405	21,399
Contingency		13,240	11,416	11,413
Escalation		9,203	7,742	7,678
Interest During Construction		43,785	37,713	37,690
Total Installed Power Station Cost		256,567	220,972	220,837
Capital Cost per kw ($)		199.6	166.7	167.2

TABLE IX

BREAKDOWN OF FPC ACCOUNT 312 - BOILER PLANT EQUIPMENT

Total for Two Identical Systems

System Steam Cycle	Scheme I Reheat	Scheme II Reheat	Scheme II Nonreheat
Gas Turbine Output Power (MW)	865.3	1083.8	1083.8
Steam Turbine Output Power (MW)	420.1	241.6	237.0
Total System Output Power (MW)	1285.4	1325.4	1320.8
Waste Heat Boiler	$36,354,300	$20,486,800	$20,691,900
Boiler Feed Pumps	725,000	406,600	383,500
Boiler Feed Tank Deaerator	134,000	75,500	94,900
Water Treatment-Demineralization	696,800	392,700	493,600
Condensate Storage Tank	35,100	19,300	26,800
Process Steam Heat Exchanger	9,058,300	18,073,700	16,874,600
Miscellaneous Pumps	105,650	84,500	108,400
Piping	5,839,400	3,358,200	3,294,300
Insulation for Piping	467,200	268,700	263,500
Controls Computer	584,000	335,800	329,400
Total: Account 312	$54,174,750	$43,501,800	$42,560,900

TABLE X

BREAKDOWN OF FPC ACCOUNT 314 - STEAM TURBINE GENERATOR

Total for Two Identical Systems

System Steam Cycle	Scheme I Reheat	Scheme II Reheat	Scheme II Nonreheat
Gas Turbine Output Power (MW)	865.3	1083.8	1083.8
Steam Turbine Output Power (MW)	420.1	241.6	237.0
Total System Output Power (MW)	1285.4	1325.4	1320.8
Steam Turbine and Generator	$17,170,800	$11,942,500	$11,435,000
Condenser and Tubes	1,220,600	688,000	864,600
Condensate Vacuume Pump and Motor	135,700	76,500	96,100
Condensate Pump and Motor	153,200	86,400	108,500
Cooling Tower	7,946,300	5,633,400	6,461,100
Circulating Water Pump	709,100	476,700	1,114,000
Make-up Structure; Screens and Pumps			
Chlorination Equipment			
Miscellaneous Pumps			
Total: Account 314	$27,335,700	$18,903,500	$20,079,300

TABLE XI

BREAKDOWN OF FPC ACCOUNT 341 - STRUCTURES AND IMPROVEMENTS

Total for Two Identical Systems

System Steam System	Scheme I Reheat	Scheme II Reheat	Scheme II Nonreheat
Gas Turbine Output Power (MW)	865.3	1083.8	1083.8
Steam Turbine Output Power (MW)	420.1	241.6	237.0
Total System Output Power (MW)	1285.4	1325.4	1320.8
Site Preparation	$1,245,700	$1,272,700	$1,269,600
Administration Building	828,000	846,000	843,600
Condensate Polishing System	1,544,600	1,047,100	839,100
Turbogenerator Building	7,557,000	6,720,000	6,720,000
Stack	452,200	535,500	535,500
Tank Farm	970,900	1,028,400	1,028,400
Total, Account 341	$12,598,400	$11,449,700	$11,236,200

TABLE XII

BREAKDOWN OF FPC ACCOUNT 343 - PRIME MOVERS (GAS TURBINES)

Total for Two Identical Systems

System	Scheme I	Scheme II	Scheme II
Steam Cycle	Reheat	Reheat	Nonreheat
Gas Turbine Output Power (MW)	865.3	1083.8	1083.8
Steam Turbine Output Power (MW)	420.1	241.6	237.0
Total System Output Power (MW)	1285.4	1325.4	1320.8
Gas Turbines (installed)	$32,412,000	$27,328,000	$27,328,000
Starting Motors	64,200	64,200	64,200
Torque Convertors	201,000	201,000	201,000
Lube Oil Purfication System	201,600	201,600	201,600
Lube Oil Fire Protection	168,000	168,000	168,000
Turbine Airfoil Cooling System	672,500	637,600	637,600
Compressor Service and Instrumentation	140,000	140,000	140,000
Breeching:			
Intake Silencers and Insulation	798,050	532,950	532,950
Enclosure	786,300	692,200	692,200
Exhaust Hoods	61,000	70,400	70,400
Ductwork	903,700	1,010,100	1,010,100
Accoustic Insulation	123,420	137,900	137,900
Inlet Air Filters	386,800	464,800	464,800
Turbine Enclosure Aircooler	168,000	168,000	168,000
Emergency Cooling Water Tank, etc.	11,200	11,200	11,200
Fuel Oil Heaters and Pumps	15,400	15,400	15,400
Miscellaneous Pumps and Tanks	42,000	42,000	42,000
Control Panels	140,000	140,000	140,000
Computer Controls	560,000	560,000	560,000
Fuel Piping	1,630,600	1,630,600	1,630,600
Fuel Pipe Insulation	244,700	244,700	244,700
Airfoil Cooling Compressors	2,482,200	2,948,800	2,948,800
Total: Account 343	$42,212,670	$37,409,350	$37,409,350

TABLE XIII

CHARACTERISTICS OF HEAT EXCHANGERS IN SCHEME II

Unit	Heat Exchanger (Million of Btu/hr)	Surface Area	Total Price
Preheater (Methane Effluent)	180.5	629,500	$ 5,099,100
Economizer	301.7	814,900	6,600,500
Boiler	1687.8	786,900	6,374,100
TOTAL	2170.0	2,231,300	$18,073,700
Heat-to-Process Heat Exchanger - Nonreheat Steam System			
Preheater (Methane Effluent)	123.4	292,600	2,370,400
Economizer	358.8	976,300	7,908,200
Boiler	1687.8	814,300	6,596,000
TOTAL	2170.0	2,083,200	$16,874,600
Steam Cycle Heat Exchanger - Reheat Steam System			
Methane Effluent	176.3	178,900	$ 1,449,500
Economizer	747.6	984,000	7,970,800
Boiler	474.4	521,300	4,222,200
Superheater	540.2	674,200	5,461,000
Reheater	291.6	170,800	1,383,300
TOTAL	2230.1	2,529,200	$20,486,800
Steam Cycle Heat Exchanger - Nonreheat Steam System			
Methane Effluent	233.3	238,600	$ 1,932,400
Economizer	740.3	1,201,600	9,733,300
Boiler	1002.8	781,900	6,333,400
Superheater	500.3	332,400	2,692,800
TOTAL	2476.7	2,554,500	$20,691,900

TABLE XIV

MANUFACTURING COSTS OF COMPRESSOR A - TURBINE C COMPONENTS

Scheme II

Costs Shown Are For One Unit Only

Inlet Section	$ 15,635
Low Compressor (13 Stages)	
Blades	75,396
Vanes	100,007
Disks and Front Hub	134,067
High Compressor (11 Stages)	
Blades	43,686
Vanes	81,935
Disks and Rear Hub	69,617
High Turbine (2 Stages)	
Blades	67,276
Vanes	54,383
Disks and Hubs	46,862
Low Turbine (2 Stages)	
Blades	143,701
Vanes	42,380
Disks and Hubs	59,559
Casings	180,504
Shafts	17,684
Miscellaneous Parts	333,716
Total Assembled Engine Manufacturing Cost	$1,492,309

TABLE XV

MANUFACTURING COSTS OF COMPRESSOR B-TURBINE D COMPONENTS

Scheme II

Costs Shown Are For One Unit Only

Inlet Section	$ 10,998
Low Compressor (8 Stages)	
Blades	29,007
Vanes	50,768
Disks and Front Hub	71,725
High Compressor (7 Stages)	
Blades	25,632
Vanes	46,662
Disks and Rear Hub	31,906
High Turbine (2 Stages)	
Blades	42,725
Vanes	53,360
Disks and Hubs	35,279
Low Turbine (2 Stages)	
Blades	64,024
Vanes	16,527
Disks and Hubs	62,458
Casing	162,389
Shafts	9,406
Bearings	18,866
Miscellaneous Parts	244,421
Total Assembled Engine Manufacturing Cost	$ 976,153

TABLE XVI

MANUFACTURING COSTS OF POWER TURBINE E COMPONENTS

Scheme II

Costs Shown Are For One Unit Only

Turbine (4 Steps)	
Blades	$ 621,225
Vanes	333,407
Disks, Hubs, and Tie Rods	126,140
Casings	125,767
Bearings	10,728
Miscellaneous Parts	340,439
Total Assembled Engine Manufacturing Cost	$1,557,706

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TABLE XVII

MAJOR FEATURES OF TURBOMACHINERY UNITS OF SCHEME II

Compressor A - Turbine C

Compressor Section

Stage	1	2	3	4	5	6	7	7
No. of Blades	23	24	26	28	30	32	35	38
No. of Vanes	25	26	28	30	32	34	37	40

Stage	9	10	11	12	13	14	15	16
No. of Blades	42	46	52	59	68	47	50	53
No. of Vanes	45	49	55	62	72	50	53	56

Stage	17	18	19	20	21	22	23	24
No. of Blades	56	61	66	72	80	89	103	121
No. of Vanes	59	65	70	76	84	94	109	128

Blade and Vane Material - AMS 5616 and IN718
Disk Material - Cameron Z 448 and AMS 5616

High Turbine Section

Stage	1	2	
No. of Blades	112	82	(Both Stages Cooled)
No. of Vanes	50	52	(Both Stages Cooled)

Blade and Vane Material - B1900
Disk Material - AMS 5719

Low Turbine Section

Stage	1	2	
No. of Blades	132	95	(Uncooled)
No. of Vanes	37	28	(First Stage Cooled)

Blade and Vane Material - U700
Bisk Material - AMS 5616

TABLE XVII - Continued

Compressor B - Turbine D

Compressor Section

Stage	1	2	3	4	5	6	7	8
No. of Blades	28	32	36	41	48	58	73	99
No. of Vanes	30	34	38	44	51	61	77	104

Stage	9	10	11	12	13	14	15
No. of Blades	45	49	55	62	72	87	113
No. of Vanes	48	52	58	66	76	92	119

Blade and Vane Material - AMS 5616
Disk Material - Cameron Z 448

High Turbine Section

Stage	1	2	
No. of Blades	64	55	(Both stages cooled)
No. of Vanes	49	29	(Both stages cooled)

Low Turbine Section

Stage	1	2	
No. of Blades	134	110	(Uncooled)
No. of Vanes	49	40	(Uncooled)

Blade and Vane Material - B1900
Disk Material - AMS 5616

Power Turbine E

Stage	1	2	3	4	
No. of Blades	171	151	149	102	(First two stages cooled)
No. of Vanes	56	54	57	36	(First two stages cooled)

Blade and Vane Materials - Satellite 31; IN713, B1900
Disk Materials - AMS 5719 and AMS 5616

TABLE XVIII

SELLING PRICE OF TURBOMACHINERY IN SCHEME II

Compressor A - Turbine C (2 Units)	$ 5,969,236
Compressor B - Turbine D (2 Units)	3,904,608
Power Turbine - (2 Double Ended Units)	12,461,640
Combustor - (2 Units)	2,691,200
Fuel Control (2 Units; Estimated Price)	1,000,000
Total Assembled Selling Price	$26,026,684

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TABLE XIX

COST SUMMARY FOR POWER GENERATING SYSTEMS

	Scheme I	Scheme II Reheat Steam	Scheme II Nonreheat
Net Power Station Output, mw	1,285.4	1325.4	1,320.8
Capital Investment Thousands of $	256,567	220,971	220,837
Net Capital Cost, $/kw	199.6	166.7	167.2
Annual Owning and Operating Cost, mills/kwhr			
Capital Charges(1)	5.53	4.62	4.64
Operation, Supplies and Maintenance	1.14	0.95	0.96
Fuel(2)	0	0	0
Busbar Power Cost, mills/kwhr	6.67	5.57	5.60

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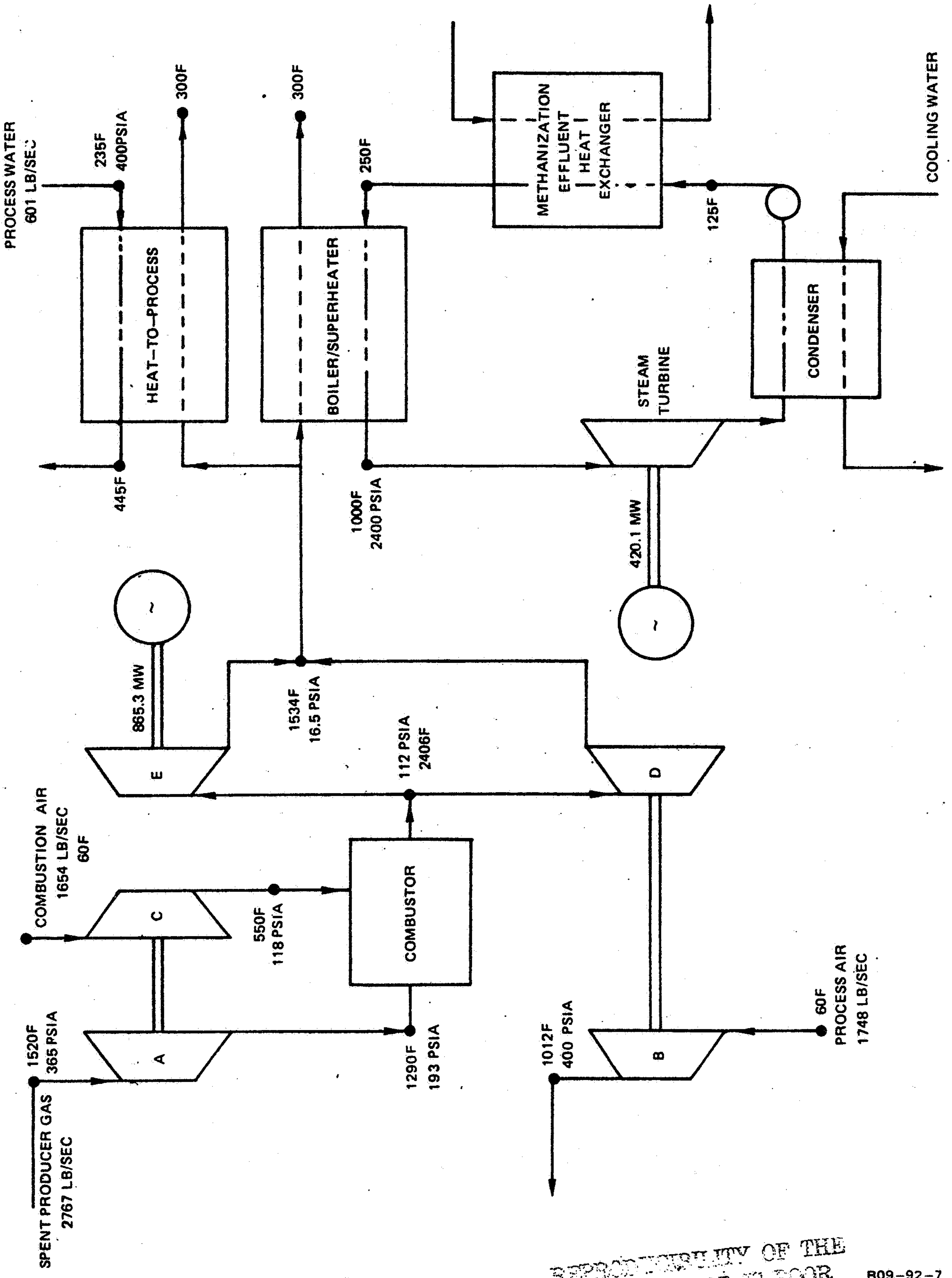
(1) Capital charges at 17% and a 70% load factor

(2) The cost of coal fuel charged against the gasification system

RECOVERY OF ELECTRICAL POWER FROM STEAM–IRON GASIFICATION PROCESS

SCHEME I



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RECOVERY OF ELECTRICAL POWER FROM STEAM–IRON GASIFICATION PROCESS

SCHEME II–REHEAT STEAM CYCLE

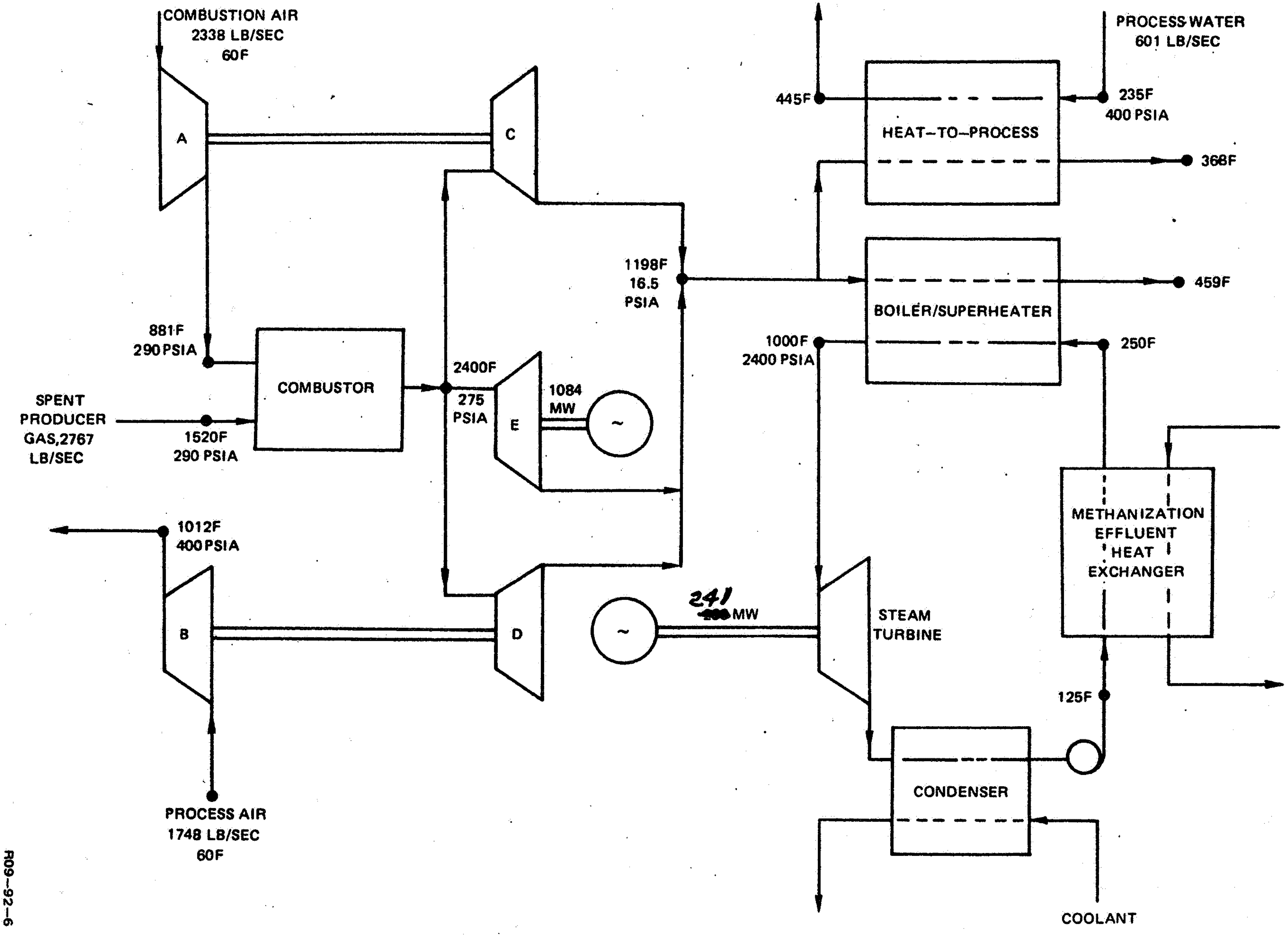

FIG.2

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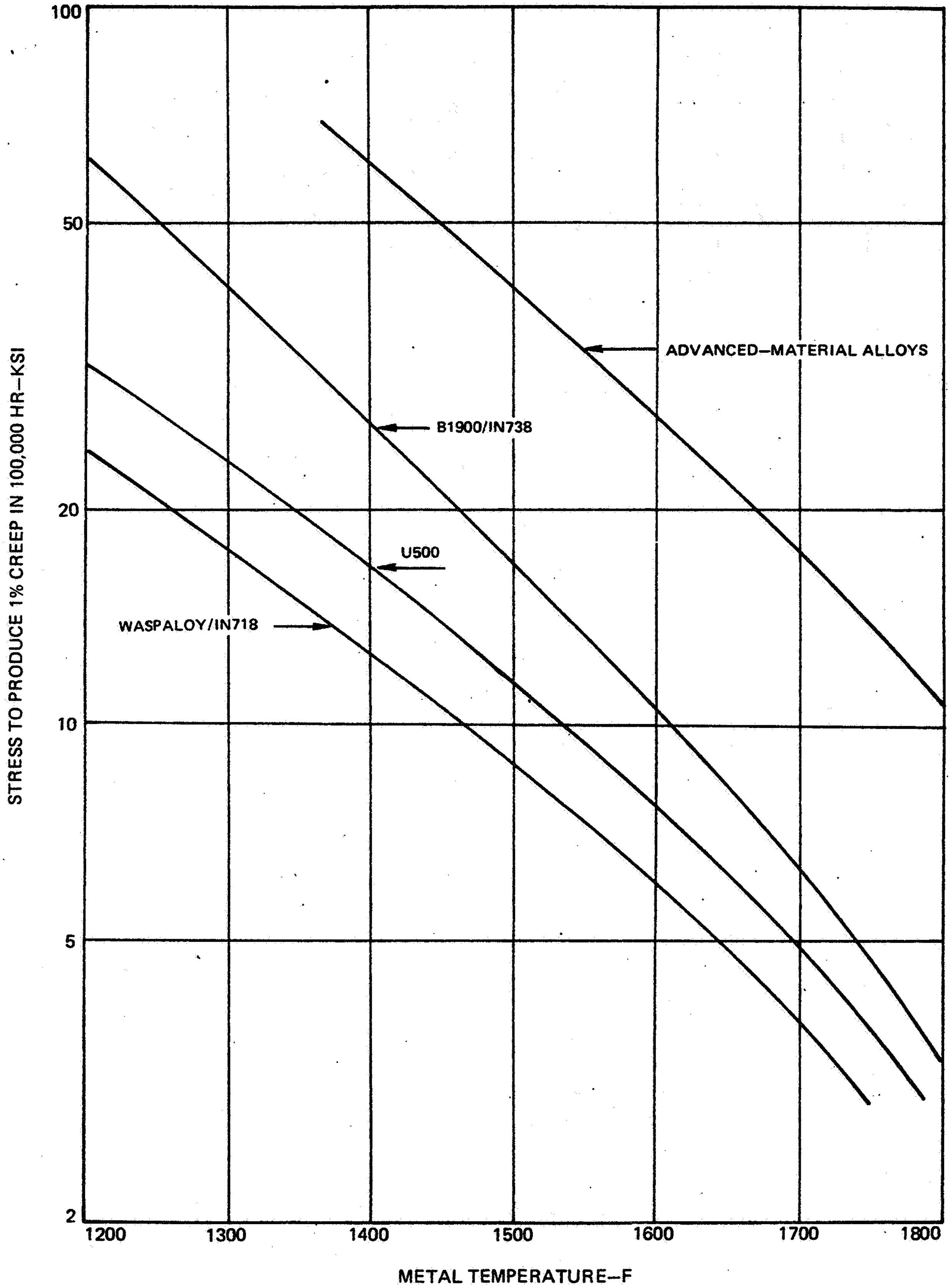
FIG. 3

CREEP STRENGTH FOR TURBINE ALLOY MATERIALS

SPECIFIC MATERIALS NOTED ARE REPRESENTIVE OF THEIR RESPECTIVE CLASSES



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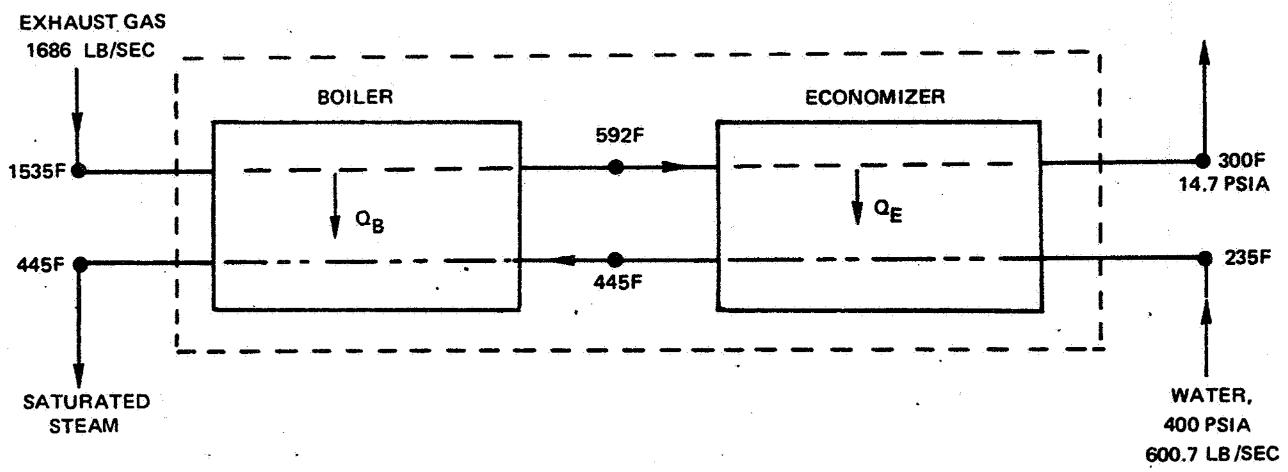
SCHEMATIC DIAGRAMS OF SCHEME–I HEAT EXCHANGERS

HEAT–TO–PROCESS HEAT EXCHANGER



STEAM CYCLE HEAT EXCHANGER

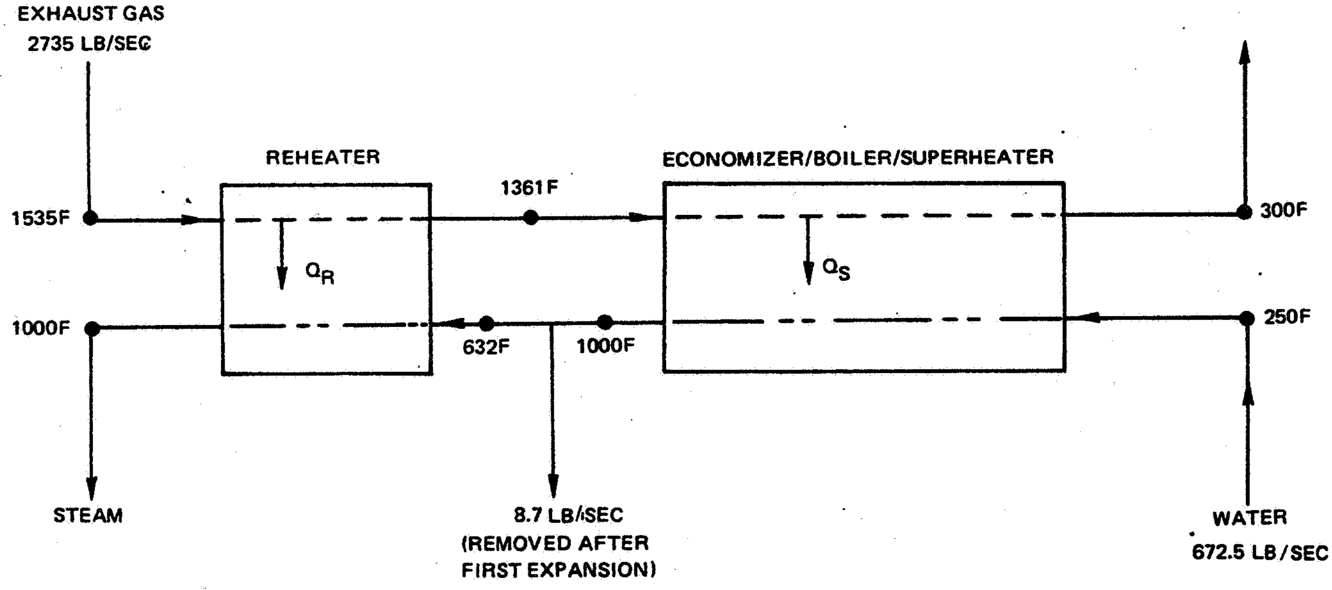


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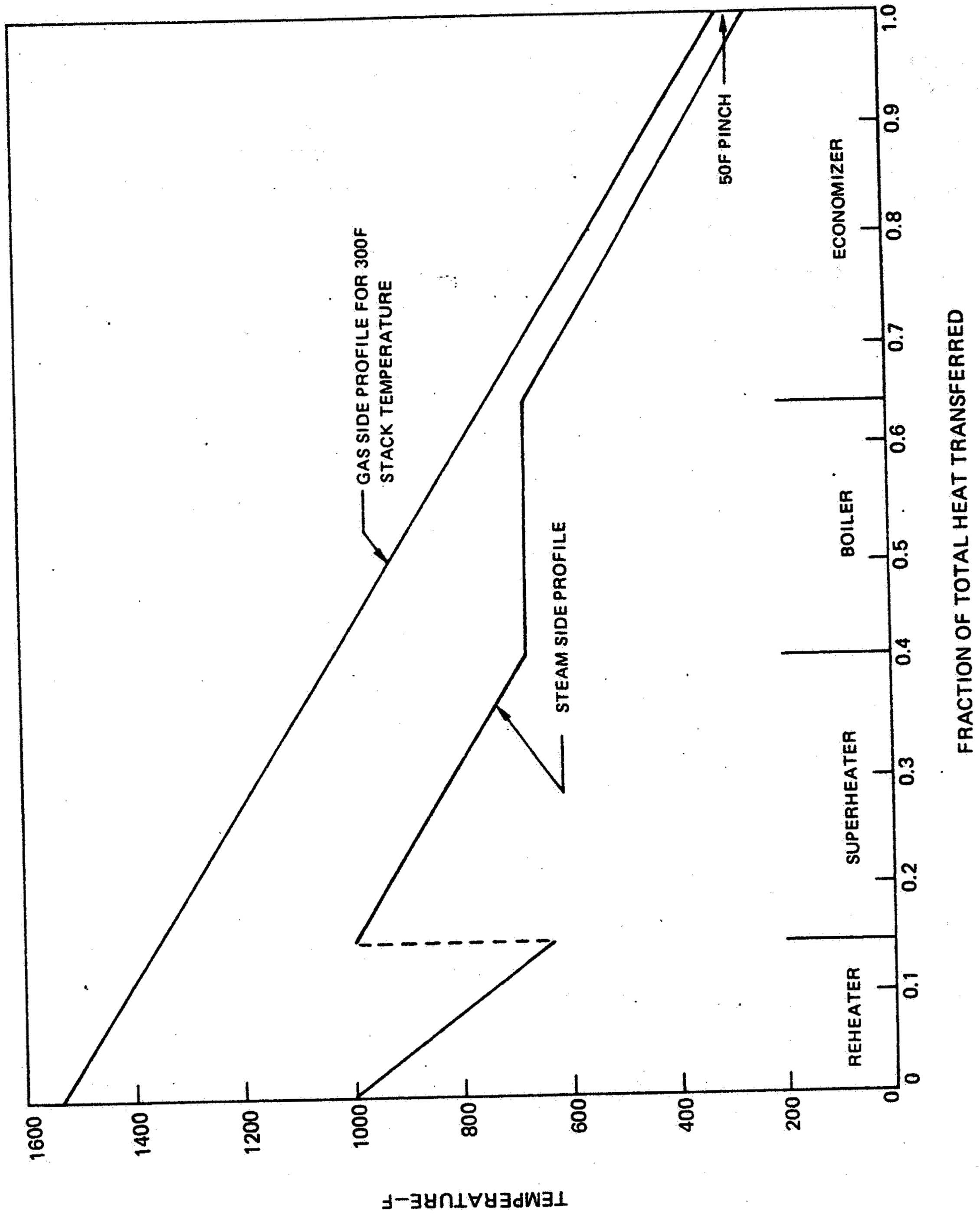
FIG.5

WASTE HEAT BOILER T–Q DIAGRAM

SCHEME I

REHEAT STEAM CYCLE: 2400 PSIA/1000F/1000F



WASTE HEAT BOILER T–Q DIAGRAM

IGT SCHEME II

2400 PSIA/1000F/1000F REHEAT STEAM CYCLE

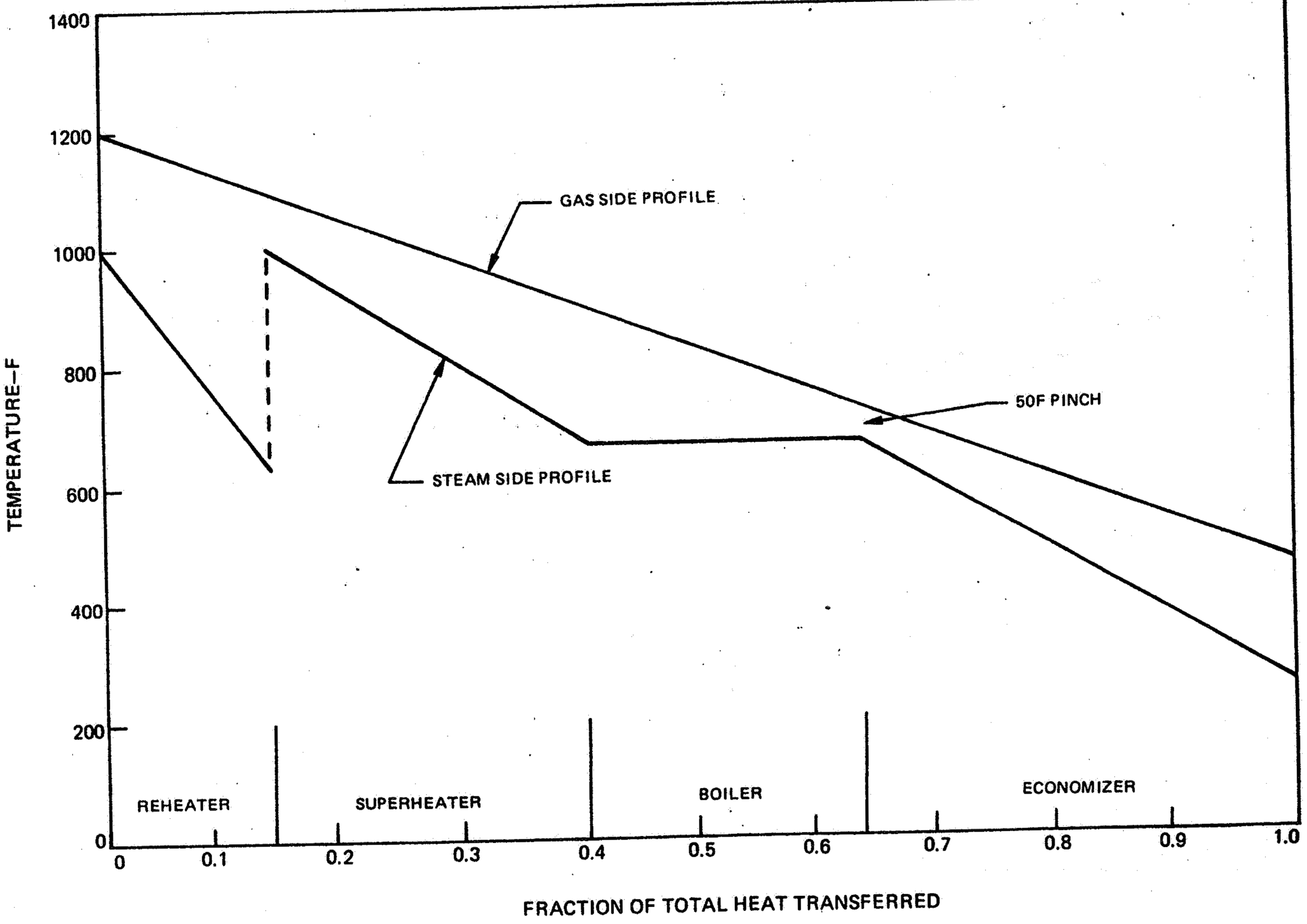


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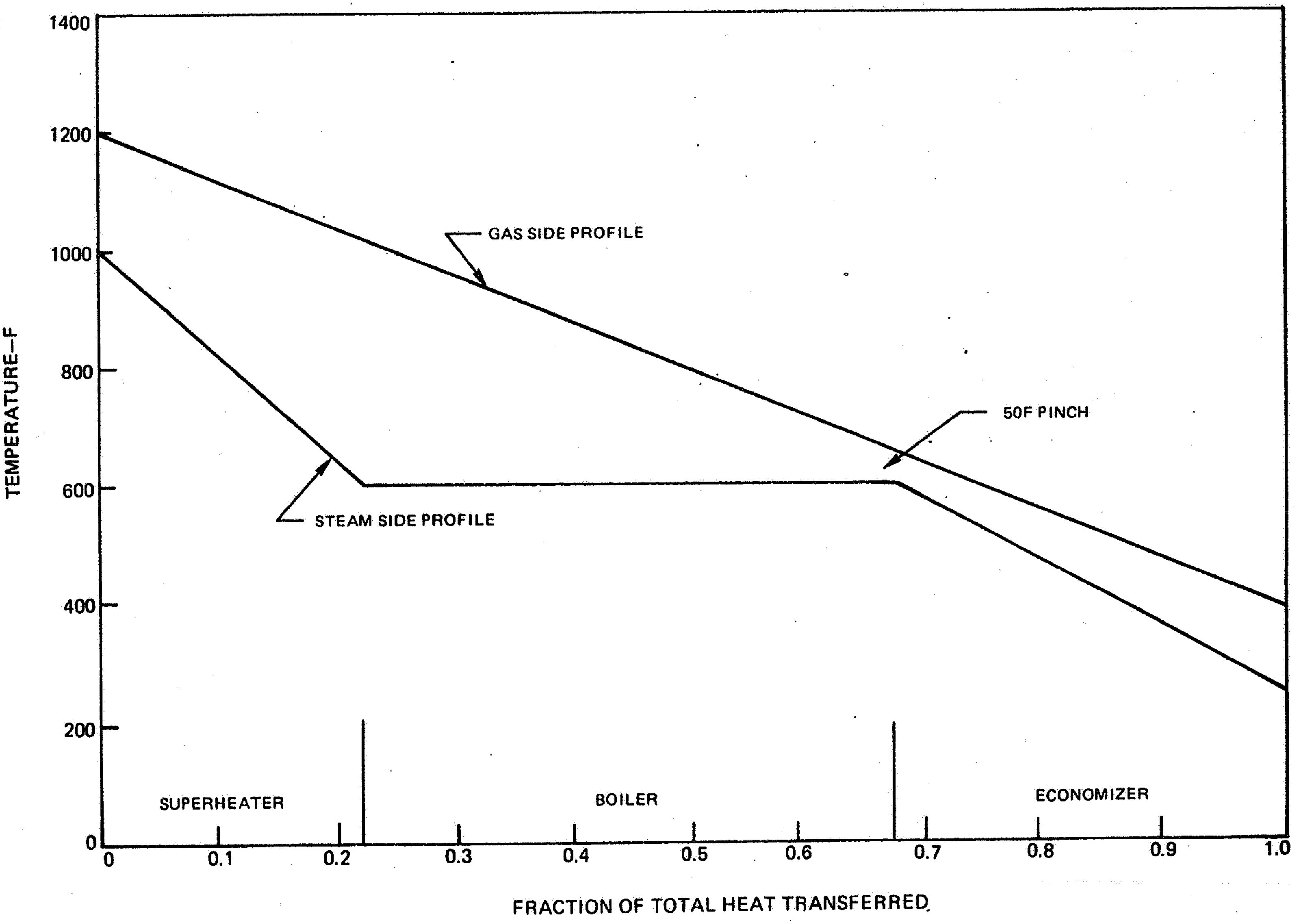
FIG.7

WASTE HEAT BOILER T–Q DIAGRAM

IGT SCHEME II

1600 PSIA/1000F NONREHEAT STEAM CYCLE



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