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IGT Project No. 65060

SYSTEMS ANALYSIS OF ELECTRICITY PRODUCTION FROM COAL USING FUEL CELLS



Final Report

Donald K. Fleming

November 1983

Prepared for

**CALIFORNIA INSTITUTE OF TECHNOLOGY
JET PROPULSION LABORATORY
4800 Oak Grove Drive
Pasadena, CA 91103**

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Prepared by
Institute of Gas Technology
IIT Center, 3424 S. State Street
Chicago, Illinois 60616



This work was performed for the Jet Propulsion Laboratory, California Institute of Technology, sponsored by the United States Department of Energy through an agreement with the National Aeronautics and Space Administration.

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EXECUTIVE SUMMARY

A program has been undertaken to evaluate alternative systems for the production of electricity from coal using coal gasification and power generation through molten carbonate fuel cells. The purpose of the work effort was to identify research and development effort that, if successful, could be cost-effective in reducing the overall cost of power from such systems.

The approach taken in the evaluation was the development of a versatile, computerized process simulator. The performance of individual subprocess elements was characterized mathematically; the subprocess elements were arranged in alternative configurations on the simulator to form a total process; and the relative performance of the systems was analyzed. For those systems showing efficiency promise, capital and operating costs for a 675 MW power plant were estimated using factored costing techniques. Finally, the 10-year levelized cost of power was obtained using standardized financial factors.

The above approach is cost-effective for the stated purpose of this program. However, it must be used with care and the results must be interpreted with mature engineering judgment for the following reasons:

1. By its nature, the overall system is quite sensitive to minor variations in inputted assumptions.
 - a. The systems analyst must use care to assure that only the desired variable has been altered in the analysis.
 - b. In a separate subtask, input assumptions were systematically varied for the same total base system. Process efficiencies varied from 46% to 57%; yet all of the systems are operable.
 - c. Comparisons of systems analyzed by different investigators must be made with extreme care.
2. The cost of electricity developed by this approach is highly sensitive to capital-related charges, with system efficiency having little impact. Yet, the factored-costing technique that was necessarily utilized cannot have great accuracy.
3. The financial assumptions used in developing the levelized cost of electricity also tend to emphasize capital costs, rather than system efficiency.

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A total of 66 cases were evaluated during the course of the study; several of these cases were iterated several times to assure greater comparability when the degree of sensitivity of the overall system became apparent.

Table ES-1 presents the nine systems that were analyzed economically; several other systems also incorporate features that warrant R&D attention. The efficiency and cost data in Table ES-1 must be assessed while considering the caveats above. In particular, the degree of conservatism used in this analysis is somewhat greater than other efforts reported in the literature — reported system efficiencies are considerably lower than estimated by other investigators.

The major conclusions drawn from this effort include

- Fluidized-bed gasification appears to be a viable alternative in this system if the relatively minor cost advantage can be realized. Although this approach has a total capital requirement that is only 2.5% lower than entrained flow gasification, this difference is apparently sufficient to counteract the three-percentage-point reduction in process efficiency.
- The development of steam reforming of methane, thermally and chemically integrated within the anode compartment of the fuel cell, offers significant efficiency and cost advantage when the system also includes a fluidized-bed gasifier. This advantage increases as the quantity of methane made in the gasifier is increased.
- High-temperature dust removal adds nearly two percentage points of process efficiency, without a capital cost penalty, by permitting process rearrangement. Additionally, this approach permits incorporation of other advanced concepts.
- High-temperature desulfurization, when integrated with high-temperature dust removal, results in still higher efficiency and cost savings due to improved system simplicity.

Other R&D effort was identified that may be cost-effective in reducing the cost of power or improving system simplicity:

- Improved definition of the carbon deposition phenomena, so that the quench system might be eliminated
- Development of countercurrent heat exchange for high-temperature, toxic, hazardous gases
- Development of a novel, single-stage, fluidized-bed gasifier that produces greater quantities of methane

Table ES-1. SUMMARY OF SYSTEMS ANALYZED

System No.	1	2	3	4	5
Gasifier Type	Entrained	Entrained	Fluidized Bed	Fluidized Bed	Fluidized Bed
Methane Make	Nil	Nil	Minimal	Minimal	Moderate
Heat Exchange Type	Countercurrent	Crossflow	Crossflow	Crossflow	Crossflow
Carbon Control	None	Quench	Quench	Quench	Quench
Dust Control	Water Wash	Water Wash	Water Wash	Water Wash	Water Wash
Sulfur Control	Selective AGR	Selective AGR	Selective AGR	Selective AGR	Selective AGR
CO ₂ Control	None	None	None	None	None
Fuel Cell	Conventional	Conventional	Conventional	Internal Reforming	Internal Reforming
System Efficiency, %	42.1	41.6	38.4	44.9	45.8
Total Capital Requirement (675 MW), \$10 ⁶	1024.1	1055.3	1029.1	981.6	970.4
10-Year Levelized Cost of Power, ¢/kWh					
Capital Charge	4.80	4.94	4.82	4.60	4.54
Coal	1.98	2.00	2.17	1.85	1.82
Other Operating	0.98	1.04	0.99	0.94	0.92
Fuel Cell	<u>0.72</u>	<u>0.76</u>	<u>0.75</u>	<u>0.87</u>	<u>0.88</u>
Total	8.5-	8.7+	8.7+	8.3-	8.2-

System No.	6	7	8	9
Gasifier Type	Fluidized Bed	Fluidized Bed	Fluidized Bed	Fluidized Bed
Methane Make	Moderate	Moderate	Moderate	Moderate
Heat Exchange Type	Crossflow	Crossflow	Crossflow	Crossflow
Carbon Control	Quench	Quench	Quench	Quench
Dust Control	High-Temperature	High-Temperature	Water Wash	High-Temperature
Sulfur Control	Selective AGR	None	Selective AGR	High-Temperature
CO ₂ Control	None	None	High-Temperature	None
Fuel Cell	Internal Reforming	Sulfur Tolerant Internal Reforming	Internal Reforming	Internal Reforming
System Efficiency, %	47.4	50.0	47.5	50.0
Total Capital Requirement (675 MW), \$10 ⁶	958.4	852.0	987.2	846.8
10-Year Levelized Cost of Power, ¢/kWh				
Capital Charge	4.49	3.99	4.62	3.97
Coal	1.76	1.67	1.76	1.67
Other Operating	0.91	0.78	0.94	0.80
Fuel Cell	<u>0.86</u>	<u>0.81</u>	<u>0.86</u>	<u>0.81</u>
Total	8.0+	7.2+	8.2-	7.2+

- Development of a dust-tolerant water-gas shift reactor, permitting the recovery of the exothermic heat of the shift reaction at elevated temperatures.

Additional R&D effort that appears warranted, based upon engineering judgment, includes the following:

- Development of a sulfur-tolerant fuel cell
- Development of high-temperature carbon dioxide removal
- Continued systems analysis effort to assess additional novel systems that could not be evaluated within the constraints of this effort
- Improved catalytic combustion systems with lower light off temperatures and higher maximum temperature limits for reduced heat exchange and improved system simplification.

The above recommendations for identified R&D effort are based upon an in-depth analysis of the many ramifications of the overall system by mature chemical process and research engineers. The conclusions are based not only upon the calculated analysis, but also upon first-hand knowledge of the processes of coal gasification and fuel cell power generation.

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OBJECTIVE

The purpose of the work effort described in this document is to evaluate, on a total systems basis, various means for integration of coal gasification with power generation (through molten carbonate fuel cells) with emphasis upon identifying research and development effort that, if fruitful, could be cost-effective in reducing the overall cost of power from such systems.

INTRODUCTION

Approach

The approach employed in evaluating the integration of coal gasification with high-temperature, molten carbonate fuel cells was the development of a versatile computer process simulator that can integrate the various subsystems and provide an analysis of the resulting overall system. Such an approach is useful because it permits the relatively quick comparison of alternative systems and, if used with care, provides a relative comparison of the benefits of process modifications.

Those process configurations that appeared to offer efficiency merit, based upon the process simulator, were then costed, using a factored-estimate approach, and the relative cost of power from each of the systems was projected. More accurate costing techniques could not be employed because the engineering basis underlying several of the novel processing systems was considered insufficient to warrant a more definitive evaluation.

The approach has certain drawbacks. Unless excessively complex, the models for certain process elements (notably the gasifier, the acid-gas removal system, and the fuel cell) are simplified approximations of actual operation. Thus, unless engineering judgement is employed, it is possible that the approach may result in exercising the simulator, rather than actually comparing the anticipated performance of alternative systems.

With the overall system under consideration, extreme care is required to make each simulator test comparable to the others. Because many hundreds of millions of Btu's are being transferred or converted at each step of the process, and because the net power recovered from the system is the algebraic sum of large quantities of energy that are being produced and consumed, an apparently minor change in an assumption can readily cause a change of 25 to 50 million Btu's in recovered power — an apparently significant improvement. Thus, many of the systems discussed have been iterated several times to assure comparability.

As a corollary to the sensitivity discussed above, one must be extremely careful in comparing the evaluations of alternative investigators. The assumptions underlying the effort must be clearly defined and be precisely identical for comparisons to be meaningful. Evaluations presented in this

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effort consistently show lower process efficiencies than others reported in the literature. In a separate subtask (Appendix Da), a systematic variation in the conservatism of the assumptions (underlying the same basic system) caused a change in apparent system efficiency from 45% to 57%. When alternative systems are to be compared, they must be compared with precisely the same assumptions.

The approach is flawed by the high leverage of capital costs upon final cost of power; yet the factored estimating technique that must necessarily be employed cannot be highly accurate. Moreover, the financial factors employed in assessing the cost of power from capital and operating charges inherently tend to favor systems with reduced capital requirements.

In spite of the drawbacks cited above, the approach is quite valuable in providing a first-cut evaluation of alternative systems and is particularly valuable in projecting the anticipated benefits of R&D on novel concepts. When used in conjunction with mature engineering judgement, it provides an excellent basis for assessing projected research effort.

Consistent with the limitations of the approach as set forth above, the following sections describe the process simulator employed and define the assumptions used throughout the effort.

Basic Process Element Simulators

Many of the elements of the process simulator were adapted from an earlier effort that evaluated alternative approaches for the production of substitute natural gas from coal.¹ The elements were modified to include oxidizing species and additional process subsections were added for this effort.

Before discussing the individual elements in the overall system, the general process configuration will be outlined. Figure 1 presents the process configuration that was found to be preferred in many of the evaluations; improved systems were generally modifications of this basic configuration. Appendix Cd presents the computerized process simulator printout for the specific case presented. Each page of the printout represents one step in the overall process; approximately 75 pages of printout are required to character-

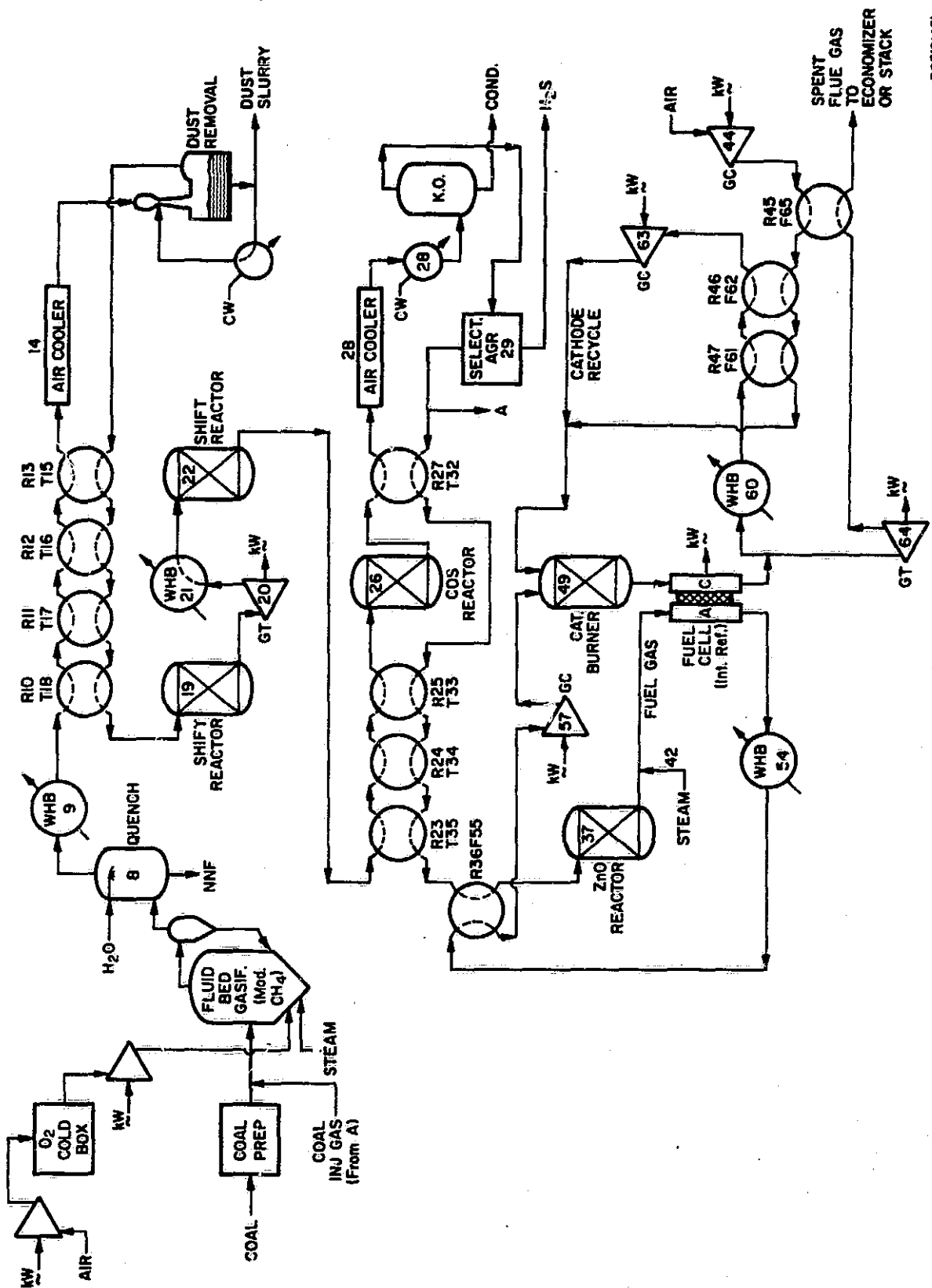
ize a given configuration.* Figure 1, the printout, and the discussion below, when analyzed together, provide the necessary background for evaluating the process simulator employed.

In Figure 1, fuel gas is produced from coal, steam, and oxygen in the primary converter. After preliminary dust removal, the gases are quenched for carbon stability and heat is recovered in a waste heat boiler (WHB). The gas is cooled in a train of crossflow heat exchangers and an air cooler, finally washing out the dust in a non-condensing venturi scrubber (operating at the temperature that is the dewpoint of the gas). The gas is then reheated in the above train, reacted for initial shift conversion, and expanded to fuel cell operating pressure at the maximum temperature of the departiculated gas. After additional heat recovery through a waste heat boiler, another stage of shift conversion is included; followed by a heat exchange train, with intermediate COS hydrolysis; final cooling for condensate removal; and selective acid/gas removal (AGR) for recovery of H₂S but only the cost-effective removal of carbon dioxide. The fuel gas is then reheated in the second heat exchange train, purified over zinc oxide, and steam is added for control of carbon stability in the feed to the fuel cell. Eighty-five percent of the moles of reductant (H₂ + CO + CH₄) are electrochemically converted to power in the fuel cell; the rest are catalytically combusted. At the fuel cell station, recycle (with heat recovery) is employed for 1) temperature control of the fuel cell and 2) maintaining correct temperatures for the booster fans (on both the anode and cathode exhaust gases) as well as the catalytic combustor. In addition to power recovery in the fuel cell, a steam bottoming cycle and a cathode exhaust-gas expander also provide net power output from the system.

Gasifiers

The entrained-flow gasifier model, used in the initial simulations, is a simplified transfer-function model employing the inputs and outputs provided by the prime contractor in the reference case for the system (Appendixes Aa and Ab). The printout page includes gas quantities and compositions, oxygen demand, and outlet stream conditions. Also included is an empirical factor to allow for the power required for coal preparation and feeding of the unit. In

* Because of the volume of the printouts, only the summary printouts of cases other than the one discussed here are appended to the report; others have been placed on file with the prime contractor.



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Figure 1. PROCESS FLOW DIAGRAM OF TYPICAL COAL GASIFICATION/MCFC POWER GENERATION SYSTEM

an analysis of this operation by others, the subsystem was found to be optimistic. For example, 100% of the carbon in the coal is converted to gaseous species with an oxygen requirement of 0.84 lb of oxygen per lb of coal, while pumping a slurry with 70% feed solids concentration.

Four different operating conditions were considered for fluidized-bed gasifiers during this program. The gasifier models are based upon extrapolation of pilot plant performance to higher operating pressures, employing known effects of pressure upon gasification rates and the product gas spectrum. In each case, carbon conversion is 96.9%, reflecting pilot plant data for the amount of carbon lost to dust and gasifier ash. Three of the gasifiers were specified with a steam/carbon ratio of 1:1 molar and an operating pressure of 500 psi. The fourth gasifier was operated with a steam/carbon ratio of 0.75:1 and a pressure of 600 psi — operating conditions that are less conservative considering the state of current technology. (This fourth gasifier was employed in the analysis of variations of initial assumptions, presented in Appendix Da.) The quantity of methane in the fluidized bed gasifier product varies, depending upon the mode of operation as predicted from pilot plant and theoretical evaluations. In one conceptual gasifier, higher methane yield was allotted to correspond with predictions for the novel gasification system. Heat and material printouts for the nonconfidential fluidized-bed gasifiers are presented in Appendix Ce.

The printout for the fluidized-bed gasification output includes gas composition and quantity, off-gas temperature and pressure, steam demand (as inputted), oxygen requirements (as dictated by material and energy balance), and the quantity of recycle gas required to inject the coal into the gasifier. A power demand is indicated to account for coal preparation and gasifier requirements.

Heat Transfer

Each side of a heat exchanger was calculated independently. In many cases, heat transfer was calculated from known inlet and assumed outlet temperatures, using the composition of the flowing gas and three-factor heat capacity correlations. In other cases, for example when heat is being transferred from one gas to another, the quantity of heat to be transferred is fixed and the outlet temperature of the exchanger is calculated.

When integrating the flow sheet, care was taken to avoid temperature crosses in most instances. In other words, because crossflow heat exchange is generally considered, the temperature to which the cool gas is heated may not be warmer than the temperature to which the hot gas is cooled. In most cases, a temperature of approach (approximately 50°F) was imposed. In future work, it is suggested that, when using gas-to-gas heat exchange, the temperature of approach be the controlling design factor and the heat exchangers be designed as a unit. Such a modification would make test-to-test comparisons more valid and minimize operator error in the analysis.

The typical heat exchanger printout page provides gas compositions and quantities, inlet and outlet temperatures, process gas sensible enthalpies, and heat exchanger duty.

Gas Stability

The thermodynamic stability of the gas for carbon deposition is calculated from thermodynamic equilibria (based upon the carbon activities for both amorphous and graphitic carbon)², using free energies based upon six-term* functions of temperature. The quantity of water required to inhibit the carbon deposition is calculated as a function of the assumed temperature, pressure, and initial gas composition.

Quench

The addition of water (as either liquid or steam) to the gas stream results in a temperature redistribution based upon heat capacity data and an empirical representation of water or steam enthalpies. The quantity of water used for this quench is determined from the preceding step, as derived for inhibition to the limits described by the activity of either amorphous carbon (in the assumption sensitivity study -- Appendix Da) or for most of the work effort, midway between amorphous and graphitic carbon, which includes a safety factor in the operation.

Water-Gas Shift Reaction

Steam, carbon monoxide, carbon dioxide, and hydrogen are equilibrated according to the water-gas shift reaction based upon the free energy

* Three terms for heat capacity factors and three additional terms for correlating the heats and free energies of reactions.

calculated at an assumed temperature by using standard free energy data and six-term thermodynamic expressions. The heat of reaction is calculated and the exit gas temperature is determined assuming an adiabatic reactor.

In this step, the equilibrium temperature is assumed and the resulting calculated exit gas temperature is manually compared with the assumptions, to determine if the temperature of approach to equilibrium is satisfactory. The preferred future system would incorporate an inputted desired temperature of approach to equilibrium and iterate the process element to close the heat balance and degree of reaction.

When the water-gas shift reaction occurs within the fuel cell, the equilibrium temperature is assumed to be the "characteristic" temperature of the cell, as discussed later.

Reforming-Methanation

The steam-reforming of methane, or the reverse reaction of methane formation from hydrogen and carbon monoxide, is calculated similarly to the water-gas shift reaction. Although not used as an independent reactor step in this effort, the capability exists in the generalized computer simulator. It was used as a subroutine to modify fuel cell gas composition within the anode compartment of the fuel cell in many of the cases studied.

Other Catalytic Reactions

Most systems include a COS hydrolysis reactor in the heat exchange train as the gas is cooled. The reaction of COS with water to produce H_2S is favored at a lower temperature; therefore, the reactor is located at a point where the gas is slightly above its dewpoint. The reported thermodynamics of this reaction vary over a wide range, depending upon the data source. This effort has assumed equilibrium conversion of COS according to the recent UCI data³ (which are more conservative than JANAF or NASA tables).

After acid-gas removal, traces of H_2S and COS are removed by zinc oxide at about 750°F. The consumption of zinc oxide is presented, based on the estimated sulfur leakage through the acid-gas removal system (as discussed below).

Compressors and Expanders

Rotating power equipment is calculated at the inputted efficiency of enthalpy change into work based upon isentropic pressure differential. Process inefficiencies appear as heat in the outlet gas. A mechanical efficiency is also assumed in the conversion of shaft horsepower into work.

For multiple stage compression, interstage cooling to 90°F is assumed, and the cooling duty as well as the energy requirements per stage, are printed.

Acid-Gas Removal

Process element algorithms are available for amine, hot potassium carbonate, and physical solvent systems for acid-gas removal (AGR). Although other acid-gas systems were checked during the course of the work to determine if advantages existed, the AGR system employed throughout this system was the "Selexol" acid-gas removal process.

The algorithm for this process is a transfer function that removes 1) H₂S from the process gas, 2) CO₂ in accordance with an assumed H₂S concentration in the resultant acid-gas, 3) approximately 50% of the COS, and 4) almost all of the water content of the feed gas (a small quantity of water is left in the gas to avoid computational problems in the downstream fuel cell). Steam and power demands are based on an extensive in-house data bank and are functions of the operating pressure, acid-gas partial pressures, and degree of concentration of the H₂S in the resulting acid-gas.

In this work, the Selexol acid-gas removal system was not specified for total carbonyl sulfide removal because this AGR process option requires extensive power demands for pumping and refrigeration. Rather, in most cases, a COS hydrolysis reactor was included upstream of the AGR and the estimated distribution of COS in the H₂S-selective system was reported.

The simulator algorithm appears directionally correct. For example, it predicts increased steam usage with changes in the subsystem parameters that would increase the solvent flowrate in the AGR subsystem. Similarly, it predicts increased power demands for modifications that would increase the pumping required, the refrigeration load, or the flash gas recompression duty. However, the algorithm is based upon an empirical correlation of published results, rather than upon the basic principles of the AGR system. Therefore,

the algorithm cannot be extrapolated beyond the normal operating regime of such a system.

In addition to the selective AGR described above, the simulator also includes algorithms for bulk removal of acid-gas. Such a process unit was employed in a few tests of the overall system. In this case, the nitrogen required for stripping the solvent of CO_2 is also included.

The Fuel Cell

The fuel cell simulator is also a simplified approximation of the complex operation of the subsystem. The simulator was derived from a Physical Sciences, Inc. (PSI) package that employs a characteristic temperature input for calculations of free energy, open circuit voltage, and cell polarizations. The Nernst concentration polarizations are calculated at the characteristic temperature, using the log mean concentrations of the inlet and outlet gases at both the anode and the cathode. In its original form, the program also calculated electrode polarizations, using a wetted-pore model, and the same gas compositions. The total heat released in the system was calculated and both the anode and cathode exhaust gases leave at the same temperature for an adiabatic cell.

The initial PSI simulator was modified twice during the project. First, the features external to the fuel cell within the PSI-supplied tape were dropped and the fuel cell simulator was incorporated as a subroutine in the overall project computer package. Reforming and shift equilibria were drawn from the project subroutines rather than from the initial PSI package. The PSI simulator, in this form, still contained a relatively thick electrolyte and electrode polarizations that were based upon data published by IGT about seven years ago. Thus, the simulator is not representative of state-of-the-art fuel cell electrodes, nor the new, thinner electrolyte systems being developed in several laboratories. However, at a 10-atmosphere operating pressure, the simulator projects about 925 mV, IR-free, when operated on simulated reformed methane (at 160 mA/cm^2). Extrapolated experimental data⁴ indicate a potential of 935 mV under these conditions. With allowances made for the differing fuel utilizations, the correspondence is within 5 mV. Thus, the simulator, after the initial modifications, gave reasonable results at high-pressure operation. However, at 1-atmosphere operating pressure, the simulator projected only 520 mV (IR-free), much less than the 852 mV

experimental output. Thus, the use of the simulator at pressures much different than 10 atmospheres (or at gas partial pressures that vary greatly from the experimental conditions) should be avoided.

Midway through the project, it was found that the fuel cell subroutine would not function in conjunction with internal reforming within the fuel cell when the gas contained moderate methane concentrations. The conversion technique employed in the simulator was too sensitive with the changing quantities of hydrogen produced by reforming. At this point in the project, the internal programming of the simulator was modified in three ways:

1. The fuel cell hydrogen conversion was changed to a stepwise approach, with methane reforming, in shift equilibrium, after each step.
2. The electrode polarizations were modified to simple, current-dependent approximations with a factor that resulted in the same polarization that had been noted in the prior tests.
3. The expression for the free energy of the reaction was adjusted to be more consistent with the JANAF tables.

The elimination of the wetted-pore model, and the substitution of a simpler expression, further limits the applicability of the subsystem simulator to near the range in which it was utilized. Application of the model to a higher pressure fuel cell, as had been considered in this project, was therefore not attempted.

The modification of the expression for the free energy caused a 54 mV loss in cell potential. This voltage loss, in the systems considered, caused a 39 MW change in the power delivered by the fuel cell, or about a 25 MW change in total plant output, a variation of 1.6 percentage points in system efficiency.* Thus, the modified fuel cell simulator is somewhat more conservative than results shown by 1980 fuel cell data. It is significantly more conservative (approximately 2.5 percentage points in efficiency) than recent fuel cell data.⁵

The fuel cell simulator is simplified. For example, it does not differentiate the directions of gas flow with the cell. Considering the imperative requirements of 1) uniform gas flow across the face of a single cell and 2) precise gas distribution, cell-to-cell, in a fuel cell stack, it is likely

* As presented in Appendix Da.

that future systems will be cocurrent in operation. Several of the projected designs were double-checked against total Nernst polarization, assuming zero current flow at the common outlet of a cocurrent cell. These results did not give a voltage less than the average total cell polarization provided by the simulator and were, therefore, not controlling. However, as is shown later, the total system performance is quite sensitive to the fuel cell operating characteristics and its off-gas temperature. An integrated, differential simulator for a cocurrent cell, including temperature effects, would be highly desirable for an improved process simulator.

Catalytic Combustor

The combustion of the anode exhaust into the cathode feed is assumed to take place catalytically. The subsystem simulator integrates the heat of reaction with the chemical changes taking place and calculates an off-gas temperature. Warnings appear if the mixed feed temperature is less than the light-off temperature for conventional catalysts, or if the adiabatic reaction temperature is sufficiently high to cause long-term catalyst degradation.

Major Assumptions -- System Simulator

System Pressure Drops

Pressure drops of 5 psi were assumed across packed bed reactors such as the water-gas shift reactor, carbonyl sulfide hydrolysis reactor, zinc oxide bed, and catalytic combustion. Similarly, 5 psi was taken across the acid-gas removal absorbers.

A 3-psi drop was taken across particulate scrubbers. A 2-psi drop was taken across each shell of a heat exchanger train and the fuel cell itself. Splitting or merging streams were assumed to require 1-psi drop.

The above pressure drops would appear quite optimistic compared with normal chemical engineering practice. Yet several system designs in the literature have shown much lower pressure drops than presented here. The impact of lower pressure drops is improved energy recovery in the overall system. A few cases were evaluated at lower pressure drops (in the analysis of sensitivity to assumptions presented in Appendix Da) to indicate the magnitude of this effect.

Efficiency of Rotating Power Equipment

In most instances, turboexpanders were taken at 92% isentropic efficiency and large axial compressors were taken at 88% efficiency. A mechanical efficiency of 98% was also imposed. These values were taken from the base case provided at the beginning of the project. Although slightly optimistic relative to the anticipated performance of generally available machinery, the values were maintained throughout the project for consistency. One series of calculations was made to indicate the impact of driver efficiency upon the overall system.

The compression efficiency of the multiple-stage oxygen compressor, feeding the gasifier, was taken at 62%, to reflect the higher power demand of these non-lubricated, slow-speed, reciprocating machines.

The maximum compression ratio was varied as a function of the gas being compressed. For air, this maximum was normally a ratio of 3.5. However, with final stages of air compression, specialized outboard designs were considered that would permit higher compressed air temperatures and compression ratios as high as 4.0. Note that the humid air to the system is assumed to contain 14.3-psi partial pressure dry air, at reasonable site elevations above sea level.

Heat Exchangers

Generally, heat exchangers were assumed to be crossflow with an approach temperature differential of 50°F between the maximum temperature of the cooler gas and the minimum temperature of the warmer gas. The assumption of counter-current heat exchanger and/or reduced temperature of approach will decrease the heat losses of the system and improve the overall system efficiency.

Although countercurrent heat exchangers were not normally employed, if the heat exchange duty was large and multiple shells of heat exchange would normally be required, the shells were arranged in series to provide more efficient heat transfer. This approach improves the heat recovery within the system at the expense of greater system pressure drops; however, it does not approach the temperature recovery for countercurrent heat exchangers by perhaps 100°F temperature loss in each heat exchanger train.

Chemical Reactions

The water-gas shift reactors were assumed to have a temperature of approach to equilibrium of approximately 50°F to simulate end-of-life for the catalyst. A reduced approach to equilibrium will improve the overall system efficiency by converting more of the water vapor present into hydrogen, reducing heat loss to condensation, and reducing downstream humidification requirements.

The water-gas shift and reforming operations were assumed to reach equilibrium when these reactions take place within the fuel cell where the gas flow is laminar. However, the equilibrium is based upon the "characteristic" temperature of the cell. Should the outlet gas temperature of the cell be used, slightly improved system performance would be expected.

Complete combustion was assumed in the catalytic reactors on the anode tail gas.

The carbonyl sulfide hydrolysis reactors were assumed to reach equilibrium, using the free energies of carbonyl sulfide recently developed by United Catalysts, Inc.

Zinc oxide beds were assumed to operate at 750°F and provide complete removal of both hydrogen sulfide and carbonyl sulfide.

Steam System

Because the total facility for conversion of coal to power has not been described, and certain steam sources and demands have not been quantified, the steam system was not defined in detail. Additional energy demands or production that might be expected of the total system include:--

1. Both high- and low-pressure steam generation in the sulfur plant
2. The steam and reductant demand in the sulfur plant tail gas treatment unit
3. The steam demand, both low and moderate pressure, for waste water stripping and ammonia recovery
4. The power demands for activated sludge treatment of water before recycle
5. Boiler feedwater and cooling water pumping (for process units)
6. The power demands for waste solids handling

7. Miscellaneous power demands for facilities, instrument air, flare, and other systems that will be required in the overall facility.

The impact of these miscellaneous steam, power, and reductant demands, or sources, indicates that the total system cannot be quantified with the available information. Therefore, the steam system was not optimized.

In the project, the steam required for humidification of the anode fuel was assumed to be recovered as high-pressure, superheated steam, similar to the reference case supplied by the prime contractor. Likewise, that steam recovered into power was also assumed as high-pressure, superheated steam, with reheat. Other steam demands within the system were assumed to be recovered at the conditions for which they were required. For example, the gasifier steam requirement was assumed to be recovered at 600 psi, 750°F, and the AGR steam demands were assumed to be recovered at 50 psi.

In the appended discussion on variation of assumptions, the total heat recoverable to high-pressure steam was also assumed to be manageable for the production of high-pressure, superheated steam. Boiler feed water was preheated in an economizer and by contact with the cooling gas stream. Lower temperature heat was assumed to be available for vaporizing the water and higher temperature heat was available for providing superheat. The effort was not made to quantify the source of heat as a function of temperature; rather, it was assumed that the total heat discharged to the system could be recovered in this fashion. In that optimization study, significant power could be recovered by expanding the high-pressure steam to the duty required in the system.

The high-temperature, superheated steam was assumed to be recoverable to power at 35.28% efficiency. This efficiency includes allowances for boiler feedwater heating and pumping, as well as cooling tower duty and cooling water pumping. This efficiency might be considered low by modern power plant standards (if stack gas scrubbing is not required) and, in the assessment of the assumptions, an alternative efficiency of 40% was also assumed.

In future efforts, it is recommended that the energy sources and requirements outside the basic plant be identified and factored into the overall energy balance. Then, the steam system could be quantified to identify heat sources by quality of heat and verify that sufficient heat is available at the necessary quality to provide the steam power at the design conditions.

The Fuel Cell Subsystem

Appendix Cb presents a discussion of the voltage and power produced in the fuel cell. The determination of voltage losses in the cell, as projected by the modified PSI simulator, was presented earlier. In general, current density was taken at 150 mA/cm^2 and the conversion of total fuel gas species ($\text{H}_2 + \text{CO} + \text{CH}_4$) was 85%. Although short optimization tests indicate that these assumptions are not ideal, they were maintained for comparability to other work.

An effort was made to specify operating units around the fuel cell subsystem that would be operable. The fans, or low-compression-ratio boosters, that are used on both the anode and cathode exhausts have been protected against excessive temperature by heat exchange. Additionally, feeds to the catalytic combustor were properly temperature-conditioned such that this unit should be in the proper operating temperature range. Additional heat exchange equipment was required for this design consideration.

In general, the temperature of the feed gases to the fuel cell were greater than 1000°F to avoid excessive polarization at the gas inlet. These streams were temperature-conditioned with additional heat exchange equipment. In most instances, the outlet gas temperatures were about 1260°F . By careful tuning of the simulator, through adjustment of the cathode gas recycle rate, the exit gas temperature could be increased to 1300°F , with a resultant increase in overall system power output.

Major Assumptions — System Costing

The prime contractor provided a base case reference design and economic analysis to estimate the investment, operating costs, and levelized power costs for a 675 MW fuel cell power plant using an entrained-flow coal gasification system. That complete report is included with this document as Appendix Aa. The costing assumptions employed herein, together with the financial factors, came from the base case economics (included as Appendix Ab) and were used directly in the economic assessment of the current effort when applicable.

The nine cases evaluated economically in this document often contain gasification and other processing steps that are different from the base case, requiring additional costing information. The assumptions listed in Table 1

were added to those presented in the reference design to evaluate alternative processing schemes.

Table 1. SYSTEM COSTING ASSUMPTIONS

<u>Fuel Processing Unit</u>	<u>Cost Base</u>	<u>Scaling Capacity Factor</u>	<u>Scaling Exponent</u>
Fluidized Bed Gasification and Ash Handling	M. W. Kellogg report on Westinghouse and U-Gas Processes for Coal to SNG ⁶	Tons/day coal	0.6
Gas Cooling and Quench	M. W. Kellogg report on Westinghouse and U-GAS Processes for Coal to SNG ⁶	Gas molar flow	0.82
Water-Gas Shift	M. W. Kellogg report on Westinghouse and U-GAS Processes for Coal to SNG ⁶	Moles of CO shifted	0.6
High-Temperature CO ₂ Removal	In-House	Tons/day absorbent	0.78
High-Temperature Sulfur Removal	In-House	Tons/day mixed oxides	0.78
Fuel Cells	Base Design	Cell area	1.0
Gas Turbine and Steam Cycle	Base Design	Power output	1.0
DC/AC Inverters	Base Design	Fuel cell output	1.0
Accessory Electrical Equipment	Base Design	Total plant output	1.0

RESULTS: ECONOMIC COMPARISON

General

During this project, 66 cases were examined for the integration of coal gasification with molten carbonate fuel cells for power generation. Of these, 20 cases were analyzed for the change in system efficiency by variation in process configuration. These initial cases were based upon a constant 4991 tons/day of candidate coal input (the quantity required for the production of 675 MW in the reference case in an early iteration) and the resulting power output was assessed. When the sensitivity of the overall system to input assumptions was realized, many of these cases were iterated several times for more accurate comparisons. The results of this analysis are presented in Appendix Ca.

As discussed elsewhere, the system is very sensitive to the engineering assumptions inputted by the systems analyst. A subprogram was undertaken to indicate the effect of various input assumptions; this assessment (21 cases) is presented in Appendix Da. Summary printouts for these cases are presented in Appendix Db.

Many of the systems analyzed are not specifically presented herein. In general, many of these cases represent minor variations in a specific analysis or involved optimization studies by varying a single process parameter. In these cases, the results are discussed but not specifically analyzed in depth.

Nine of the conceptual systems were reevaluated at a constant 675 MW power output to provide a constant basis for the economic analysis of the alternatives. When recalculated on the process simulator, the necessary flows at all stages of the system were evaluated to aid in the system costing. The reevaluation of these systems for the economic analysis was made after the sensitivity assessment and, consequently, an effort was made while reassessing these data to ensure comparability between the systems. Therefore, the system efficiencies vary slightly from the initial work presented in Appendix Ca.

The nine cases were selected for economic analysis based upon anticipated efficiency improvements, as derived from the earlier work, the systematic incorporation of system advances to be expected through R&D, and the potential for success in the R&D effort. Because the overall goal of the program was to identify cost-effective R&D efforts as applied to this system, the major

thrust was for advanced concepts, rather than shorter-term, engineering solutions.

A number of concepts were identified that could not be costed within the constraints of the program. These concepts are discussed, directionally, based upon the analyses of similar efforts.

The nine cases evaluated in the economic analysis, and the rationale for their selection, were as follows:

1. **The Base Case:** This system includes an entrained-flow gasifier, thermodynamically unstable raw gas, no shift reactor, and countercurrent heat exchange. The case is included to provide a direct comparison to earlier work.
2. **Modified Base Case:** The base case was reengineered for operability, adding a quench for carbon stability, water-gas shift reaction for improved raw gas hydrogen content, and crossflow heat exchange for improved operability. Location of the turboexpander was moved for greater energy recovery. This case was included to indicate the cost impact of the desired modifications and becomes a new reference for the remainder of the systems that are designed to a similar degree of operability.
3. **Fluidized-Bed Gasifier:** This case is similar to the Case 2 above, with the substitution of a fluidized-bed gasifier, producing low methane off-gas (substituted for the entrained-flow gasifier). Although this case showed lower efficiency than Case 2 above, it was included because it is a starting point for further improvements indicated below.
4. **Internal Reforming Fuel Cell:** Internal reforming capability was added to the fuel cell of the system in Case 3 (using the fluidized-bed gasifier) to indicate the merit of this concept.
5. **Modified Fluidized Bed Gasifier; to produce greater methane content:** Case 4, above, was modified only by the mechanical configuration of the gasifier so that it would produce greater quantities of methane for greater efficiency both in the gasifier and in the internally reforming fuel cell.
6. **High-Temperature Dust Removal:** The addition of high-temperature dust removal, per se, adds little to the overall system. However, it permits system modifications. In this case, the turboexpander could be relocated to provide significant improvement in process efficiency, as well as a cost reduction by elimination of much heat exchange equipment. The high-temperature dust removal is also employed in other systems below.
7. **Sulfur-Tolerant Fuel Cell:** The combination of high-temperature dust removal with a conceptual sulfur-tolerant fuel cell represents an extremely simple system, with minimum equipment, of relatively high thermal efficiency.

8. High-Temperature CO₂ Transfer: The concept of removal of carbon dioxide from the raw fuel gas and transferring it to the cathode is attractive because of the higher-purity fuel feeding the fuel cell and greater fuel cell output. The system in Case 5 was modified to include this effect; somewhat greater efficiency was achieved.
9. High-Temperature Sulfur Removal: The integration of high-temperature dust removal with high-temperature sulfur removal provides a system that is similar in efficiency and complexity to the sulfur-tolerant fuel cell (Case 7) discussed above.

Other cases considered, but not specifically analyzed for cost, include the following:

10. Incorporation of a Novel Fluidized-Bed Gasifier: A novel, single-stage fluidized-bed gasifier has been conceived that promises still higher methane production. From the efficiency analyses, incorporation of this gasifier into the system with an internally reforming fuel cell promises improved system efficiency.
11. Integration of High-Temperature Sulfur Removal with a Sulfur-Tolerant Fuel Cell: The system including the sulfur-tolerant fuel cell does not remove sulfur dioxide from the final stack. High-temperature sulfur removal, to the degree required for environmentally satisfactory utilization of coal, has now been demonstrated on the laboratory and bench scale. Although laboratory results on high temperature sulfur removal to the degree required for conventional fuel cells appear promising, this combination appears attractive as a backup position.
12. Dust-tolerant Shift Reactor: A system incorporating a dust-tolerant shift reactor indicated improved process efficiency and system simplicity. It was not separately costed because extrapolations from Case 6 above would be applicable.
13. Advanced Systems: Several systems were conceived during the course of the program that could not be evaluated, even on an efficiency basis, because of the constraints of time, budget, or inapplicability of the fuel cell simulator. Such systems appear promising, based upon first principles, but were beyond the scope of this investigation.

Table 2 presents the efficiency comparisons for the nine cases outlined above, as developed from the computerized process simulator. The reader is specifically directed to Appendix Ca for the discussion of the individual cases and the factors that impact the efficiency. More detailed discussions of the effects that occur in the system are presented in Appendix Cb and Appendix B presents summary printouts of the computerized analysis for the nine cases outlined above. Each of the cases is described briefly later in this report section, after the economic analysis is discussed.

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Table 2. EFFICIENCY COMPARISONS OF ALTERNATIVE SYSTEMS

Case	Reference (Appendix Ca)	Features	Input 10 ⁹ Btu/h	Power Output			Eff	
				Net Mech	Fuel Cell	Steam Cycle		
1	I	Base Case, Entrained Gasifier Unstable Raw Gas - No Shift, Counter-current Heat Exchange	5.47	8	553	114	675	42.1
2	IF	Add Quench, Shift, Crossflow Heat Exchange	5.54	40	570	65	675	41.6
3	II	Change to Fluidized-Bed Gasifier, Low CH ₄	6.00	27	558	69	675	38.4
4	III	Add Internal Reforming to Fuel Cell	5.13	46	617	11	675	44.9
5	IVe	Change Gasifier to Moderate CH ₄ Production	5.03	51	620	4	675	45.8
6	VI	High-Temperature Dust Removal	4.86	71	601	3	675	47.4
7	VII	Sulfur-Tolerant Fuel Cell	4.61	47	555	73	675	50.0
8	VIII	High-Temperature CO ₂ -Transfer	4.85	7	635	33	675	47.5
9	IX	High-Temperature Sulfur Removal	4.60	49	554	72	675	50.0

Table 2 illustrates one of the basic tenets of systems analyses — an optimized overall system is not necessarily made up of components operating at their optimal condition. For example, the fluidized-bed gasifier of Case 3 has a higher efficiency than the entrained-flow gasifier of Case 2, with both operating in the same system. However, the entrained-flow gasifier yielded a higher overall system efficiency. This efficiency was recovered, including the benefits of the improved gasifier, when internal reforming was added to the fuel cell, as in Case 4.

Systems 7 and 9, incorporating either high-temperature sulfur removal or the sulfur-tolerant fuel cell (both in conjunction with high-temperature dust removal) produce nearly the lowest fraction of the total power from the fuel cell itself. This is because the gas fed to the fuel cell has relatively high concentrations of carbon dioxide and water vapor, suppressing the fuel cell voltage. Nevertheless, on the total systems basis, they show the highest overall efficiency. Thus, compromises taken in the fuel cell operating conditions result in favored overall system performance. Conversely, System 8, with high-temperature carbon dioxide transfer, shows the highest voltage from the fuel cell; yet the overall system performance is not as high as in Systems 7 and 9. Note that System 8 did not include the high-temperature dust removal in conjunction with the high-temperature carbon dioxide transfer. Should this combination have been included, it is conceivable that the overall system efficiency would again be higher.

Table 2 also indicates that, in general, high-temperature purification systems showed efficiency advantages over their low-temperature counterparts. High-temperature dust removal, CO₂ transfer, and sulfur removal are all preferred system options.

Of note in Table 2 is the relative power recovered in system elements. For example, Systems 5 and 6 provide very little power from the steam cycle; it may be reasonable in these cases to simplify the overall system by optimizing the conversion taken across the fuel cell and completely eliminating the steam cycle from the system.

Tables 3, 4, and 5 provide the economic analysis for the nine cases evaluated. The techniques of calculation follow the format suggested by the prime contractor and were derived EPRI costing bases. The assumptions used in arriving at the capital cost and operating expenses were presented earlier in

Table 3. SUMMARY OF PROCESS (On-Site) COSTS FOR 675 MW POWER PLANT USING
COMBINED COAL GASIFICATION/FUEL CELL OPERATION

Case No.*	1	2	3	4	5	6	7	8	9
	\$10 ⁶								
<u>Component</u>									
Gasification Section									
Oxygen Plant	87.2	88.2	77.3	68.6	63.8	62.4	59.8	62.1	59.8
Coal Handling	18.8	19.1	20.3	18.0	17.6	17.3	16.6	17.2	16.6
Coal Feed, Gasification and Ash Handling	23.5	23.6	15.5	14.1	13.9	13.6	13.2	13.6	13.2
Gasifier Effluent Cooling, Quench and Dust Removal	45.4	45.9	19.3	16.9	17.0	15.6	14.7	15.4	15.9
High-Temperature CO ₂ Removal	--	--	--	--	--	--	--	28.5	--
High-Temperature Sulfur Removal	--	--	--	--	--	--	--	--	18.6
CO-Shift and Heat Recovery	--	31.6	26.0	23.6	22.6	22.2	--	24.0	--
Acid-Gas Removal	38.5	47.0	50.6	46.0	43.7	42.9	--	38.4	--
Zinc Oxide Treatment	4.6	5.7	5.9	5.4	5.0	4.9	--	4.4	--
Sulfur Recovery	<u>12.1</u>	<u>12.6</u>	<u>13.4</u>	<u>12.2</u>	<u>12.0</u>	<u>11.8</u>	--	<u>11.6</u>	<u>1.0</u>
Fuel Processing Section Subtotal	230.1	273.7	228.1	204.8	195.6	190.7	104.3	215.2	124.9
Power Recovery Section									
Fuel Cells Including Piping, etc.	145.2	147.1	146.8	169.9	172.1	166.7	176.1	166.1	158.0
Gas-Turbine and Steam Cycle	91.9	57.2	95.3	51.5	50.8	58.3	81.9	44.1	73.7
Inverters	47.3	50.6	46.1	55.4	55.6	53.8	46.7	57.2	49.0
Other Electrical Equipment	<u>25.6</u>	<u>25.6</u>	<u>25.6</u>	<u>25.6</u>	<u>25.6</u>	<u>25.6</u>	<u>25.6</u>	<u>25.6</u>	<u>25.6</u>
Power Recovery Section Subtotal	310.0	280.6	313.8	302.4	304.2	304.4	330.3	293.0	306.3
Total Process On-Site	540.1	554.3	541.9	507.2	499.8	495.1	434.6	508.2	431.2

* Description of cases presented in Table 2.

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Table 4. SUMMARY OF CAPITAL REQUIRED FOR 675 MW POWER PLANT USING COMBINED COAL GASIFICATION/FUEL CELL OPERATION

Case No*	1	2	3	4	5	6	7	8	9
	\$10 ⁶								
A. Process (On-Site) Costs (Table 3)	540.1	554.3	541.9	507.2	499.8	495.1	434.6	508.2	431.2
B. General Facilities Total Constructed Cost - 10% of Process Capital -	54.0	55.4	54.2	50.7	50.0	49.5	43.5	50.8	43.1
C. Engineering and Home Office Overhead plus Fee - 10% of Process Capital -	54.0	55.4	54.2	50.7	50.0	49.5	43.5	50.8	43.1
D. Process Contingency - Composite by Process Unit and State of Art -	97.9	100.1	93.9	106.7	107.9	104.6	106.6	109.1	100.3
E. Project Contingency - 30% of Total Plant Cost - (A + B + C) x 0.3	194.4	199.5	195.1	182.6	179.9	178.3	156.5	182.9	155.2
F. Total (TPC)	940.4	964.7	939.3	897.9	887.6	876.9	784.7	901.8	772.9
<u>Allowance for Funds During Construction (AFDC)</u>									
AFDC = TPC (0.0373)	35.1	36.0	35.0	33.5	33.1	32.7	29.3	33.6	28.8
<u>Prepaid Royalties</u>									
= 0.5% (TPC - General Facilities - Project Contingencies on General Facilities)	4.4	4.6	4.5	4.3	4.3	4.2	3.8	4.3	3.7
<u>Preproduction or Start-Up Costs</u>	28.9	28.0	27.8	26.0	25.7	25.3	22.8	25.9	22.7
1 Month Fixed Operating Costs									
1 Month Variable Operating Costs									
1 Month Capacity Fuel x 0.25									
2% of TPC									
<u>Inventory Capital</u>									
60 Days Inventory of Fuel, Catalyst, and Chemicals at Full Rate	12.6	13.0	14.0	12.0	11.8	11.4	10.5	11.5	10.8
Initial Catalyst and Chemical Charge	1.8	8.1	7.6	7.0	7.0	7.0	--	9.2	7.0
Land Cost, 164 Acres at \$5500/Acre	0.9	0.9	0.9	0.9	0.9	0.9	0.9	0.9	0.9
Total Capital Required (TCR)	1024.1	1055.3	1029.1	981.6	970.4	958.4	852.0	987.2	846.8

* Description of cases given in Table 2.

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I N S T I T U T E O F T E C H N O L O G Y

Table 5. ANNUAL OPERATING COSTS FOR PRODUCING 675 MW POWER FROM
COMBINED COAL GASIFICATION/FUEL CELL OPERATION

Case No.*	1	2	3	4	5	6	7	8	9
	\$10 ⁶								
<u>Fixed Operating Costs</u>									
Operating Labor	3.74	3.74	3.74	3.74	3.74	3.74	3.74	3.74	3.74
Maintenance Cost									
Material	10.32	11.02	10.45	9.71	9.52	9.40	7.74	9.82	7.85
Labor	6.89	7.35	6.97	6.48	6.34	6.27	5.15	6.54	5.24
Overhead Charges	<u>3.19</u>	<u>3.33</u>	<u>3.21</u>	<u>3.06</u>	<u>3.02</u>	<u>3.00</u>	<u>2.67</u>	<u>3.09</u>	<u>2.69</u>
Subtotal	24.14	25.44	24.37	22.99	22.63	22.41	19.31	23.19	19.52
<u>Annual Variable Operating Costs</u>									
Fuel at 65% Operating Factor	49.24	49.92	54.09	46.22	45.28	43.86	41.52	43.72	41.48
Water	0.34	0.30	0.30	0.30	0.30	0.30	0.30	0.30	0.30
Ash Disposal	0.59	0.59	0.64	0.55	0.54	0.52	0.49	0.52	0.49
Catalysts and Chemicals (Ex Fuel Cell)	<u>0.33</u>	<u>0.38</u>	<u>0.35</u>	<u>0.33</u>	<u>0.33</u>	<u>0.33</u>	--	<u>0.43</u>	<u>0.33</u>
Subtotal	50.50	51.19	55.38	47.40	46.45	45.01	42.31	44.97	42.60
<u>Levelized Revenue Required for First 10 Years</u>									
Capital Related Charges	184.3	190.0	185.3	176.7	174.7	172.5	153.4	177.7	152.4
Fixed Operating Costs	35.9	37.9	36.3	34.2	33.7	33.3	28.7	34.5	29.0
<u>Variable Operating Costs</u>									
Coal	76.0	77.1	83.5	71.3	69.9	67.7	64.1	67.5	64.0
Others	1.9	1.9	1.9	1.8	1.7	1.7	1.2	1.9	1.7
Fuel Cell Replacement Cost	<u>27.6</u>	<u>29.1</u>	<u>29.0</u>	<u>33.6</u>	<u>34.0</u>	<u>33.0</u>	<u>31.2</u>	<u>32.9</u>	<u>31.2</u>
Total Levelized Revenue Required	325.7	336.0	336.0	317.6	314.0	308.2	278.6	314.5	278.3
Annual Electric Power Production, 10 ⁹ kW									
Cost of Electricity, \$/kWh	0.085	0.087	0.087	0.083	0.082	0.080	0.072	0.082	0.072
3.8435									

* Description of cases given in Table 2.

the "Assumptions" section of this document and are illustrated in Appendix Ab. These assumptions were employed directly to give a consistent comparison basis. It should be emphasized, however, that variations in the assumed financial factors strongly influence the cost of electricity; further, such assumptions can affect the relative ranking of the systems.

The relative cost-of-power derived in this analysis must be used with care. Significant effort was expended in evaluating a number of alternative concepts to identify systems with relatively high overall system thermal efficiency. Yet, this system efficiency is often overshadowed by the extremely high leverage effects of capital upon cost of power.

The financial factors employed in the analysis of the cost of electricity for the various systems also tend to emphasize capital-related areas, rather than more highly efficient systems:

- The contingency allowances range from 43% to 50% of the estimated base capital cost. This assumption increases the relative capital cost charges in the electricity, and diminishes the relative importance of the cost of fuel.
- The levelized cost of electricity is taken over the first 10 years of plant life. This assumption tends to deemphasize the cost of fuel, which would continue to escalate and show greater relative importance over a 25- or 30-year levelized cost of electricity.
- The 65% load factor also emphasizes capital-related charges, relative to a higher load factor that might be anticipated on a base-loaded facility.

In the systems considered, the capital-related charges (including the fixed operating costs that are largely capital-driven) are approximately 70% of the cost of electricity; the coal costs, reflecting the efficiency of the system selected, represent 20% to 25% of the cost of power. Thus, with the underlying financial assumptions employed, an appreciable change in system efficiency can be readily overpowered by a moderate variation in initial capital requirement estimates.

For example, System 3 operates with a fluidized-bed gasifier in place of the entrained-bed gasifier of System 2, with the rest of the system essentially identical. System 3 had an overall system thermal efficiency of about 38%, compared with 41% for System 2. Yet, because of the somewhat simpler or smaller gasifier, oxygen plant, and raw gas quench, System 3 has a process equipment cost advantage of \$12 million out of approximately \$550 million

(about 2.2%). Yet, this small change in capital cost is sufficient to overcome the efficiency advantage of System 2, and both systems result in approximately the same cost-of-power.

It is, therefore, emphasized that mature engineering judgement must be employed in evaluating the alternative cases for cost-effective research and development, rather than absolute reliance upon the numbers that have been generated.

Discussion of Systems Analyzed

The drawings presenting simplified process flow diagrams for the nine cases are included at the end of this discussion; these drawings are repeated in Appendix Ca, where more complete discussions of the cases are presented, along with discussions of the efficiency impacts of the various alternatives.

Case 1 — The Reference Case

The reference system for integration of coal gasification with power generation through molten carbonate fuel cells was supplied to the program by the prime contractor. It employed an entrained-flow gasifier and a simplified process configuration (Figure 2). When analyzed on the process simulator, the system provided a process efficiency of 42.1% and, when costed, resulted in a cost of electricity of 8.5¢/kWh. This cost of electricity is similar to that provided by the prime contractor in his base case analysis; however, the details of the cost factors are much different: The current process simulator results in much higher power output from the fuel cell but less energy recovered from the bottoming cycles.

Case 2 — Conservative Base Case

Case 2 (Figure 3) employs the entrained flow gasifier but incorporates additional features to improve the operational stability, safety, and design conservatism. A quench step has been added to eliminate thermodynamic potential for carbon deposition and the heat exchange has been converted from countercurrent to crossflow operation. A portion of the loss in efficiency was recovered by incorporation of a shift reactor and improving the location of the fuel gas turboexpander in the system. Overall process efficiency fell to 41.6% and the cost of electricity increased to 8.7¢/kWh. Note that the decrease in efficiency caused only a minor change in the levelized revenue requirements; the coal accounts for only 23% of the total cost of power. The

major factor in the cost change is a 3% increase in the total capital required for the facility.

Case 3 — Fluidized-Bed Gasifier

The process design employed for Case 2 was modified only by changing the gasifier to a fluidized-bed type; the remaining process elements were kept the same (Figure 4). Temperatures and flows varied because of the different characteristics of the gasifier. The gasifier selected was designed for minimal methane production at the 500 psi operating pressure. This methane is not reacted in the fuel cell but, rather, is only recovered as heat in the steam bottoming cycle. The overall process efficiency dropped to 38.4%. However, the cost of power remained the same at 8.7¢/kWh. The reduced efficiency was economically counterbalanced by only a 2.5% change in total capital requirement. This reduction in cost was due to simpler heat recovery from the off-gases of the lower temperature gasifier and the reduced oxygen demand of the system.

Although the reduced cost of the system is probably directionally correct, the technique of factored estimates used in the costing procedure can not have great accuracy. Because of the high leverage of capital costs on the final price of electricity, the relatively high loss in system efficiency is not of great importance. Thus, the results of this study, particularly the comparison of Cases 2 and 3, must be used with care. Nevertheless, the use of a fluidized-bed gasifier in such a system, although apparently of lower efficiency, merits further evaluation.

Case 4 — Internal Reforming

Case 4 (Figure 5) employs the identical system used in Case 3, incorporating a low-methane, fluidized-bed gasifier, but adds the developmental concept of reforming the methane within the fuel cell with chemical and thermal integration to the fuel cell. The remainder of this system is nearly identical, except for the variations in output energy allocation. In spite of the capital-related developmental charge applied against the novel fuel cell concept, the total capital requirement dropped by over 4%, the efficiency increased by 6.5 percentage points, and the overall cost of power dropped to 8.3¢/kWh. Apparently, the concept of internal reforming in the fuel cell is desirable, when integrated with the fluidized bed gasifier.

Case 5 — Higher Methane Production

By modification of operating techniques, the fluidized-bed gasifier can produce higher quantities of methane (Figure 6 — a repeat of Figure 1). Such a configuration was employed in the system described above; the cost-of-power dropped to 8.2¢/kWh.

One case not analyzed economically (Case IVa of Appendix Ca), included a conceptual fluidized gasifier that would permit even greater methane production. Analysis of the directional trends of Cases 4 and 5 indicates that this gasifier warrants development.

Case 6 — High-Temperature Dust Removal

The process of Case 5 was modified to include removal of dust at high temperature from the raw gasifier product (Figure 7). This approach simplifies the system, eliminates much heat exchange and gas washing equipment, and enables the gas expansion turbine to operate at higher inlet temperatures. The overall process efficiency increased by 1.6 percentage points, the capital costs dropped slightly, and the cost-of-electricity decreased to 8.0¢/kWh. Therefore, it is concluded that current work on high-temperature dust removal should be accelerated.

An alternative system that was not evaluated economically employs a dust-tolerant water-gas shift reactor (Case V of Appendix Ca). This option resulted in the same process simplicity as the high-temperature dust removal and had a similar process efficiency. The concept is relatively straightforward and only a minor portion of the capital in the plant would be impacted by developmental charges. It is, therefore, suggested that this concept also merits investigation.

Case 7 — Sulfur-Tolerant Fuel Cell

This analysis assumed that a fuel cell can be developed that will operate directly on the raw gas after departiculation and cooling (Figure 8). Catalysts for this duty have been preliminarily identified. The system is considerably simplified and overall process efficiency increased to 50.0%. The cost-of-power decreased to 7.2¢/kWh.

It should be noted, however, that the system described does not treat the sulfur in the coal. Some type of stack gas scrubber should be added to the

system. If that stack gas scrubbing costs \$150/kW, the apparent benefits of the sulfur-tolerant cell are negated.

Nevertheless, the system has merit. High-temperature sulfur removal systems (similar to those discussed later) have now been identified and tested (on a PDU scale) to provide sulfur removal in excess of EPA New Source Performance Standards for coal combustion, although not to the extreme purity required for fuel cell application. The integration of this subsystem with a sulfur-tolerant cell may have significant potential. The concept also has value as a backup position should the current high-temperature desulfurization work, directed at the extreme purities required for fuel cells, not be fruitful when scaled up.

Case 8 — High-Temperature CO₂ Transfer

A concept has been identified that permits energy-efficient transfer of carbon dioxide from the hot, high-pressure fuel gas directly to the cathode inlet (Figure 9). This approach improves the purity of the gas feeding the fuel cell because the carbon dioxide concentration has been reduced and the concentration polarization within the fuel cell is minimized. By incorporation of this subsystem, the system efficiency is improved by nearly 2 percentage points. However, the capital cost of this option appears to be about 1.7% greater than without the option, negating the efficiency improvement and resulting in an identical cost of power of 8.2¢/kWh (when referred to Case 5 as the starting point).

The integration of high-temperature CO₂ transfer with high-temperature dust removal may be an attractive option (similar to integration of the sulfur-tolerant cell with high temperature dust removal). The improved process simplicity and efficiency of the two systems would appear to be synergistic; that is, the high-temperature dust removal improves the fuel processing section and the high-temperature CO₂ removal improves the fuel cell section such that the combination should work better than either of the alternatives alone.

Case 9 — High-Temperature Sulfur Removal

Several systems for high-temperature sulfur removal are under development. The approach considered here is a spinel-based system, which regenerates directly to elemental sulfur, under development at IGT (Figure 10). It

should be noted that this case is not strictly comparable to the remainder of the examples — an overall process is considered that results in elemental sulfur production. Other systems have included only the efficiency of the basic system, without the energy penalties associated with sulfur production and tail-gas treatment. Nevertheless, the system offers a process efficiency of 50.5% and an apparent cost of power of 7.2¢/kWh.

Summary

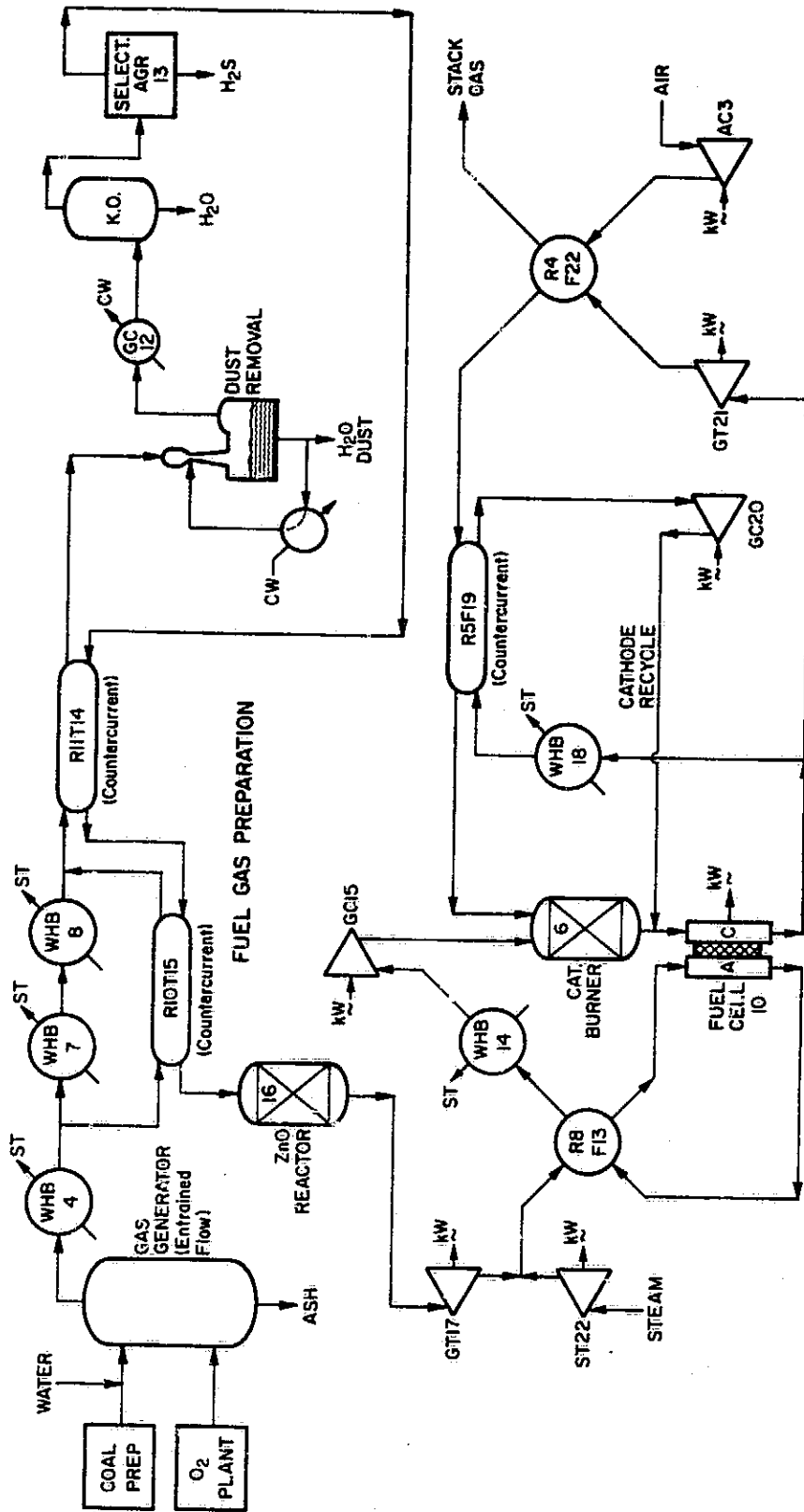
Nine systems for integration of coal gasification with molten carbonate fuel cells have been evaluated economically to assess potential for cost effective research in this application. The analysis is constrained by the high leverage of capital cost upon the cost-of-electricity and the necessary inaccuracies in the factored-costing approach utilized. The financial factors assumed for the analysis also impact the relative ranking of alternative systems. The data, however, when combined with mature engineering judgement, indicate that a number of process subsystems appear to offer merit for future development:

- Internally reforming fuel cell, chemically and thermally integrated, to be incorporated with fluidized bed gasifiers
- High-temperature desulfurization
- High-temperature dust removal
- High methane content, single-stage fluidized-bed gasifier
- Dust-tolerant shift catalyst
- High-temperature CO₂ transfer
- Sulfur-tolerant fuel cell.

Additionally, a number of minor process improvements have been identified that, although apparently desirable, have not been specifically analyzed in the economic assessment.

- Improved data base on carbon deposition phenomena
- Countercurrent heat exchangers for hazardous, high-temperature gases
- Improved catalytic combustion
- Continued systems analysis.

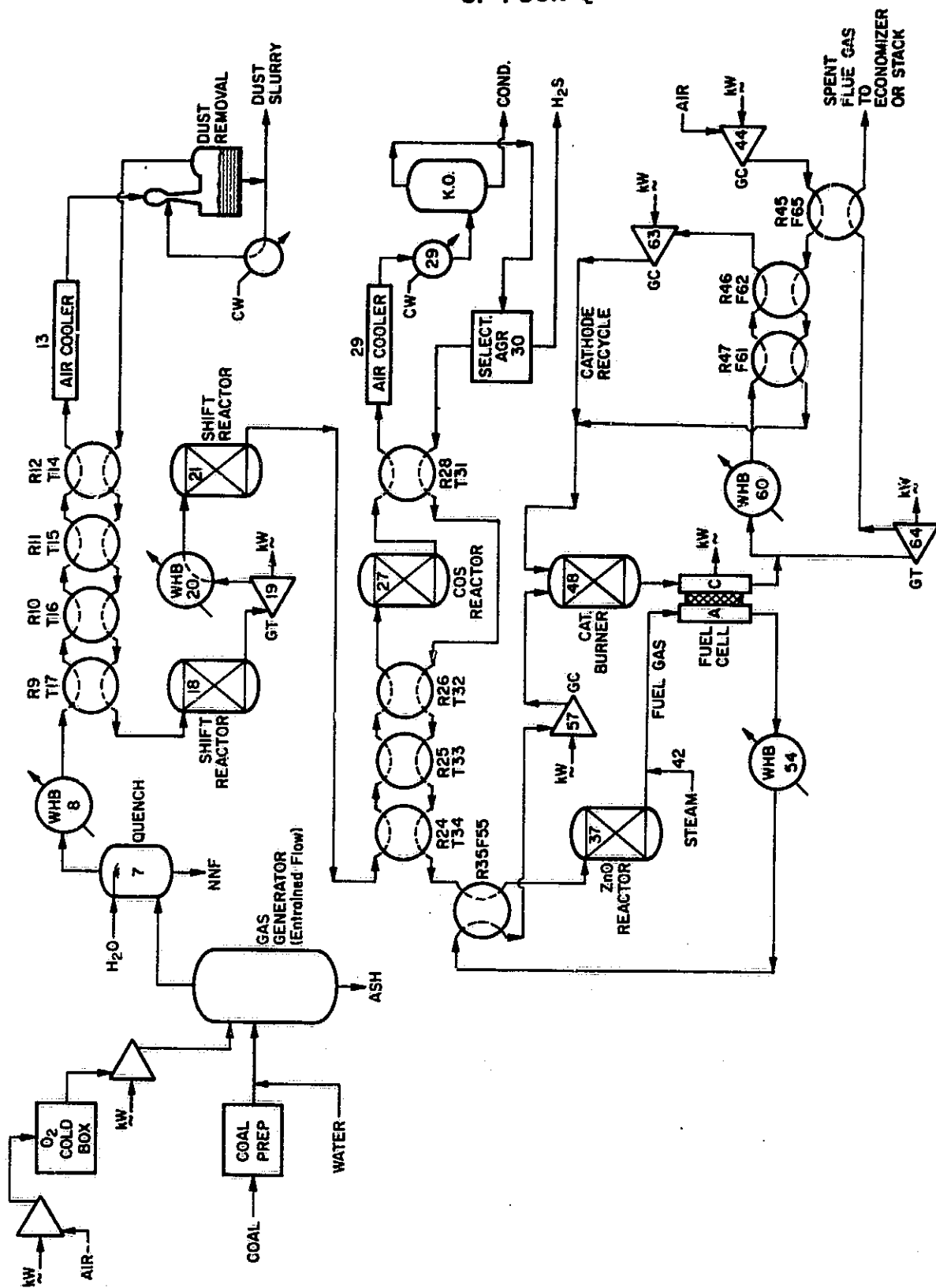
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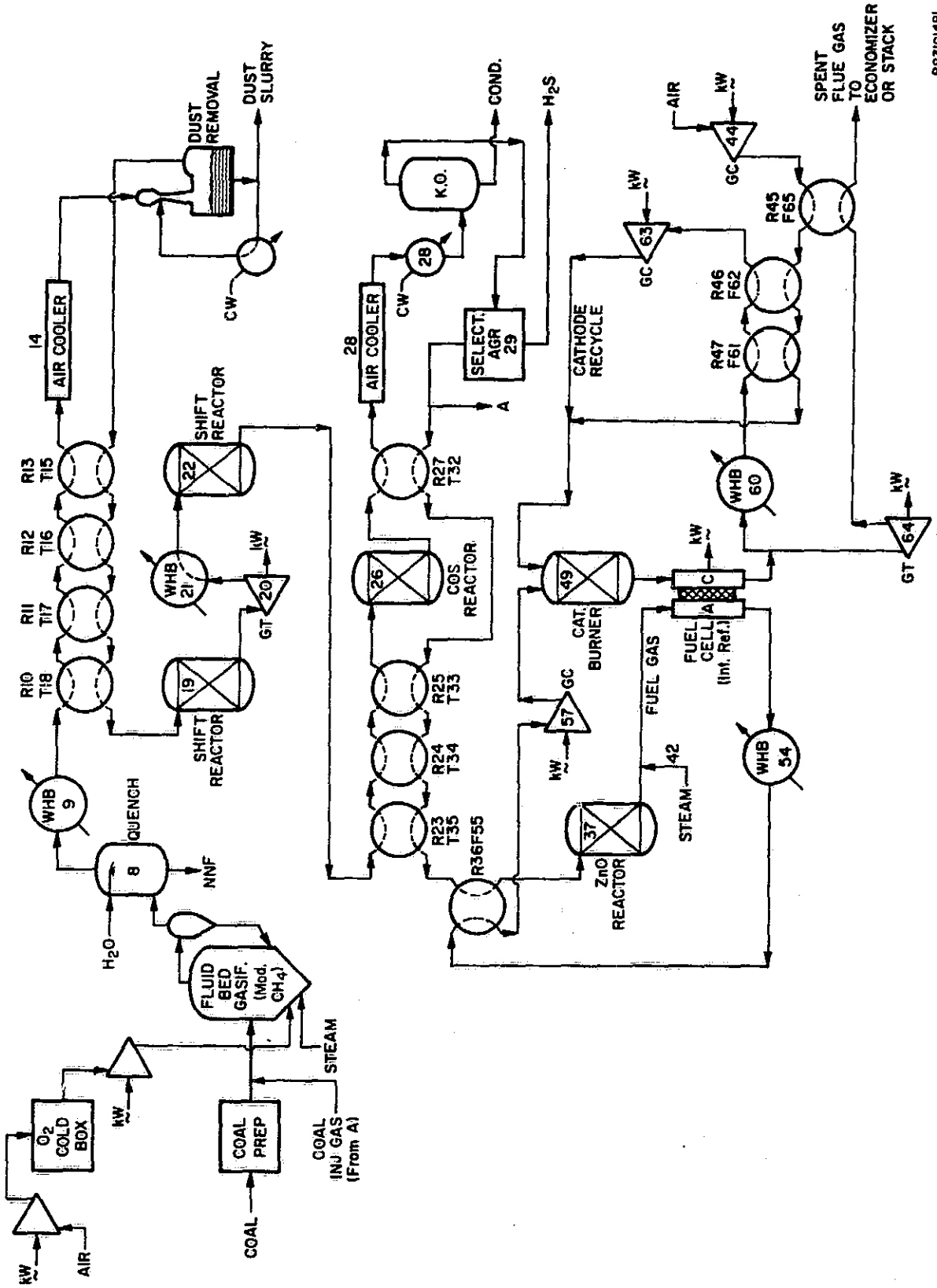
Figure 2. PROCESS FLOW DIAGRAM OF BASE CASE

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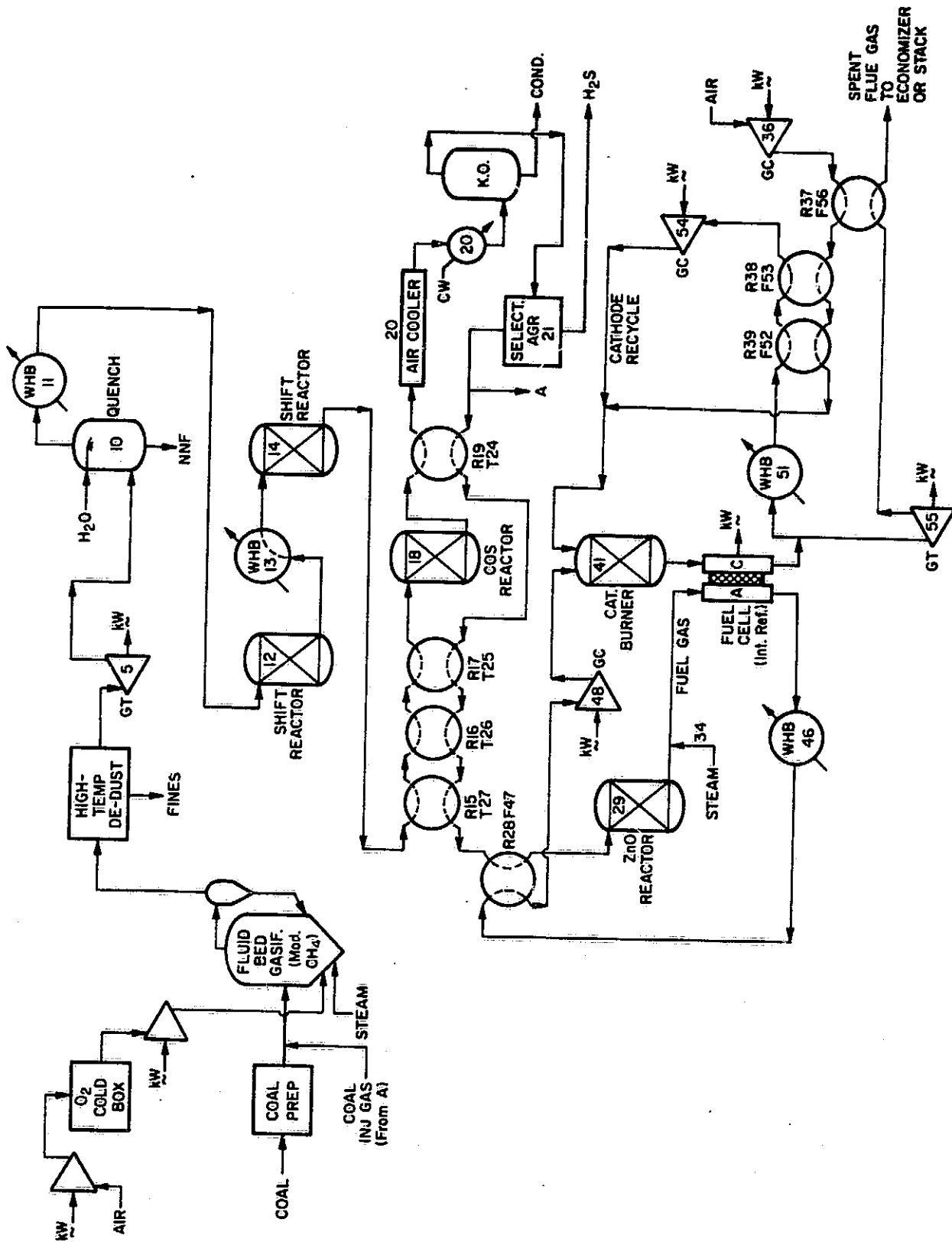
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Figure 3. PROCESS FLOW DIAGRAM OF BASE CASE AFTER REDESIGN FOR OPERABILITY



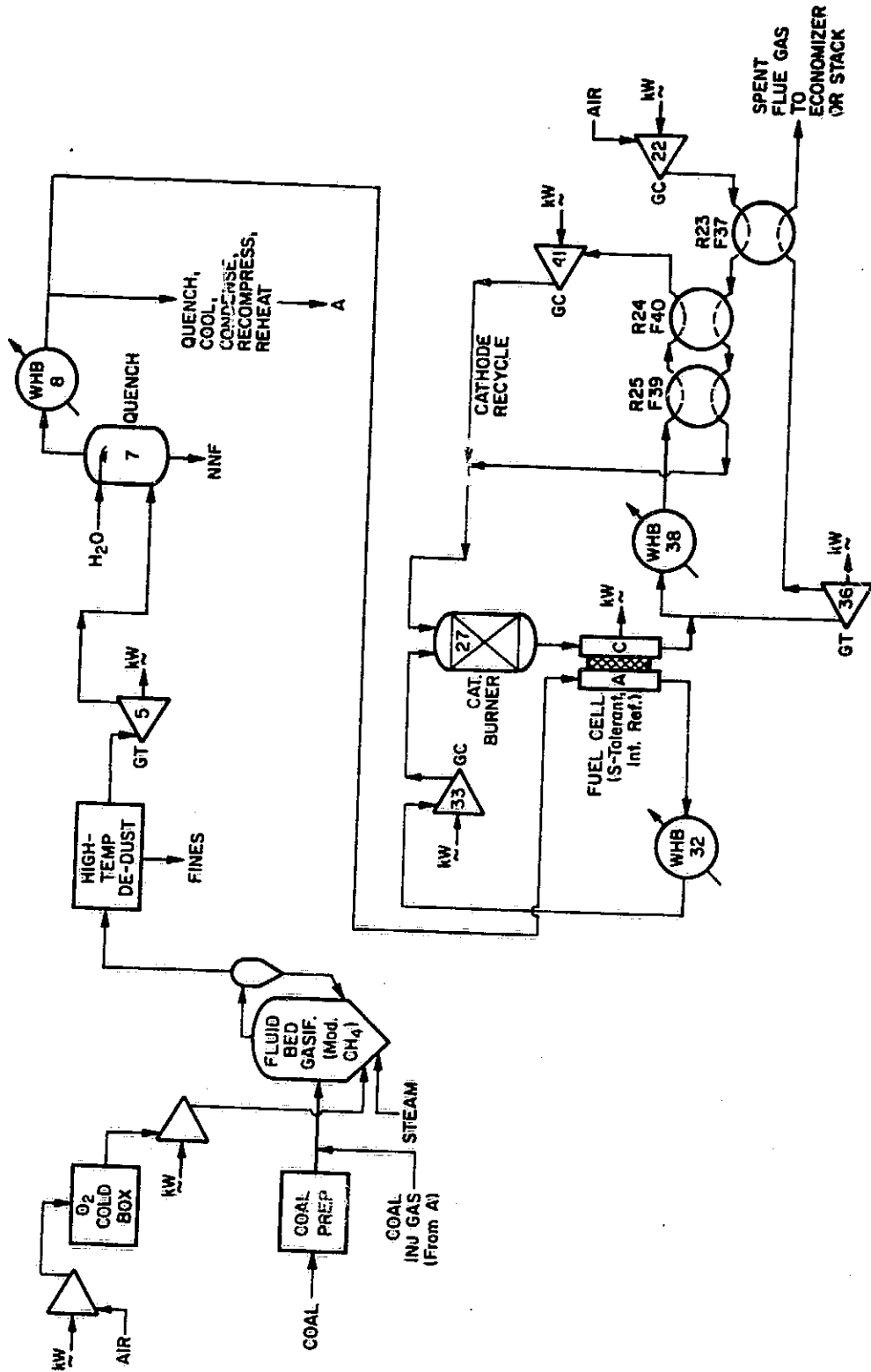
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Figure 6. PROCESS FLOW DIAGRAM OF SYSTEM WITH INTERNALLY REFORMING FUEL CELL AND FLUIDIZED-BED GASIFICATION WITH MODERATE METHANE PRODUCTION



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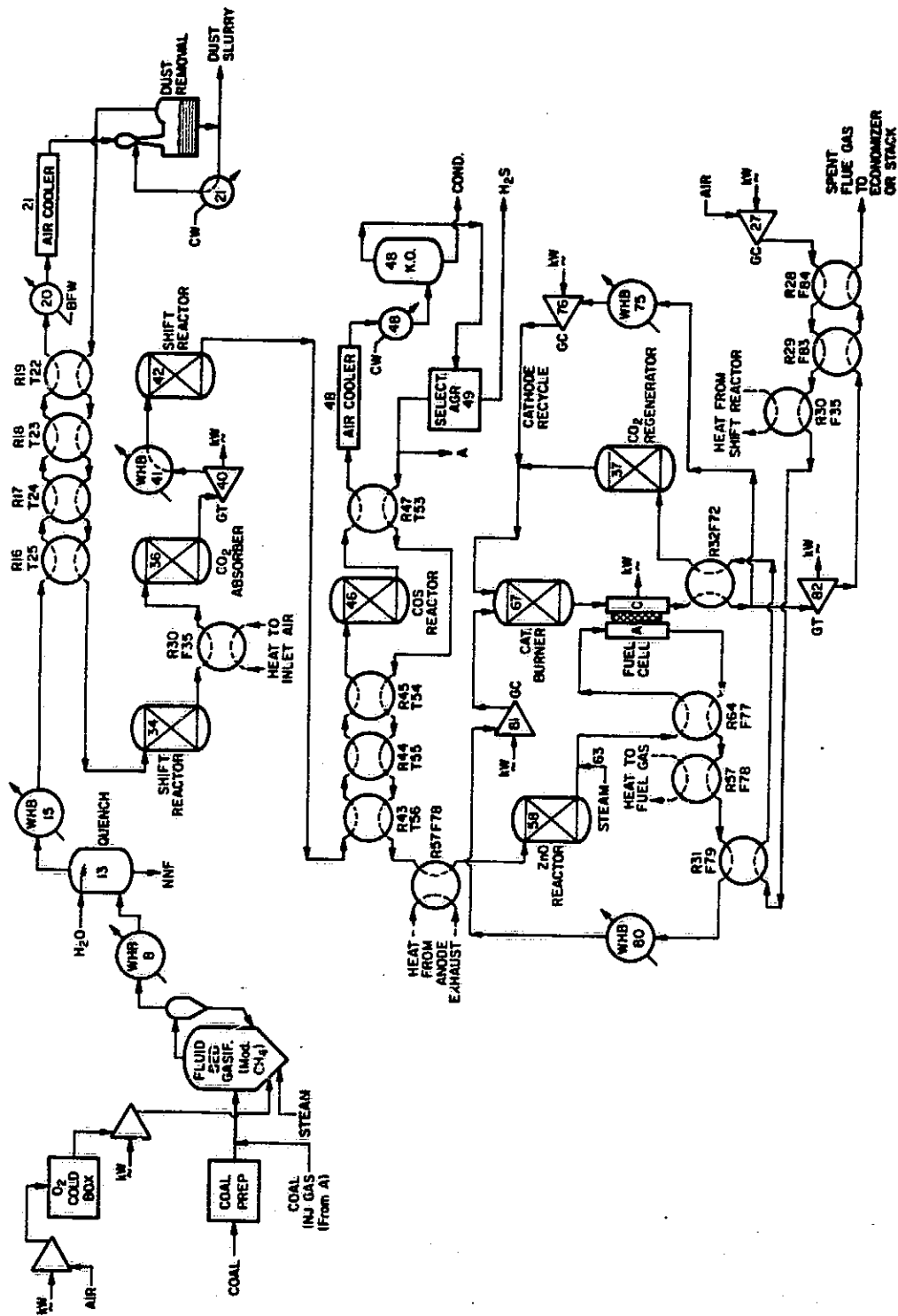
Figure 7. PROCESS FLOW DIAGRAM INCORPORATING HIGH-TEMPERATURE DUST REMOVAL



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Figure 8. PROCESS FLOW DIAGRAM INCORPORATING SULFUR-TOLERANT FUEL CELL

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Figure 9. PROCESS FLOW DIAGRAM INCORPORATING HIGH-TEMPERATURE CO₂ TRANSFER

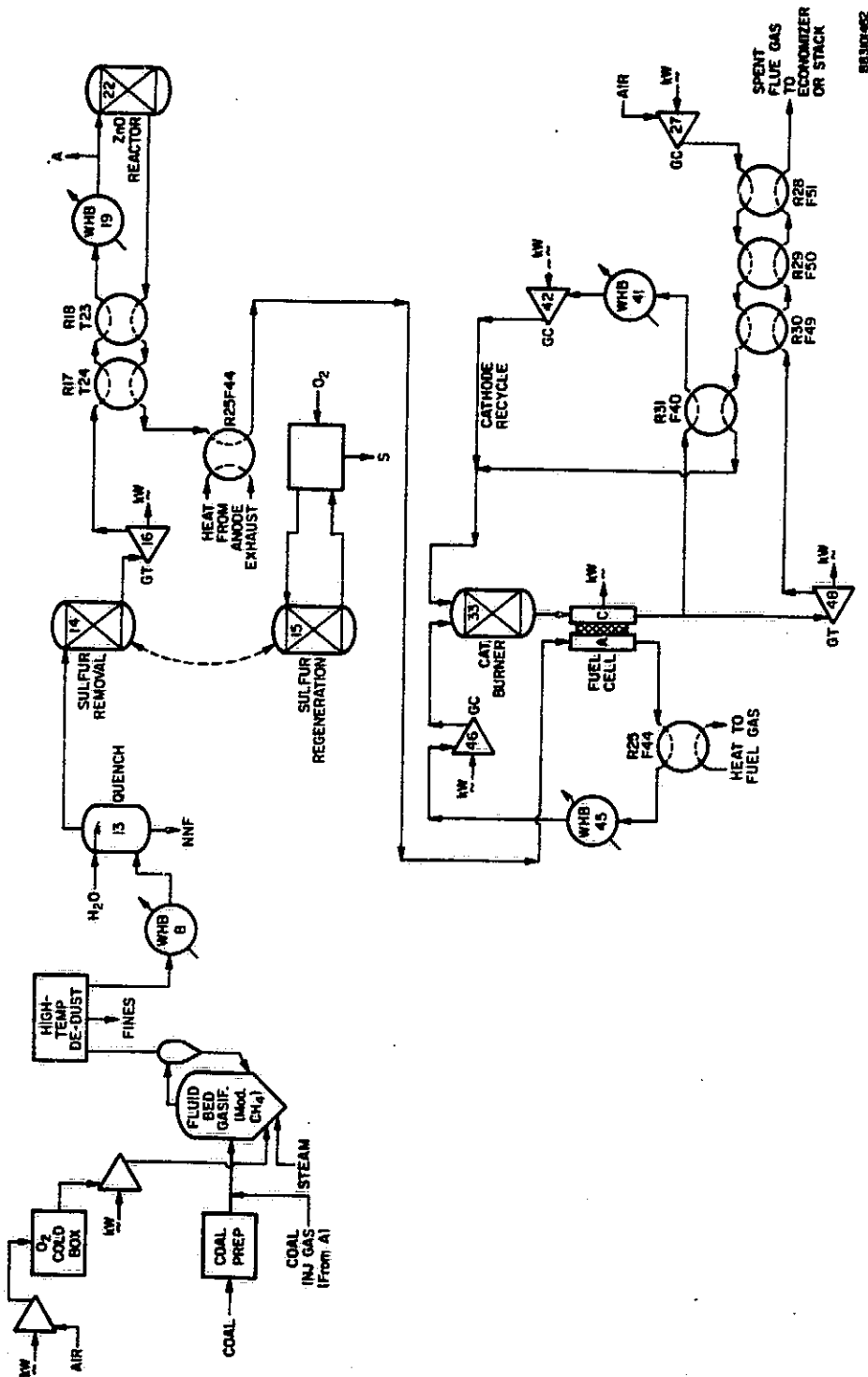


Figure 10. PROCESS FLOW DIAGRAM INCORPORATING HIGH-TEMPERATURE SULFUR REMOVAL

CONCLUSIONS

Based upon the conduct of the project and the results obtained, the following conclusions have been drawn:

- The technique of systems analysis is a cost-effective approach for comparing the efficiency of alternative systems for power production from coal using gasification and molten carbonate fuel cells.
 - In this system, the results are extremely sensitive to the process assumptions
 - a. The systems analyst must use great care to assure that two systems are identical in all respects except for that quantity being varied.
 - b. Caution must be exercised in comparing work of different investigators because of the probable difference in initial assumptions.
 - Overall system efficiencies, for approximately the same system, may vary from 46% to 57%, depending upon the initial assumptions employed.
- The cost of power is highly leveraged by capital cost, minimizing the impact of system efficiency. Additionally, the choice of financial assumptions does not tend to favor systems of higher efficiency.
 - A 3-percentage-point loss in efficiency, from 41% to 38%, is counterbalanced by only a 2.5% change in capital cost.
 - The factored costing techniques employed, although directionally correct, cannot have high accuracy.
 - Knowledgable engineering judgment must be employed in the interpretation of the relative cost data.
- Several systems were found that indicate promise for additional research and development. Such recommendations are listed in the following section.
 - The incorporation of a methane-producing fluidized-bed gasifier and an internal reforming fuel cell shows a 4.2-percentage-point gain in efficiency and a 10% reduction in capital cost when compared with a conservative analysis of the reference case.
 - The incorporation of high-temperature dust removal adds nearly 2 percentage points of process efficiency, without a capital cost penalty, because of improved placement of the fuel gas turboexpander. Additionally, this concept permits incorporation of other advanced concepts, listed below.

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- The incorporation of high-temperature sulfur removal, together with high temperature dust removal, improves the system efficiency by another 2.5 percentage points with a significant capital savings due to system simplicity, resulting in a preferred cost of power.
- A number of other subsystem elements were identified that either a) were not costed so relative value could not be determined or b) showed lesser efficiency and/or cost gains but, in total, might be significant.
 - Improved definition of the carbon deposition phenomena so that the quench system might be eliminated
 - Development of countercurrent heat exchange for high-temperature, toxic, hazardous materials
 - Development of dust-tolerant water-gas shift reactors
 - Development of a sulfur-tolerant fuel cell
 - Development of high-temperature carbon dioxide removal
 - Development of a novel, single-stage, fluidized-bed gasifier that produces greater quantities of methane.
- Optimization of a subsystem element does not necessarily result in a preferred overall system.
 - Optimization of the fluidized bed gasifier to the limits of currently available designs, compared to the conservative designs specified, did not improve the overall system efficiency.
 - Systems with highest overall efficiency had relatively poor fuel cell operating characteristics because of high back pressures of carbon dioxide and water vapor in the fuel.
- Additional systems have been conceived that are directed at minimizing the primary areas of energy loss in the system. These systems could not be evaluated within the budget constraints of the program.

RECOMMENDATIONS

The recommendations presented below are based upon engineering assessment of the direct calculated results of the program and include factors for both efficiency and cost improvement of novel processing schemes in recommending where identified R&D may be fruitful. Also, additional systems analysis work is suggested.

Identified R&D Effort

A number of mechanical and/or chemical processing suggestions appear to offer promise for improving the efficiency and, in some cases, the cost-effectiveness of production of electricity from coal through molten carbonate fuel cells. The suggestions are briefly outlined below and discussed in detail within the report. The suggestions below are not prioritized.

Improved Definition of Carbon Deposition Phenomenon

The raw gas from the gasifier, and also the purified fuel cell feed gas, have been humidified to the extent required to render the gases thermodynamically stable and eliminate the potential for carbon deposition.

The data base underlying the carbon deposition phenomenon is relatively sparse and includes only methane cracking and the Boudouard reactions, not the combination of gases that will be experienced in such a system. Further, a potential may exist for the inhibition of carbon deposition in the presence of sulfur compounds.

Although systems have been developed that nearly overcome the efficiency penalty inherent with quenching the raw gas, these systems are cumbersome and expensive.

Cost-Effective, Countercurrent, High-Temperature Gas-Gas Heat Exchange

Although countercurrent heat exchange between combustion products and combustion air is often practiced in the power industry, it is rarely employed in chemical engineering service, particularly with hazardous, toxic, or flammable materials at elevated pressure, because of the extreme complexity, poor maintainability, and costs of available designs. The problem with such units is the thermal expansion of the shell and tube components and the available means to allow for differential expansion within the exchanger. Should such units be available for operation with toxic, explosive gases at elevated

pressure, the heat exchange within the system would be simplified, efficiency would be improved, and cost of service would be reduced.

The Combination of Methane-Producing Gasifiers and Internally Reforming Fuel Cells

A fluidized-bed gasifier, compared with an entrained flow gasifier, has lower parasitic power demands and is more efficient in converting the coal to fuel gas species rather than heat. In a conventional MCFC cell, however, the methane produced in such a gasifier cannot be readily consumed electrochemically to power. Introduction of reforming capability within the anode compartment of the cell, thermally and chemically integrated with the fuel cell operation, permits substantial improvement in overall system performance. Initial R&D efforts at internally reforming fuel cells are promising. Efforts should be accelerated, because the improved cost-of-electricity appears to be significant.

High-Methane-Content Gasifiers

Patent action has been initiated on a conceptual, single-compartment, fluidized-bed gasifier that should produce higher methane content in the fuel gas through control of the mixing patterns within the single fluidized bed. The gasifier cost should be similar to a conventional, single-stage, fluidized bed gasifier. If developed, such a gasifier would permit improved efficiency and lower cost in the overall system, when integrated with an internally reforming fuel cell.

High-Temperature Departiculation

By removing particulates from the raw gas while it is hot, the turbo-expander can be moved within the system to operate on a hotter gas and provide greater power recovery. Process efficiency can be improved with this development; cost-of-power is apparently reduced.

Dust-Tolerant Water-Gas Shift Reactors

Dust-tolerant catalytic reactors have been conceptualized and appear to offer little cost penalty relative to conventional systems. If implemented in the overall system, the process efficiency will improve; moreover, the system complexity is significantly reduced and the development appears to be cost-effective.

High-Temperature Desulfurization

The concept of high-temperature desulfurization, when used in conjunction with high-temperature particulate removal, offers improved system efficiency relative to currently available, low-temperature sulfur removal processes. The improvement in efficiency, however, is not the major benefit. Rather, this approach eliminates most of the heat exchangers, pressure vessels, and contacting towers associated with gas cooling, thus providing a significant cost reduction in the selling price of the electricity.

Sulfur-Tolerant Fuel Cell

If a sulfur-tolerant fuel cell could be developed and integrated with high-temperature departiculation described above, the overall system efficiency would be improved and the overall system significantly simplified. The cost of such a system is currently conjectural, but the benefits appear to warrant an effort in this direction.

High-Temperature CO₂ Transfer

A subsystem for transfer of carbon dioxide from the fuel gas to the cathode feed has been conceived and analyzed. The overall system efficiency improved by over 2 percentage points; however, with the high leverage of capital costs used in this analysis, the benefit did not outweigh the cost. Nevertheless, the CO₂-transfer system appears to offer merit (particularly in advanced systems that are beyond the scope of this program) and should be further evaluated.

Improved Catalytic Combustion

The operating temperature range of commercial catalysts in the combustion of the anode tail gas requires significant heat exchange equipment for temperature control. Non-noble catalysts with lower light-off temperature and higher maximum-temperature limits are required for system simplification and cost reduction.

Continued Systems Analysis

Several systems were identified, but could not be evaluated within the time and budget constraints of the program. Specific areas of evaluation include the following:

- Recycle Systems
 - Effective 100% conversion, rather than 85% conversion
 - Recover latent heat of water vapor
- Air Blown Systems
 - Save cost of oxygen system
 - Use fuel cell as heat pump
 - Incorporation of membranes
- The total system — from coal in the ground to delivered electricity
 - May be cost effective to manufacture methane in large, central, mine-mouth facilities, transport methane, and produce electricity in smaller substations
 - Relative transportation costs of coal, electricity, and gas, favor gas
 - Makes maximum use of economies of scale.

The suggested effort is discussed in greater detail below.

Improved Fuel Cell Simulator

The calculated system efficiency is sensitive to the fuel cell performance; yet the fuel cell simulator is simplified and based on out-of-date data. Additionally, the overall system is quite sensitive to the cathode off-gas temperature, but the current simulator does not separately calculate this parameter. Current, publishable fuel cell performance data should be acquired and incorporated into an improved simulator.

Identified Systems Analysis Effort

Air-Blown Systems

It was discovered during the program that the molten carbonate fuel cell is an excellent heat source for a heat engine. The theoretical Carnot efficiency from a fuel cell exhaust temperature of 1300°F approaches 69%; the maximum fuel cell efficiency, considering isentropic losses, is 79%. Thus, the heat content of the exhaust gases, when recovered by a gas turbine and a steam bottoming cycle, can approach high theoretical efficiency. Note that transferring the heat to the exhaust gases is preferred to a cathode recycle that exhausts heat directly to a steam cycle.

The system considered is to feed the gasifier with air, rather than oxygen, and allow the gas to pass through the total cycle. The gasifier off-gas would contain approximately 50% nitrogen, which would be a significant diluent and increase the size of much of the processing equipment. Additionally, it would increase the polarization of both the anode and the cathode of the fuel cell. However, significant thermal energy will be stored in this nitrogen at the fuel cell cathode exhaust, and this energy may be recovered effectively by the gas turbine bottoming cycle downstream of the fuel cell. The system appears to offer merit; however, the relative tradeoff of positive and negative features cannot be ascertained without a careful comparison on the overall process simulator.

An alternative system considers the use of an air-blown gasifier with semipermeable membrane separation of the fuel hydrogen from the remainder of the species in the gas. Conceptually, the system would consist of a gasifier, quench, heat recovery, dust-tolerant shift reaction, cooling, and sulfur removal. The high-pressure fuel gas can then be separated by a semipermeable membrane recovering a hydrogen-rich stream. The membrane would operate between the pressure of the sweetened gas and the fuel cell operating pressure for the permeated hydrogen. Depending upon the operating conditions selected for the membrane, either a moderate fraction of the hydrogen could be recovered at high purity, or a much higher fraction of the hydrogen could be recovered at lower purity. The primary impurity in the hydrogen would be carbon dioxide, which has a moderate transport rate through commercially available membrane materials. Both the permeate hydrogen and the non-permeated, low-Btu fuel gas would then be reheated through the heat exchanger train. The hydrogen would pass to the fuel cell after zinc oxide treatment; the low-Btu gas could be fired into a combined cycle system, using hot cathode off-gas as the oxidant. A significant increase in energy recovery from the exhaust gas turbine would be expected because of its much higher inlet temperature.

Note that the above system has increased merit if high-temperature membranes (which are now being developed) were available. After high-temperature sulfur removal, the hydrogen could be recovered directly at elevated temperature and passed to the fuel cell. Cell operating voltage would be higher than in the high-temperature sulfur removal case considered herein because of the cleaner fuel gas. The cost associated with the heating and cooling trains would be eliminated.

Recycle Systems

Analysis of the energy balances of the various systems considered indicates that the major losses in the system are 1) the 15% of the fuel value that is combusted directly and 2) the latent heat of vaporization of water vapor that is discharged through the cathode to the stack. A number of systems can be envisioned that would employ recycle to enhance the water vapor balance.

On the total systems basis, the primary fuel is carbon, but the fuel cell only consumes hydrogen. The carbon values are converted into hydrogen reductant by reacting with water. Because these reactions take place at elevated temperature, the reactant water must have been vaporized and the latent heat of that water vapor is lost as available energy to the system.

Carbon fuel is converted to hydrogen fuel by the steam-carbon reaction, in the gasifier, and by the water-gas shift reaction, either within an external shift reactor or within the fuel cell.* Also, additional water vapor was added to the fuel cell feed to avoid carbon deposition. All of the water added to the system -- either for the chemical reactions or to maintain carbon stability in the off-gas -- has been vaporized. That heat of vaporization is a heat debit to the system, either in raw gas cooling or in the latent heat present in the final stack from the system. That latent heat consumed feed energy that was not recoverable to product power.

By recycling the hot, wet, anode exhaust back upstream in the system, the energy penalty for water vaporization can be minimized. As discussed in Appendix Cb, the theoretical system need not require water make-up or discharge; rather, it may be in water balance. By recycling the wet anode exhaust upstream, all of the hydrogen value required for the fuel cell operation can be maintained in the recycle loop, without adding heat to vaporize water. Additionally, all of the fuel value in the raw gas is eventually converted to electricity in the fuel cell, without burning the anode exhaust into the Carnot-limited steam cycle.

* See Appendix Cb for more complete discussion.

Case X of Appendix Ca was an initial attempt at a recycle system. In this case, the fuel cell was fed with dry, nearly CO₂-free gas by recycle of the spent anode tail gas back to the feed of the acid-gas removal system. The system permitted 100% recovery of the hydrogen value in the fuel within the fuel cell and eliminated the need to add steam to the fuel cell feed, thus minimizing some of the heat losses from the system. Although that system had high fuel cell voltage and power recovery, the approach was found not to be a preferred alternative.

Other points for reinjection of the spent anode exhaust would result in further minimization of the water demand. The gas could be fed directly into the raw off-gas of the gasifier, displacing the steam required for quenching. Alternatively, the spent anode off-gas could be fed directly into the gasifier itself, displacing all of the steam or water required for the gasifier or the quench system. The chemistry discussed in Appendix Cb illustrates that this approach is feasible; preliminary calculations indicate that it is almost possible, although a small bleed stream must be taken from the recirculating gas loop to eliminate nitrogen build-up from the system.

Either approach would require removal of carbon dioxide from the recirculating loop and, preferably, would require the operation of the fuel cell at the gasifier operating pressure. In order to maintain the water in the vapor phase in the recirculating loop, a high-temperature CO₂ recovery system is required. The proper location for this recovery system is on the anode discharge, where the partial pressure of carbon dioxide would be the greatest (assuming that the gasifier and fuel cell operate at the same pressure). Sufficient carbon dioxide could be transferred from the recirculating anode loop to the cathode feed to close the carbon dioxide balance of the system.

Preferably, the fuel cell should operate at the pressure of the gasifier. The volume of tail gas from the anode is much greater than the feed; therefore, expansion and recompression would be uneconomical. Conversely, the volume of the cathode exhaust would be similar to its feed, so the penalty of air compression should not be severe. One of the reasons that improved recycle systems were not considered in the current program was the nonapplicability of the current fuel cell simulator to high-pressure fuel cell operation. The simulator, although relatively accurate at 150 psi, showed significant errors at lower operating pressures. The extrapolation to higher

operating pressure would, therefore, be tenuous. Part of the required program would be the development and evaluation of a fuel cell simulator that would operate well at elevated pressures.

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APPENDIX Aa. Reference Case

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**BASE CASE REFERENCE DESIGN
AND ECONOMIC ASSUMPTIONS
FOR A COAL-BASED
FUEL CELL POWER PLANT**

**Jet Propulsion Laboratory
4800 Oak Grove Drive
Pasadena, California 91109**

August, 1982

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I. Description of Base Case Design

The base case, as shown in Figure 1 is a relatively simple two pressure level configuration incorporating a Texaco gasifier and gas clean-up system operating at 600 psia and a fuel cell system operating at 100 psia. Data on the composition, pressure, and temperature of designated streams is given in Table 1. The salient feature of this system is the injection of low pressure steam into the anode fuel gas to create a carbon-free equilibrium composition prior to entry into the fuel cell.

The Texaco gasifier is an entrained bed gasifier fed with a coal/water slurry and oxygen producing a high temperature fuel gas. A performance summary of the Texaco gasifier is given in Table 2. Product gas at 2450°F leaves the gasifier and passes through the heat recovery steam generator. Downstream of the steam generator, the raw gas flow is divided into two streams, one passing through a steam generator heat exchanger and the other passing through a regenerative heat exchanger for reheating clean gas. These heat exchangers are followed by a wet particulate scrubber in which the gas enters at a temperature slightly above its dew point. Following the scrubber is a second parallel arrangement of steam generators and regenerative heat exchangers, followed by a gas cooler, knock-out drum, and the acid gas removal subsystem.

Clean fuel gas leaving the acid gas removal subsystem is regeneratively reheated to a temperature of 750°F and passed through a ZnO polishing scrubber to remove any residual sulfur compounds. The gas is then admitted to an expander turbine in which pressure is reduced to the fuel cell pressure. Anode inlet gas is conditioned to a carbon-free equilibrium composition by mixing it with steam extracted from the steam turbine and then it is heated by the anode outlet gas stream. The anode discharge stream is mixed

Table 1

BASE CASE FLOW STREAM GAS COMPOSITIONS

	1	2	2'	3	3'	4	5	6	7	6	9	10
H ₂ MF	0.288				↑	0.361		0.360	0.034			
CO MF	0.425				↑	0.532		0.116	0.017			
CH ₄ MF	0.001				↑	0.001		0.0005	0.0003			
C ₂ H ₄ MF												
C ₂ H ₆ MF												
CO ₂ MF	0.087				↑	0.096		0.219	0.509	0.110		0.137
H ₂ O MF	0.179				↑		1.00	0.300	0.436	0.218		0.207
N ₂ MF	0.008				↑	0.010		0.005	0.004	0.594	0.790	0.565
H ₂ S MF	0.010				↑							
COS MF	0.001				↑							
O ₂ MF										0.079	0.210	0.091
NH ₃ MF	0.002				↑							
Total Moles/ Mole Raw Gas	1	0.79	0.21	0.73	0.27	0.80	0.70	1.50	2.10	4.55	3.41	19.12
Pressure PSIA	600	800	800	330	330	75	600	100	860	1300	14.7	1120
Temp (°F)	2450	800	800	330	330	75	600	1000	860	1300	60	1120

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Table 2
TEXACO GASIFIER PERFORMANCE SUMMARY

GASIFIER PRODUCT GAS COMPOSITION (Mole Fraction)

CH ₄	0.0008
H ₂	0.2884
CO	0.4245
CO ₂	0.0871
H ₂ S	0.0101
COS	-0.0006
N ₂	0.0078
H ₂ O	0.1788
NH ₃	0.0019

FEEDS

O₂: 0.84 Lb/Lb Coal

Water: 0.44 Lb/Lb Coal

Coal/Water Slurry Feed Temperature 140°F

Oxygen Feed Temperature 300°F

Product Gas Temperature 2450°F

Moles Product Gas/Lb Coal 0.10809

HHV (CO + H₂ + CH₄) HHV Coal .772

with air from the bottoming cycle gas turbine compressor and passed through a catalytic burner, to combust the remaining H_2 , CO and any CH_4 . The burner discharge stream is mixed with additional air and cooled recycled cathode discharge gas before it is admitted to the cathode. The cathode discharge stream, after being partially recirculated through a steam generator heat exchanger for removal of fuel cell waste heat, is admitted to the bottoming cycle gas turbine. Downstream of this turbine is a compressor discharge preheater and an economizer heat exchanger for recovery of exhaust heat ahead of the stack. Steam generation for a 2400 psi/1000°F/1000°F reheat steam bottoming cycle is accomplished in the economizer (feedwater heating), the gasifier heat recovery steam generator (final feedwater heating and evaporation) and in the bottoming cycle heat exchanger (superheat and reheat) which cools the fuel cell cathode recirculation stream.

The gasifier heat recovery steam generator heat exchanger, a standard feature of the Texaco gasification system is a combination radiant/convective unit with slag solidification and separation in the radiant portion and further gas cooling in the convective section. Tube metal temperatures are maintained below 800°F by restricting the heat exchanger duty to final feedwater heating and evaporation. Low pressure steam for stripping use in the acid gas removal process is generated in the secondary heat exchangers operating in parallel with the clean gas regenerative heat exchangers.

Air discharge flow from the gas turbine compressor is preheated by heat exchange with the expansion turbine discharge in order to increase the amount of high level heat available to the steam cycle through the cathode recirculation loop heat exchanger.

The fuel cell anode operates at 0.85 utilization of the CO and H₂ in the fuel gas. Unreacted CO and H₂ in the anode exhaust, together with a small amount of methane generated in the gasifier which passes through the anode, are burned in the catalytic combustor. Most of the CO₂ required by the cathode is generated by electrochemical conversion of the fuel in the anode, however, small contributions are produced in the gasifier and in the catalytic combustor.

The gasifier pressure level of this system (600 psi) is that of an established Texaco design. It is also a level which is suitable for relatively low cost heat exchangers in the heat recovery train, and for a relatively low cost physical absorption type of acid gas removal process. Power recovery in the fuel gas expander turbine, in which pressure is reduced to the level of the fuel cell, exceeds the power required by the intercooled oxygen compressors in the oxidant supply to the gasifier.

The fuel cell pressure level (100 psi) has been selected at an approximately optimized value for maximum power output from the bottoming cycle. Similarly, the overall excess air ratio (100%) has been chosen to maximize the combined output of the gas and steam turbine generators.

Design requirements for the acid gas removal process specify a reduction of sulfur bearing gas species to a level less than 1 ppm (0.1 ppm), and that the loss of CO₂ from the clean gas stream not exceed 12%. It is also desirable that stripping steam requirements be low in order to minimize or avoid the need for turbine extraction steam to supplement that which can be generated from the lower temperature level heat sources in the heat recovery

train. A single stage Selexol process appears to be capable of meeting these requirements for the conditions of this cycle. Some problem with COS removal may exist, but if necessary, final COS removal can be handled in the ZnO polishing units placed downstream of the clean gas regenerative heat exchangers.

An overall energy balance is given in Table 3. Although this system provides a modest fuel cell output ratio of 0.30, it does achieve an excellent ratio of waste heat conversion in the gas and steam turbine bottoming cycle. Thus, very little low level heat is uselessly rejected in the cooling train. The net efficiency* of this system is 46.7%, which is considered satisfactory for a near term plant and would probably be increased by advancements in molten carbonate fuel cell technology.

Compared to other alternative cycles, this system configuration has an attractive simplicity. In addition, the gas composition at the fuel cell anode inlet is similar to that of fuel gases that have been used in laboratory testing of molten carbonate fuel cells. Thus, uncertainties relative to carbon and methane formation and relative to the accuracy of fuel cell voltage predictions are minimized.

The principal disadvantage of this system is the calculated fuel cell DC voltage of 0.69 volts (at the design current density of 160 mA/cm²). Fuel cell voltage this low may result in electrochemical corrosion phenomena within the fuel cell stack. Although improvements in molten carbonate fuel cell

*Thermal cycle efficiency, coal pile to busbar.

Table 3.

GAS STREAM ENERGY BALANCE AND SYSTEM PERFORMANCE SUMMARY

Gas Stream Energy Balance		
Energy Inputs	Energy Input	Energy Output
Coal HHV Input	1.0	
Coal/H ₂ O Oxidant Sensible Energy	0.008	
Injection Steam Energy	0.144	
Plant Compressors	0.02	
Total Energy Input to Gas Stream	1.172	
<u>Energy Outputs</u>		
Fuel Cell DC Output		0.305
Gas Turbine Shaft Output		0.121
Heat Delivered to Steam Turbine Cycle		0.450
Heat Available for Stripping Steam Generation		0.011
Stack Loss		0.223
Low Level Heat Rejection in Gas Heat Recovery Train		0.016
Residual Heat Loss		0.046
Total energy Output		1.172

Performance Summary	
Electric Generation Outputs	
Fuel Cell AC Output	0.299
Gas Turbine Generation Output	0.119
Steam Turbine Generation Output	0.133
Gross Electric Output	0.551
Auxiliary Power Requirements	
Oxygen Plant	0.034
Oxygen Compressors	0.015
Plant Gas Compressors	0.02
Other Auxiliary Power	0.015
Total Auxiliary Power	0.084

Plant Efficiency = 0.551 - 0.084 = 0.467

Fuel Cell DC Voltage: 0.69

performance will probably ameliorate this concern in the future, the present uncertainty in predicting the voltage level at which this electrochemical corrosion will occur provides impetus for developing system configurations that result in higher fuel cell voltages than are currently predicted for this system configuration.

II. Technical Assumptions Used in Base Case Design

For the Base Case the following assumptions apply:

1. Fuel Stack Performance

o Fuel Cell Temperatures

The anode inlet temperature is 1000°F. Cathode outlet temperature is 1300°F for this system and the cathode inlet temperature is selected to provide an anode outlet temperature under counterflow conditions of 1150°F.

o Carbon Formation Criterion

The anode inlet gas composition satisfies the criterion of freedom from carbon deposition under the condition of complete C-H-O equilibrium.

o Methane Formation

It is assumed that the kinetics of the methanation and reforming reactions are sufficiently slow inside the anode stream of the fuel cell under load that methane can be treated as an inert species for purposes of determining gas composition variation through the cell anode.

o Fuel Cell Average Current Density

The design value of the fuel cell stack average current density is 160 mA/cm².

o Calculation Procedure for Cell Terminal Voltage

Cell terminal voltage is calculated by use of the fuel cell nodal computer program described in EPRI Report EM-1670. For

the Base Case, the fuel cell was represented by 10 segments. Nernst voltage is calculated for each node on the basis of average nodal anode and cathode gas concentration and the local temperature. Terminal voltage is determined by an iterative procedure which establishes the current density distribution such that the difference between local Nernst voltage and the product of the polarization constant and local current density is equal for all segments. The program is capable of handling counterflow, coflow or crossflow relationships between the anode and the cathode gas streams.

o Cell Polarization Factor

The cell polarization constant for use in the calculation of cell voltage has been determined from an empirical correlation of state-of-the-art MCFC test data.

o Fuel Cell Stack Inverter

A D.C. to A.C. power conversion efficiency of 0.98 is assumed.

2. Plant Auxiliary Power Requirements

Auxiliary power requirements are assumed to consist of the following:

o Oxygen Plant

Oxygen plant power is assumed to be 108% of the power for electrically driven high efficiency air separation process compressors of 6.8 pressure ratio.

o Oxygen Compressors

Electrically driven high efficiency compressors with three stages of intercooling are assumed.

o Plant Gas Recirculation Compressors

High efficiency electrically driven compressors are assumed.

- o Additional auxiliary power for acid gas removal pumps and compressors, coal handling equipment, gasifier auxiliaries, etc., is assumed to equal 1.5 percent of the coal higher heating value (HHV) energy (approximately 3% of the plant gross output).

3. Gas Turbomachinery Components

o Compressors

Efficiency is assumed to be 0.88.

o Turbines

Efficiency is assumed to be 0.92.

4. Electrical Machines

All motors and generators are assumed to have an efficiency of 0.98.

5. Acid Gas Removal Processes

o Gas Purification

Processes are assumed to remove all sulfur bearing species and HCN to less than 1 ppm.

o Stripping Steam Requirements

A minimum requirement of 0.13 lbs of 50 psig saturated steam per pound of coal was assumed. This assumes the combined use of steam and N₂ for stripping. This corresponds to an energy requirement of 0.012 of the coal HHV, and 80% of this energy must be available above a temperature level of approximately 330°F. For systems similar to the Base Case, the 50 psia steam energy available may be lower than the stripping steam requirements. This would require the excess to be supplied by extraction from the steam turbine, at a cost per pound of coal of approximately 2 percentage points system performance loss per pound of steam required.

o CO₂ Recovery

The Base Case employs a one stage acid gas removal process and 88% CO₂ recovery is assumed.

III. Economic Analysis Assumptions & Methods to be Used

1. Capital costs include the following basic components:

- o TOTAL PLANT COST (TPC)
 - Process (On-Site) Capital
 - General Facilities (Off-Site) Capital
 - Engineering and Home Office Overhead Including Fee
 - Process Contingency
 - Project Contingency

- o TOTAL CAPITAL REQUIREMENT (TCR)
 - Total Plant Investment
 - Prepaid Royalties
 - Preproduction (Start-up) Costs
 - Inventory Capital
 - Initial Chemical and Catalyst Charge
 - Allowance for Funds During Construction
 - Land

Process capital is the total constructed cost of all on-site fuel processing and power generating units including all direct and indirect construction costs. Capital costs for fuel processing items can be scaled from previous literature sources or in-house data, according to the following rule:

$$\text{Cost Unit A} = \text{Cost Unit B} \times \left(\frac{\text{Size Unit A}}{\text{Size Unit B}} \right)^C$$

Scaling exponent C and cost basis for several fuel processing units are given as follows:

<u>Fuel Processing Unit</u>	<u>Cost Base</u>	<u>Scaling Capacity Factor</u>	<u>Scaling Exponent C</u>
Oxygen Plant	AP-642*	Ton/Day O ₂	0.56
Coal Storage & Reclaiming/ Handling & Preparation	AP-1543*	Ton/Day Coal	0.78
Gasification & Ash Handling	AP-1543*	Ton/Day Coal	0.60
Gasifier Gas Cooling		Gas Molar Flow	0.82
Gas Cooling & Ammonia Scrubbing		Gas Molar Flow	0.67
Acid Gas Removal	AP-1543*	Gas Molar Flow	0.60
ZnO Unit		Gas Molar Flow	0.65
Sulfur Recovery	AP-1543*	Gas Molar Flow	0.60

The capital cost of the fuel cell should be estimated as if it were a "mature" technology; i.e., the cost would be that of the "nth" fuel cell unit and not the first, second, etc., unit ever to be built.

General Facilities Capital includes roads, buildings, shops, laboratories, etc., and are generally in the range of 5-20% of the Process Capital Cost. For this analysis, 10% should be used.

Engineering and Home Office Overhead Including Fees is generally about 10-15% of the process capital. For this analysis, 10% should be used.

* Report published by Electric Power Research Institute, Palo Alto, California.

Process contingencies include costs added to Process Plant Investment for improved technology. Process Contingencies for those units considered improved are as follows:

Gasification - commercial units operated	5%
- 100-1000 TPD plant operated	15%
- 10-100 TPD pilot plant operated	25%
- concept with bench scale data	50%
- new concept with limited data	60%
Ash Handling	5%
High Temperature Gas Cooling	15%
High Temperature Sulfur Cleanup	50%
Tail Gas Treating	5%
Molten Carbonate Fuel Cell (Base Case)	50%
Molten Carbonate Internal Reforming Fuel Cell	60%

Other process contingencies can be coded as follows:

<u>State of Technology Development</u>	<u>Percentage of Installed Section Cost</u>
New concept with limited data	60%
Concept with bench-scale data available	50%
Small pilot plant data (e.g., 1 MW) available	25%
A full size module has been operated (e.g., 10-100 MW)	15%
Process used commercially with different feed or application	5%

Project Contingency includes costs added to the Total Plant Cost that would result from a more definitive design. A 30% Project Contingency cost should be used for this analysis.

Prepaid royalties include process royalty fees; a value of 0.5% of the Process Capital (excluding contingencies) should be used for this analysis.

Preproduction, or Start-up Costs, includes the following:

- a) One month fixed operating costs (operating and maintenance labor, administrative and support labor, and maintenance materials);
- b) one month of variable costs (consumables) at full capacity excluding fuel;
- c) twenty-five percent of full capacity fuel cost for one month;
- d) two percent of the TPC.

Inventory Capital includes the value of 60 days inventory of fuels, other consumables, and by-products at full load (100% capacity).

Initial Chemical and Catalyst Charge Costs include the initial cost of the charge of catalysts or chemicals contained within the process equipment.

Allowance for Funds During Construction is based on the annual interest rate with annual end-of-year compounding and the following construction expenditure schedule:

<u>Year</u>	<u>Percent of Total Plant Cost</u>
1	33 1/3
2	33 1/3
3	33 1/3

The capital estimates should be based on the following:

<u>Year</u>	<u>Mid-1981</u>
Debt/Equity Ratio	50/50
Debt Cost	8%/Yr.
Preferred Stock Ratio	15%
Preferred Stock Cost	11%/Yr.
Common Stock Ratio	35%
Common Stock Cost	15.3%/Yr.
Discount Rate	12.5%/Yr.
Federal and State Income Tax Rate	50%
Property Taxes and Insurance	2%/Yr.
Investment Tax Credit	10%
Book Life	30 Yrs.
Tax Life	15 Yrs.
Inflation Rate	8.5%/Yr.
Real Escalation Rate	0%/Yr.
Total Plant Capacity	675 MW

2. Operating costs are based on the following components:

- o FIXED
 - Operating Labor
 - Maintenance Costs
 - Overhead Charges

- o VARIABLE
 - Consumables
 - o Fuel
 - o Water
 - o Ash Disposal
 - o Catalysts and Chemicals

 - Variable Maintenance Charges (if any)

- o FUEL CELL REPLACEMENT EXPENSE (40,000 HR LIFE)

An average labor rate(ALR) of \$15.25/person hour should be used . This ALR includes payroll burden.

Annual maintenance costs were estimated as a percentage of the installed capital cost of each plant section. The percentage applied varies with the nature of the processing conditions and the type of design. The following percentages were used in this study as shown in Table 4.

The maintenance costs can be broken down and expressed as maintenance labor and maintenance materials. A maintenance labor/materials ratio of 40/60 can be used for this breakdown.

Table 4. MAINTENANCE COSTS

<u>Process Unit Function</u>	<u>Percent/year. of Installed Plant Section Cost</u>
Oxygen Production	2.0
Coal Storage & Preparation	3.0
Gasification & Ash Handling	4.5
Gas Processing	
A. Gas Cooling	3.0
B. Gas Cooling & Ammonia Scrubbing	3.0
C. Gas Reheat	2.0
Acid Gas Removal	
A. Selexol	2.0
B. Trim ZnO Unit	0.5
Sulfur Recovery & Tail Gas Treating	2.0
Waste Water Treating	3.0
Steam Generation	1.5
Fuel Gas Expansion & Air Compression	3.0
Process Condensate Treating	3.0
Steam, Condensate and BFW	1.5
Support Facilities	1.5
Fuel Cell Combined Cycles	2.0
Fuel Cell Modules	0
Fuel Cell Inverters	1.0

The Overhead Charges is for administrative and support labor; 30% of the operating and maintenance labor should be used.

The variable operating costs are based on a 65% yearly utilization of the total plant capacity and are composed of the following charges:

- o Fuel (Coal) Cost
- o Raw Water
- o Ash Disposal
- o Catalysts and Chemicals

These items are discussed below.

The 1981 delivered coal cost is \$1.65/10⁶ Btu (HHV) for an Illinois No. 6 coal with analyses as given in Table 5. Coal costs are expected to escalate in real year terms by 0.7%/year (apparent escalation of 9.3% including 8.5% inflation).

The 1981 raw water acquisition cost to be used is 50¢/1000 gallons. Intake structures, treating costs and pumping costs are to be included in the process capital and operating and maintenance charges.

The catalyst, chemicals and other consumable costs should be estimated from literature, in-house, or vendor sources.

The 1981 ash disposal charge to be used is \$5/ton.

Coal analysis is shown in Table 5 for an Illinois No. 6 coal.

Table 5
COAL ANALYSIS

<u>Type</u>	<u>Illinois No. 6</u>
PROXIMATE ANALYSIS (Wt %)	
Moisture	4.2
Ash	9.6
Fixed Carbon	52.0
Volatile Matter	<u>34.2</u>
	100.0
ULTIMATE ANALYSIS - DAF COAL (Wt %)	
Carbon	77.26
Hydrogen	5.92
Oxygen	11.14
Nitrogen	1.39
Sulfur	4.29
Other	<u>-</u>
	100.00
HEATING VALUE - AS RECEIVED	
High Heating Value (HHV) (Btu/lb)	12,235
Net Heating Value (LHV) (Btu/lb)	11,709

3. Economic Analysis Method

The objective of the study is to identify R&D activities that have the potential for making significant improvements over the Base Case Reference Design . Improvements should be quantified, for purposes of this study, as reductions in capital cost (\$/kW), reductions in operating costs (\$/year), and reductions in cost of electricity (COE; \$ /kWh). Capital and operating cost reductions should be calculated using the factors given above. COE reductions should be calculated using the revenue requirement method as explained in EPRI's 1982 Technical Assessment Guide (Report No. P-2410-SR). More specifically, the COE should be calculated using the 10-year, short-term, levelized analyses.

Potential economic advantages of alternatives over the Base Case Reference Design should then be compared to estimated R&D costs required to utilize the alternative in a coal based fuel cell plant.

APPENDIX Ab. Economic Report for Reference Case

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ECONOMICS FOR
BASE CASE REFERENCE DESIGN

Jet Propulsion Laboratory
4800 Oak Grove Drive
Pasadena, California 91109

October 1982

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INTRODUCTION

The purpose of this report is to provide a standardized format for calculating the Cost of Electricity (COE). The report includes a Summary Section, which highlights the major capital and operating costs, plus Sections 1 through 5 which give further details on the calculation of capital costs, operating costs, and the COE calculation method. An Appendix section is included to give details of how units within the Reference Design were scaled and how the constructed costs were estimated.

Many of the economic assumptions were taken from EPRI's Technical Assessment Guide (TAG), Report No. P-2410-SR.

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SUMMARY

Calculation of the Cost of Electricity (COE) includes estimating the Total Capital Requirement, the Variable Operating Costs, the Fixed Operating Costs, and the Fuel Cell Replacement Expense.

The Total Capital Requirement can be broken into seven major elements as follows:

	<u>\$MM</u>
1.1 <u>Total Plant Cost (TPC)</u>	
A. Process (On-Site) Costs	547.73
B. General Facilities Total Constructed Cost - 10% of Process Capital -	54.77
C. Engineering & Home Office Overhead plus Fee - 10% of Process Capital -	54.77
D. Process Contingency - composite by process unit and state of art -	90.9
E. Project Contingency - 30% of Total Plant Cost - (A + B + C) x 0.3	197.18
F. Total	945.35
1.2 <u>Allowance for Funds During Construction (AFDC)</u>	
AFDC = TPC	35.28
(see TAG B-5-IV-1) = 945.35 (0.0373)	
1.3 <u>Prepaid Royalties</u>	
= 0.5% (TPC - Gen'l Facilities - Project Cont. on Gen'l. Facilities)4.37	
= 0.005 (945.35 - 54.77 - 0.3 x 54.77)	

1.4 Preproduction or Start Up Costs

1 month fixed oper. costs	24.350 MM/yr	- 12	=	2.03 MM
1 month variable oper. costs	47.67 MM/yr	- 0.65 - 12	=	6.1 MM
1 month capacity fuel x 0.25	46.35 MM/yr	- 0.65 - 12 x 0.25	=	1.5 MM
2% of TPC	945.35	x 0.02	=	18.9 MM
Total				28.53

1.5 Inventory Capital

60 days inventory of fuel, catalyst, and chemicals @ full capacity

$$\text{coal} = \frac{46.35}{0.65} \times \frac{60}{365} = \$11.7 \text{ MM}$$

$$\text{Catalyst \& Chemicals} = \frac{0.31}{0.65} \times \frac{60}{365} = 0.08 \text{ MM}$$

Total 11.78

1.6 Initial Catalyst & Chemical Charge

Basis: Prorate based on AP-1543

$$\frac{\text{Initial Cat \& Chem}}{\text{Annual Cat \& Chem}} = \frac{2451}{455} = 5.4$$

$$5.4 \times 0.31 = 1.7 \text{ MM} \quad 1.7$$

1.7 Land (est. 164 acres @ 5500/acre) 0.9

1.8 Total Capital Requirement (TCR)

$$\text{TCR} = \sum 1.1 \text{ thru } 1.7 \quad 1027.91 \text{ MM}$$

The Variable Operating Costs can be broken into 4 elements as follows:

2.0	<u>Variable Operating Costs</u>		47.67
		<u>1981 MM \$/yr</u>	
2.1	Fuel	46.35	
2.2	Water	0.46	
2.3	Ash Disposal	0.55	
2.4	Catalysts & Chemicals	0.31	
		<u>47.67</u>	

The Fixed Operating Costs can be broken into 3 elements as follows:

3.0	<u>Fixed Operating Costs</u>		24.41
3.1	Operating Labor	3.74	
3.2	Maintenance Costs (60% Matis, 40% Labor)	17.45	
3.3	Overhead Charges (30% Oper + Maint. Labor)	3.22	
		<u>24.41</u>	24.41

4.0 The Fuel Cell Replacement Expense is treated as a yearly levelized sinking fund (for the first 10 years only). The cost is \$25.66 MM/yr.

The Total Annual Operating Costs equals the sum of 2.0, 3.0, and 4.0, or \$97.74 MM/ yr in 1981 \$.

The COE based on the levelized annual revenue requirement for the first 10 years is made up of the following charges:

		<u>Levelized</u>	
		<u>Revenue Requirement</u>	
		MM \$/YR	
185.0	•	Capital Related Charges	
36.3	•	Fixed Operating Costs	
73.5	•	Variable Operating Costs	
25.7	•	Fuel Cell Replacement Expense	
<u>\$320.5</u>	=	Total Annual Revenue Requirement	

COE = \$0.0834 per kW-hr for 675 MW plant at 65% operating factor

1.0 TOTAL CAPITAL REQUIREMENT (TCR)

1.1 Total Plant Cost (See Appendix A for Unit Costing Bases):

= \$945.35 MM

1.2 Allowance For Funds During Construction:

Basis: 3 yr Construction Period; i = discount rate = .125; e_i = inflation rate = 0.085

mid'78 mid'79 mid'80 mid'81
|-----|-----|-----|

$$\begin{aligned} \frac{1}{3} \text{ TPC} &= \frac{1}{3} \text{ TPC} \\ \frac{1}{3} \text{ TPC} \xrightarrow{0} &= \frac{1}{3} \text{ TPC} \cdot \frac{1+i}{1+e_i}^1 = \frac{1}{3} \text{ TPC} \frac{1.125}{1.085}^1 \\ \frac{1}{3} \text{ TPC} \xrightarrow{0} &= \frac{1}{3} \text{ TPC} \cdot \frac{1+i}{1+e_i}^2 = \frac{1}{3} \text{ TPC} \frac{1.125}{1.085}^2 \\ &= \text{TPC} \frac{1}{3} (1 + 1.0751 + 1.0369) \\ &= 1.0373 \times \text{TPC} \end{aligned}$$

$$\text{AFDC} = .0373 \times \text{TPC}$$

$$= .0373 \times 945.35$$

$$= \$35.28 \text{ MM}$$

(See also EPRI TAG P-2410 SR p. B-46)

1.3 Prepaid Royalties:

$$= (0.005) \times (\text{T.P.C.} - \text{Gen'l Facilities} - \text{Project Cont. on Gen'l Facilities})$$

$$= (.005) \times (945.35 - 54.77 - 0.3 \times 54.77)$$

$$= \$4.37 \text{ MM}$$

1.4 Preproduction Or Start-up Costs:

a) 1 Month Fixed Oper. Costs

$$= 1/12 \times \$24.35 \text{ MM/Yr}$$

$$= \$2.03 \text{ MM}$$

b) 1 month of variable operating costs at full capacity

$$= \frac{\$47.67}{.65} \times \frac{1}{12} = \$6.1 \text{ MM}$$

c) 25% of full capacity fuel cost for 1 month:

$$= \frac{\$46.35}{.65} \times \frac{1}{12} \times .25$$

$$= \$1.5 \text{ MM}$$

d) 2% of the Total Plant Cost

$$= .02 \times \$945.35$$

$$= \$18.9 \text{ MM}$$

Total Pre-Production or Start-Up Costs

= a \$ 2.03

b 6.1

c 1.5

d 18.9

$$\$28.53 \text{ MM (1981 \$)}$$

1.5 Inventory Capital:

= 60 days inventory of fuel + catalysts + chemicals
@ full capacity

Coal: $\frac{\$46.35}{.65} \times \frac{60}{365} = \11.7

Cat. & Chemicals: $\frac{0.31}{.65} \times \frac{60}{365} = \$ 0.08$

$$\underline{\$11.78 \text{ MM (1981 \$)}}$$

1.6 Initial Chemical & Catalyst Charge:

Basis: Prorate based on AP-1543

$$\frac{\text{Initial Chem. + catalyst Charge} = 2,451}{\text{Annual Cat. \& Chem. charge} = 455} = 5.4$$

$$\text{Initial Charge} = 5.4 \times \$0.31 \text{ MM/Yr}$$

$$= \$1.7 \text{ MM/Yr (1981)}$$

Total Capital Requirement: (TCR Cont'd)

1.7 Land Cost (scaled from AP-1543)

AP-1543: 1200 MW (10,000 TPD Coal) requires 210 acres

- Coal Storage Area = 105 acres
- Process Area = 105 acres

Assume for 675 MW plant that process area is the same while the coal storage area is proportional to power output.

$$\text{Area} = 105 + 675/1200 \times 105 = 164 \text{ acres}$$

$$\text{Cost @ } \$5500/\text{acre} = \$902,000$$

1.8 Total Capital Requirement

$$= \sum 1.1 \text{ thru } 1.8$$

$$= \$ 1027.91 \text{ MM}$$

2.0 VARIABLE OPERATING COSTS

Fuel	46.35
Water	0.46
Ash Disposal	0.55
Catalysts & Chemicals	0.31

\$47.67 MM @ 65% capacity (1981 \$)

2.1 Fuel:

Yearly Coal Costs @ 65% operating factor:

$$403, \frac{199 \text{ lbs}}{\text{hr}} \times \frac{12,235 \text{ Btu}}{\text{lb}} \times 365 \times 24 \times .65 \times \frac{\$1.65}{10^6 \text{ Btu's}}$$
$$= \$46.35 \text{ MM (1981 \$)}$$

2.2 Water:

assume the following

1. 90% of water use is from cooling tower
2. 90% of cooling tower use is for steam cycle condenser
3. Steam Cycle efficiency = 38%
 $\frac{\text{Ratio of Condenser Duty}}{\text{Steam turbine power}} = \frac{1-.38}{.38}$

$$\text{Steam turbine input} = 192 \text{ MW}$$
$$= 655.3 \text{ MM Btu/hr}$$

$$\text{Condenser cooling} = \frac{655.3 \times 1-.38}{.38}$$
$$= 1,069 \text{ MM Btu/hr}$$

$$\text{Total cooling Twr duty} = \frac{1,069 \text{ MM Btu/hr}}{.9}$$
$$= 1,188 \text{ MM Btu/hr}$$

$$\text{CTW make-up} = 1,188 \times 10^6 \frac{\text{Btu}}{\text{hr}} \times \frac{1 \text{ lb H}_2\text{O M.U.}}{1000 \text{ Btu's}}$$
$$= 1,188 \times 10^3 \text{ lbs H}_2\text{O/hr}$$
$$= 142,600 \text{ gal/hr}$$

$$\text{Total Water use/hr} = \frac{142,600}{.9} = 158,440 \text{ gal/hr}$$

Variable Operating Costs (Cont'd)

Costs per year @ 65% capacity

$$158,440 \frac{\text{gal}}{\text{hr}} \times 365 \times 24 \times .65 \frac{\$.50}{1000 \text{ gal}}$$

$$= \$0.46 \text{ MM/Yr (1981 \$)}$$

$$\text{Note: Total Water Use} = \frac{3,900 \text{ gpm}}{1000 \text{ MW}}$$

2.3 Ash Disposal Costs:

Assume Costs = \$5/ton

Costs @ 65% Cap. =

$$\begin{array}{l} \text{TPD} \quad \text{Ash} \\ 4838 \frac{\text{coal}}{\text{day}} \times .096 \times 365 \times .65 \times \$5 \\ = \$0.55 \text{ MM/Yr (1981 \$)} \end{array}$$

2.4 Catalysts + Chemicals:

prorate from AP-1543

1976 costs = \$.455 MM/Yr 1976
for 10,000 TPD coal

1981 costs for 65% oper. factor:

$$= \$.455 \times 10^6 \times \frac{4838}{10,000} \times 1.54 \times \frac{.65}{.7}$$

$$= \underline{\underline{\$.031}} \text{ MM/yr 1981 \$}$$

C.2

3.0 FIXED OPERATING COSTS

Operating Labor	=	3.7
Maintenance costs	=	17.45
Overhead Charges	=	<u>3.2</u>
		\$24.35

3.1 Operating Labor:

1981 ALR = \$15.25/person-hour

- using AP-1543 as basis:

Total Plant operating labor = 28 people per shift
 = 245,280 man-hrs/year
 = \$3.74 MM/Yr for 1981 \$

3.2 Maintenance Costs: Materials (60%) + Labor (40%)

	% of Capital Cost	Unit Maintenance Cost MM\$	Unit Capital Cost MM\$
O ₂	2		118.56
Coal Handling	3		25.74
Gasif + Ash Handling	4.5		36.07
Gasifier Gas Cooling + Trtng	3.7		68.95
AGR + Zn Oxide	2.0/0.5		52.26/6.37
SRU + TGU	<u>2.0</u>		<u>16.11</u>
Sub-total Fuel Processor		8.75	324.06
Fuel Cell Modules, Piping & Sub-system	1.0		268.75
Exp. Turbines + Stm. Cycle	2.25		183.56
Inverters	1.0		61.08
Land, Struct & Improvements	1.5		71.20
Accessory Elect. Equip., Step-up & Trans.	<u>1.5</u>		<u>36.66</u>
balance sub-total		8.7	<u>621.29</u>
		17.45	945.35

Total maintenance Costs = Maint. Materials + Maint. Labor

Total Maint. Costs = Total (.60) + Total x (.40)

\$17.45 MM/Yr = \$10.45 MM/Yr + \$7 MM/Yr

3.3 Overhead Charges: (30% of Oper. + Main. Labor)

= 0.3 (3.74 + 7)
 = \$ 3.2 MM/yr

$$\begin{aligned} \text{Cost to replace @ end of 10 yrs} &= 58 (1.085)^{10} \\ &= \$131.14 \end{aligned}$$

$$\begin{aligned} \times [(1+i)^5 + (1+i)^4 + (1+i)^3 + (1+i)^2 + (1+i)] &= 7.21829X \\ &= 5 \text{ yr. Sinking Fund} \end{aligned}$$

$$\begin{aligned} \text{Amount remaining after 1st replacement:} & \\ & 7.21829X - 87.21 \end{aligned}$$

$$\begin{aligned} \text{Amount @ end of 10 yr} &= \$131.14 \text{ mm} = \\ & \text{Amount Remaining after 1st Payment + Accumulated 5 yr. Sinking Fund} \end{aligned}$$

$$= (7.21829X - 87.21)(1.125)^5 + 7.21829X = 131.14$$

$$\begin{aligned} 20.23X - 157.16 &= 131.14 \\ X &= \$14.25 \text{ mm/yr} \end{aligned}$$

$$\begin{aligned} + \text{ Process Cont. @ 50\%} &= 7.13 \text{ mm/yr} \\ + \text{ Project Cont. @ 30\%} &= \underline{4.28 \text{ mm/yr}} \\ &= \underline{\underline{\$25.66 \text{ mm/yr}}} \end{aligned}$$

$$\frac{25.66 \times 10^6}{442,000 \text{ KW} \times 0.65 \times 365 \times 24} = \$0.01068/\text{KW-hr} \text{ of F.C. power}$$

5.0 COST OF ELECTRICITY (COE) CALCULATION

5.1 Calculate short-term 10 yr, levelized COE:

1st 10 Years Levelized Revenue Requirement

$$= (\text{TCR}) P_f + (\text{Expenses}) \times \text{LF}_f \quad - \text{Notes (1) \& (2)}$$

$$= 1027.91 \times .180 + (\text{Expenses}) \times \text{LF}_f$$

$$\underline{\text{Expenses}} = \text{Fixed} + \text{Coal} + \text{Other Variable} + \text{Fuel Cell Replacement}$$

	<u>1981</u> <u>Exps</u>	Int. Rate (i)	Real Escal. (e _r)	Inflat. Rate (e _i)	Apparent Esc. (e _a)	Level Factor (L ₁₀)	<u>Level</u> <u>Exps</u>
o Fixed	24.41	.125	0	.085	.085	1.4878	36.32
o Var.-Coal	46.4	"	0.007	"	.0926	1.5434	71.6
-Others	1.3	"	0	"	.085	1.4878	1.9
o Fuel Cell	25.66	"	"	"	"	-	<u>25.66</u>
							135.48

Levelized Revenue Requirement for first 10 years =

(mm \$/yr)	(%)	
185.0	57.8	• Capital Related Charges
36.3	11.3	• Fixed Operating Costs
		• Variable Operating Costs
71.6	22.3	- Coal
1.9	0.6	- Others
<u>25.7</u>	<u>8.0</u>	• Fuel Cell Replacement Expense
\$320.5	100.0	

$$\begin{aligned} \text{C.O.E.} &= \$320.5 \times 10^6 \div (675 \times 10^3 \times .65 \times 24 \times 365) \\ &= \$0.0834 \text{ per kWhr} \end{aligned}$$

Notes:

- (1) P_f = factor for cumulative present value of carrying charges for the first f years of project
- (2) LF_f = levelizing factor for f years
- (3) Factors found in EPRI TAG p. A-3 and A-5

APPENDIX A

**Total Plant Cost Via
Unit Costing For Base Case
Reference Design**

BASE CASE MOLAR FLOWS

	1	2	3	4	5	6	7	8	9	10	Sulfur Plant Feed	Anode Out	Cathode Out	Cathode Recycle
Molar Flow, mph														
H ₂	12,569			12,569	-	23,469	3,112	-	-	-	-	3,214	-	-
CO	18,501			18,501	-	7,601	1,559	-	-	-	-	1,560	-	-
CH ₄	35			35	-	35	35	-	-	-	-	35	-	-
CO ₂	3,796			3,347	-	14,247	46,585	21,883	-	113,864	449	46,584	87,568	65,585
H ₂ O	7,792			-	30,507	19,607	39,904	43,146	-	172,655	-	39,862	172,655	129,509
N ₂	340			340	-	340	340	117,746	117,406	471,179	-	340	471,179	353,433
H ₂ S	440			-	-	-	-	-	-	-	440	-	-	-
EOS	26			<2ppm	-	-	-	-	-	-	26	-	-	-
O ₂	-			-	-	-	15,604	31,209	75,590	-	-	-	62,442	46,838
NH ₃	83			-	-	-	-	-	-	-	83	-	-	-
Total mph	43,582	34,430	31,815	34,792	30,507	65,299	91,550	198,379	148,615	833,288		91,419	793,844	595,465

Notes: Molar flows are approximate and used for cost scaling purposes only.

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A-1-

BASES FOR UNIT ESTIMATING

UNIT	Ref.	YR	Total Plant Inv.	Process Cont.	Proj. Cont.	Const-ru ^{cted} Cost w/out Cont.	Plant Basis	Size	STRAINS	TOTAL	SPARE
			MM\$			Note(1)					
Oxygen Plant & Delivery	AP-642	Mid-76	117.4	-	15.3	102.1	O ₂ TPD	8,379	5	0	
Coal Handling	AP-1543	"	26.0	-0-	3.4	22.6	Coal TPD	10,000	2	0	
Gasification & Ash Handling	"	"	32.3	3.5	3.8	25.0	"	"	6	1	
Gasifier Gas Cooling	"	"	68.3	3.9	8.4	56.0	Gasifier Effluent mph	90,078	6	1	
Acid Gas Removal	"	"	46.3	0	6.0	40.3	Gasifier Dry Effluent (mph)	73,972	2	0	
Zn Oxide Treatment	"	"	5.6	0	0.7	4.9	Fuel Gas Flow mph	67,621	2	0	
Sulfur Recovery & Tail Gas	"	"	20.8	1.3	2.6	16.9	Acid Gas Flow (mph)	6,428	3	1 (14%)	
								H ₂ S (3-50% Units)			
								8,227			
Fuel Cells	AP-1670;	Mid-78					Fuel Cell AC Output (MM)		5	5 Islands,	
	P. B-4	"	158.7	49.6	9.9	99.2	(Fuel Cell Module @\$97.63/kM w/out contingency)		18	Fuel cells/ Islands,	
Remainder Fuel Cell Sub-Systems	"	"							5	MM/Fuel Cell	
Steam Bottoming Cycle	AP-1670	"	82.4	-	7.5	74.9	Gross Shaft Output (MM)	125.7	5		
	AP-1670	"					Gross ST Output (MM)	171.6	1		
Inverters	AP-1670	"	43.0	2.9	2.9	29.0	Fuel Cell AC Output (MM)	442	45	(1 Island)	
Accessory Elect. Equip. Step Up & Transmission	"	"	20.1	-	1.8	18.3	Total Plant Output (MM)	675			
General Facilities	"	"	49.5	-	4.5	45.0	10% of sub-total less Engineering and Home Office overhead plus fees	675			

Note (1): Constructed Costs include Engineering and Home Office Overhead plus Fees.

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UNIT COST ESTIMATES FOR BASE CASE REFERENCE DESIGN

UNIT	Plant Size	# Trains Total/Spare	Costing Scale Expn	Mid. 76 Cost w/out Cont.	Total Constructed Cost Note (1)(2) Mid. 81 Cost	Process (On-Site) Costs	E.H.O. Ovhld. + Fees Note (3)	Note (3) Process Cont. (1981)	Proj. Cont.	Total Plant Cost	\$/KW
				\$MM	\$MM	\$MM	\$MM	\$MM	\$MM	\$MM	
Oxygen Plant & Delivery	4,054 TPD	3/0	0.75	59.2	91.2			-0-	27.36	118.56	176.0
Coal Handling	4,838 TPD	1/0	0.78	12.8	19.8			-0-	5.94	25.74	38.0
Gasification & Ash Handling	4,838 TPD	3/1	0.60	16.2	24.9			3.7	7.47	36.07	53.0
Gasifier Gas Cooling	43,582 mph	3/1	0.82	30.9	47.5			7.2	14.25	68.95	102.0
Acid Gas Removal	35,790 mph	2/0	0.60	26.1	40.2			-	12.06	52.26	77.0
Zn Oxide Treatment	34,866 mph	2/0	0.65	3.2	4.9			-	1.47	6.37	9.4
Sulfur Recovery & Tail Gas [Acid Gas=44% H ₂ S]	2,167 mph	2/1									
Fuel Processor Sub-Totals			0.60	7.6	11.7			0.9	3.51	16.11	24.0
Fuel Cells	432 MW		1.0	97.0	149.3			11.8	72.06	324.06	480.0
Fuel Cell Piping, etc.								74.7	44.79	268.79	398.0
Steam Bottoming Cycle	172 MW		1.0	91.7	141.2			0	42.36	183.56	272.0
Expansion Turbines + Steam Cycle	192 MW		1.0	28.3	43.6			4.4	13.08	61.08	90.5
Inverters	432 MW		1.0	18.3	28.2			0	8.46	36.66	54.0
Accessory Elect. Equip. Step Up & Transmission	675		1.0								
Power Cycles Sub-totals				362.3	602.5			79.1	108.69	550.09	814.5
Process (On-Site) Sub-Total				602.5	602.5	547.7	54.77	90.9	180.75	874.15	1,295.0
General Facilities (=10% of Process (On-Site) Costs)				54.77	54.77				16.43	71.2	105.0
TOTAL				657.27	657.27			90.9	197.18	945.35	1,400.0

NOTES:

1. Constructed Costs include Engineering and Home Office Overhead plus Fees.
2. Time Esc. Factor from Mid'81 = 302/196.1 = 1.54 (GE Plant Cost Indices); Mid'78 to Mid'81 = 1.36
3. E.H.O. Ovhld + Fees = Engineering and Home Office Overhead plus Fees
4. Process Contingencies used as follows: Gasif. = 15%; Ash Handling = 5%; High Temp. Gas Cooling = 15%; Tail Gas = 15%; Fuel Cell = 50%; Inverters = 10%

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APPENDIX B. Summary Printouts of Cases for
Economic Analysis

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105

Case 1

The Base Case at 675 MW Output

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TABULATION OF SHAFT ENERGY AVAILABLE OR NEEDED.

102	-57167.7	KILOWATTS	AIR COMPRESSOR FOR SUPPLY TO CRYOGENIC OXYGEN PLANT
104	-24942.3	KILOWATTS	CRYOGENIC OXYGEN COMPRESSOR FOR SUPPLY TO COAL GASIFIER
2	-5907.3	KILOWATTS	2 TEXACO GASIFIER FOR JET PROP. LA3S. BASE CASE.
13	-1137.2	KILOWATTS	13 REMOVE SULFUR FROM RAW GAS BY SELEXOL SYSTEM.
17	23530.6	KILOWATTS	17 EXPAND SWEETENED RAW GAS TO FUEL CELL PRESSURE.
22	22766.0	KILOWATTS	22 FLOW STEAM DOWN POWER TURBINE. ADD TO PROCESS GAS (# 17).
3	-152026.1	KILOWATTS	3 COMPRESS COMBUSTION AIR TO FUEL CELL PRESSURE.
10	564644.5	KILOWATTS	10 THE FUEL CELL, 55% CONVERSION, SHIFT ONLY WITHIN ANODE.
15	-9204.4	KILOWATTS	15 ANODE EXIT GAS BOOSTER FAN
20	-39842.5	KILOWATTS	20 CATHODE RECYCLE GAS CIRCULATION FAN.
21	252363.0	KILOWATTS	21 EXPAND STACK GAS THROUGH TURBINE TO RECOVER POWER.

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1 JPL-1 CASE 1 JPL-1
STEAM LIST
19 MAR 93 10:20:52

BASE CASE 85% CONV., NO DRENCH, NO PRE-SHIFT, ISO EFF: COMP 20% T 92X

TABULATION OF STEAM NEEDED IN PROCESS.

13	65.0 PSIA	4687. WOLS STEAM	13 REMOVE SULFUR FROM PAM GAS BY SELEXOL SYSTEM.
22	1500.0 PSIA	16235. WOLS STEAM	22 BLOW STEAM DOWN POWER TURBINE. ADD TO PROCESS GAS (# 17).

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CASE 1 JPL-1
WASTE HEAT OUT
18 MAR 31 10:20:52

1 JPL-1 BASE CASE 55% CONV., NO QUENCH, NO PRE-SHIFT, ISO EFF: COMP 98% T 92%

TABULATION OF WASTE HEAT AVAILABLE FROM PROCESS BOILERS.

4	200.0 DEGREES F.	658.49 MILLIONS BTU	4 COOL RAW GAS TO 800 DEGREES FAHRENHEIT.
7	600.0 DEGREES F.	13.97 MILLIONS BTU	7 WASTE HEAT BOILER PARALLEL TO RAW GAS RECUP. EXCH. TRAIN.
14	550.0 DEGREES F.	121.96 MILLIONS BTU	14 COOL ANODE EXIT GAS IN WHR - RAISE STEAM.
18	1050.0 DEGREES F.	562.39 MILLIONS BTU	18 COOL CATHODE RECYCLE GAS IN WHR - RAISE STEAM.

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CASE 2

Add Quench, Shift, Crossflow Heat Exchange

CASE L-JPL-1F
POWER LIST
15 FEB 83 14:30:43

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L-JPL-1F MODIFIED BASE CASE 85% CONV, QUENCH, PRE-SHIFT, COSX, TURBO-EXP MOVED.

TABULATION OF SHAFT ENERGY AVAILABLE OR NEEDED.

102	-57995.0	KILOWATTS	AIP COMPRESSOR FOR SUPPLY TO CRYOGENIC OXYGEN PLANT
104	-25303.2	KILOWATTS	CRYOGENIC OXYGEN COMPRESSOR FOR SUPPLY TO COAL GASIFIER
2	-5992.9	KILOWATTS	2 TEXACO GASIFIER FOR JET PROP. LAGS. BASE CASE.
19	45529.9	KILOWATTS	19 EXPAND SHIFT (I) EXIT GAS THROUGH POWER RECOVERY TURBINE.
30	-7289.9	KILOWATTS	30 REMOVE SULFUR FROM RAW GAS BY SELEXOL SYSTEM.
42	6753.6	KILOWATTS	42 ADD STEAM VIA TURBINE; PREVENT CARBON DROP IN FUEL CELL.
44	-153934.7	KILOWATTS	44 COMPRESS COMBUSTION AIR TO FUEL CELL PRESSURE.
51	531552.5	KILOWATTS	51 THE FUEL CELL, 85% CONVERSION, SHIFT ONLY WITHIN ANODE.
57	-8783.5	KILOWATTS	57 ANODE EXIT GAS BOOSTER FAN, TO FEED CATALYTIC BURNER.
63	-15868.1	KILOWATTS	63 CATHODE RECYCLE GAS CIRCULATION FAN.
64	262707.6	KILOWATTS	64 EXPAND STACK GAS THROUGH TURBINE TO RECOVER POWER.



L-JPL-1F MODIFIED BASE CASE 85% CONV, QUENCH, PRE-SHIFT, COSX, TURBO-EXP MOVED.
CASE L-JPL-1F
STEAM LIST
15 FEB 83 14:30:43

TABULATION OF STEAM NEEDED IN PROCESS.

30	65.0 PSIA	12705. MOLS STEAM	30 REMOVE SULFUR FROM RAW GAS BY SELEXOL SYSTEM.
42	1500.0 PSIA	5074. MOLS STEAM	42 ADD STEAM VIA TURBINE; PREVENT CARBON DROP IN FUEL CELL.

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L-JPL.1F
CASE L-JPL.1F
WASTE HEAT OUT
15 FEB 83 14:30:43

L-JPL.1F MODIFIED BASE CASE 85% CONV, QUENCH, PRE-SHIFT, COSX, TURBO-EXP MOVED.

TABULATION OF WASTE HEAT AVAILABLE FROM PROCESS BOILERS.

	700.0 DEGREES F.	276.66 MILLIONS BTU	8 COOL RAW GAS TO 700 DEGREES FAHRENHEIT, MAKE STEAM.
20	600.0 DEGREES F.	51.44 MILLIONS BTU	20 COOL INTER-STAGE SHIFT GAS IN WHB; RAISE STEAM
54	920.0 DEGREES F.	333.70 MILLIONS BTU	54 COOL ANODE EXIT GAS IN WHB - RAISE STEAM.
60	1050.0 DEGREES F.	279.93 MILLIONS BTU	60 COOL CATHODE RECYCLE GAS IN WHB - RAISE STEAM.

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CASE 3

Change to Fluidized-Bed Gasifier,
Producing Low Methane

TABULATION OF SHAFT ENERGY AVAILABLE OR NEEDED.

102	-48638.1 KILOWATTS	AIR COMPRESSOR FOR SUPPLY TO CRYOGENIC OXYGEN PLANT
104	-20051.4 KILOWATTS	CRYOGENIC OXYGEN COMPRESSOR FOR SUPPLY TO COAL GASIFIER
2	-5929.1 KILOWATTS	2 U-GAS GASIFIER NORMAL OPERATION. STREAM #2 = NET MAKE.
20	37054.1 KILOWATTS	20 EXPAND SHIFT (I) EXIT GAS THROUGH POWER RECOVERY TURBINE.
29	-3117.9 KILOWATTS	29 REMOVE SULFUR FROM RAW GAS BY SELEXOL SYSTEM.
42	10254.6 KILOWATTS	42 ADD STEAM VIA TURBINE; PREVENT CARBON DROP IN FUEL CELL.
44	-194271.1 KILOWATTS	44 COMPRESS COMBUSTION AIR TO FUEL CELL PRESSURE.
51	569473.7 KILOWATTS	51 THE FUEL CELL, 85% CONVERSION, SHIFT ONLY WITHIN ANODE.
57	-7947.2 KILOWATTS	57 ANODE EXIT GAS BOOSTER FAN, TO FEED CATALYTIC BURNER.
63	-59150.6 KILOWATTS	63 CATHODE RECYCLE GAS CIRCULATION FAN.
64	326350.8 KILOWATTS	64 EXPAND STACK GAS THROUGH TURBINE TO RECOVER POWER.
70	-2186.6 KILOWATTS	70 COMPRESS RECYCLE GAS TO INJECT FEED COAL TO GASIFIER.

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L.JPL-2 FLUID BED BASE CASE 95% CONV, QUENCH, PRE-SHIFT, COSX, 8.33% CH4 RAW

CASE L.JPL-2
STEAM LIST
15 FEB 83 16:41:31

TABULATION OF STEAM NEEDED IN PROCESS.

2	550.0 PSIA	27232.	MOLS STEAM	2	U-GAS GASIFIER NORMAL OPERATION. STREAM #2 = NET MAKE.
29	65.0 PSIA	14312.	MOLS STEAM	29	REMOVE SULFUR FROM RAW GAS BY SELEXOL SYSTEM.
42	1500.0 PSIA	7704.	MOLS STEAM	42	ADD STEAM VIA TURBINE; PREVENT CARBON DROP IN FUEL CELL.

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CASE L-JPL.2
WASTE HEAT OUT
15 FEB 83 14:41:31

L-JPL.2 FLUID BED BASE CASE 85% CONV, QUENCH, PRE-SHIFT, COSX, 8.33% CH4 RAW

TABULATION OF WASTE HEAT AVAILABLE FROM PROCESS BOILERS.

9	700.0 DEGREES F.	563.60 MILLIONS BTU	9 COOL RAW GAS TO 700 DEGREES FAHRENHEIT, MAKE STEAM.
21	600.0 DEGREES F.	10.02 MILLIONS BTU	21 COOL INTER-STAGE SHIFT GAS IN WHB; RAISE STEAM
54	920.0 DEGREES F.	375.54 MILLIONS BTU	54 COOL ANODE EXIT GAS IN WHB - RAISE STEAM.
60	1040.0 DEGREES F.	880.81 MILLIONS BTU	60 COOL CATHODE RECYCLE GAS IN WHB - RAISE STEAM.

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CASE 4

Add Internal Reforming to Fuel Cell

TABULATION OF SHAFT ENERGY AVAILABLE OR NEEDED.

102	-41552.5	KILOWATTS	AIR COMPRESSOR FOR SUPPLY TO CRYOGENIC OXYGEN PLANT
104	-17130.3	KILOWATTS	CRYOGENIC OXYGEN COMPRESSOR FOR SUPPLY TO COAL GASIFIER
2	-5064.5	KILOWATTS	2 U-GAS GASIFIER NORMAL OPERATION. STREAM #2 = NET MAKE.
20	31691.0	KILOWATTS	20 EXPAND SHIFT (I) EXIT GAS THROUGH POWER RECOVERY TURBINE.
29	-6935.3	KILOWATTS	29 REMOVE SULFUR FROM RAW GAS BY SELEXOL SYSTEM.
42	8760.7	KILOWATTS	42 ADD STEAM VIA TURBINE; PREVENT CARBON DROP IN FUEL CELL.
44	-165206.1	KILOWATTS	44 COMPRESS COMBUSTION AIR TO FUEL CELL PRESSURE.
51	629915.7	KILOWATTS	51 THE FUEL CELL, MAXIMUM CONVERSION, REFORM WITHIN ANODE.
57	-7894.3	KILOWATTS	57 ANODE EXIT GAS BOOSTER FAN, TO FEED CATALYTIC BURNER.
63	-18950.0	KILOWATTS	63 CATHODE RECYCLE GAS CIRCULATION FAN.
64	270629.7	KILOWATTS	64 EXPAND STACK GAS THROUGH TURBINE TO RECOVER POWER.
70	-18556.7	KILOWATTS	70 COMPRESS RECYCLE GAS TO INJECT FEED COAL TO GASIFIER.

L-JPL.3 CASE L-JPL.3
STEAM LIST
15 FEB 83 15:00:37

L-JPL.3 FLUID BED, REFORM, 85% CONV, QUENCH, PRE-SHIFT, COSK, 8.33% CH4 9AM

TABULATION OF STEAM NEEDED IN PROCESS.

2	550.0 PSIA	23265. MOLS STEAM	2 U-GAS GASIFIER NORMAL OPERATION. STREAM #2 = NET MAKE.
29	65.0 PSIA	12227. MOLS STEAM	29 REMOVE SULFUR FROM RAW GAS BY SELEXOL SYSTEM.
42	1500.0 PSIA	6592. MOLS STEAM	42 ADD STEAM VIA TURBINE; PREVENT CARBON DROP IN FUEL CELL.

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CASE L.JPL.3
WASTE HEAT OUT
15 FEB 83 15:00:37

L.JPL.3 FLUID BED, REFORM, 95% CONV, QUENCH, PRE-SHIFT, COSX, 8.33% CH4 RAW

TABULATION OF WASTE HEAT AVAILABLE FROM PROCESS BOILERS.

9	700.0 DEGREES F.	481.51 MILLIONS BTU	9 COOL RAW GAS TO 700 DEGREES FAHRENHEIT, MAKE STEAM.
21	600.0 DEGREES F.	3.44 MILLIONS BTU	21 COOL INTER-STAGE SHIFT GAS IN WHB; RAISE STEAM.
54	925.0 DEGREES F.	311.66 MILLIONS BTU	54 COOL ANODE EXIT GAS IN WHB - RAISE STEAM.
60	1090.0 DEGREES F.	204.95 MILLIONS BTU	60 COOL CATHODE RECYCLE GAS IN WHB - RAISE STEAM.

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CASE 5

Change Gasifier to High Methane Production

CASE L.AIR.4E
POWER LIST
15 FEB 83 15:36:37

L.AIR.4E FLUID BED, REFORM, 85% CONV, QUENCH, PRE-SHIFT, COSX, 13.5% CH4 RAW

TABULATION OF SHAFT ENERGY AVAILABLE OR NEEDED.

102	-37732.1	KILOWATTS	AIR COMPRESSOR FOR SUPPLY TO CRYOGENIC OXYGEN PLANT
104	-15555.3	KILOWATTS	CRYOGENIC OXYGEN COMPRESSOR FOR SUPPLY TO COAL GASIFIER
2	-4962.5	KILOWATTS	2 U-GAS GASIFIER NORMAL OPERATION, STREAM #2 = NET MAKE.
20	30072.5	KILOWATTS	20 EXPAND SHIFT (I) EXIT GAS THROUGH POWER RECOVERY TURBINE.
29	-6281.9	KILOWATTS	29 REMOVE SULFUR FROM RAW GAS BY SELEXOL SYSTEM.
42	10102.0	KILOWATTS	42 ADD STEAM VIA TURBINE; PREVENT CARBON DROP IN FUEL CELL.
44	-162012.6	KILOWATTS	44 COMPRESS COMBUSTION AIR TO FUEL CELL PRESSURE.
51	632918.0	KILOWATTS	51 THE FUEL CELL, MAXIMUM CONVERSION, REFORM WITHIN ANODE.
57	-7883.5	KILOWATTS	57 ANODE EXIT GAS BOOSTER FAN, TO FEED CATALYTIC BURNER.
63	-19220.3	KILOWATTS	63 CATHODE RECYCLE GAS CIRCULATION FAN.
64	268004.1	KILOWATTS	64 EXPAND STACK GAS THROUGH TURBINE TO RECOVER POWER.
70	-1702.0	KILOWATTS	70 COMPRESS RECYCLE GAS TO INJECT FEED COAL TO GASIFIER.

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L-AIR.4E FLUID BED, REFORM, 85% CONV, QUENCH, PRE-SHIFT, COSX, 13.5% CH4 RAW

CASE L-AIR.4E
STEAM LIST
15 FEB 83 15:36:37

TABULATION OF STEAM NEEDED IN PROCESS.

2	550.0 PSIA	22796. MOLS STEAM	2 U-GAS GASIFIER NORMAL OPERATION. STREAM #2 = NET MAKE.
29	65.0 PSIA	11258. MOLS STEAM	29 REMOVE SULFUR FROM RAW GAS BY SELEXOL SYSTEM.
42	1500.0 PSIA	7500. MOLS STEAM	42 ADD STEAM VIA TURBINE; PREVENT CARBON DROP IN FUEL CELL.

L-AIR.4E FLUID BED, REFORM, 85% CONV, QUENCH, PRE-SHIFT, COSX, 13.5X CH4 RAW

CASE L-AIR.4E
WASTE HEAT OUT
15 FEB 83 15:36:37

TABULATION OF WASTE HEAT AVAILABLE FROM PROCESS BOILERS.

9	700.0 DEGREES F.	434.34 MILLIONS BTU	9 COOL RAW GAS TO 700 DEGREES FAHRENHEIT, MAKE STEAM.
21	600.0 DEGREES F.	9.53 MILLIONS BTU	21 COOL INTER-STAGE SHIFT GAS IN WHB; RAISE STEAM.
54	925.0 DEGREES F.	315.70 MILLIONS BTU	54 COOL ANODE EXIT GAS IN WHB - RAISE STEAM.
60	1115.0 DEGREES F.	175.57 MILLIONS BTU	60 COOL CATHODE RECYCLE GAS IN WHB - RAISE STEAM.

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CASE 6

Employ High-Temperature Dust Removal

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L.OEX.2 HIGH TEMPERATURE DE-DUSTING, RAW GAS LIKE JPL6, QUENCH AFTER TURBINE.

TABULATION OF SHAFT ENERGY AVAILABLE OR NEEDED.

102	-36546.9	KILOWATTS	AIR COMPRESSOR FOR SUPPLY TO CRYOGENIC OXYGEN PLANT
194	-15066.7	KILOWATTS	CRYOGENIC OXYGEN COMPRESSOR FOR SUPPLY TO COAL GASIFIER
2	-4200.4	KILOWATTS	U-GAS GASIFIER NORMAL OPERATION. STREAM #2 = NET MAKE.
5	50794.8	KILOWATTS	EXPAND HOT RAW GAS VIA TURBINE; COLLECT KILOWATTS.
21	-6097.0	KILOWATTS	REMOVE SULFUR FROM RAW GAS BY SELEXOL SYSTEM.
34	9746.6	KILOWATTS	ADD STEAM VIA TURBINE; PREVENT CARBON DROP IN FUEL CELL.
36	-156923.6	KILOWATTS	COMPRESS COMBUSTION AIR TO FUEL CELL PRESSURE.
43	613087.3	KILOWATTS	THE FUEL CELL, MAXIMUM CONVERSION, PERFORM WITHIN ANODE.
48	-7636.0	KILOWATTS	ANODE EXIT GAS BOOSTER FAN, TO FEED CATALYTIC BURNER.
54	-13637.0	KILOWATTS	CATHODE RECYCLE GAS CIRCULATION FAN.
55	257893.1	KILOWATTS	EXPAND STACK GAS THROUGH TURBINE TO RECOVER POWER.
63	-1249.8	KILOWATTS	COMPRESS RECYCLE GAS TO INJECT FEED COAL TO GASIFIER.

CASE L.QEX.2
STEAM LIST
17 FEB 83 7:33:26

L.QEX.2 HIGH TEMPERATURE DE-DUSTING, RAW GAS LIKE JPL4, QUENCH AFTER TURBINE.

TABULATION OF STEAM NEEDED IN PROCESS.

2	550.0 PSIA	22090. MOLS STEAM	2 U-GAS GASIFIER NORMAL OPERATION. STREAM #2 = NET MAKE.
21	65.0 PSIA	10912. MOLS STEAM	21 REMOVE SULFUR FROM RAW GAS BY SELEXOL SYSTEM.
34	1500.0 PSIA	7323. MOLS STEAM	34 ADD STEAM VIA TURBINE; PREVENT CARBON DROP IN FUEL CELL.

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L.OEX.2 HIGH TEMPERATURE DE-DUSTING, RAW GAS LIKE JPL4, QUENCH AFTER TURBINE.

CASE L.OEX.2
WASTE HEAT OUT
17 FEB 83 7:33:26

TABULATION OF WASTE HEAT AVAILABLE FROM PROCESS BOILERS.

11	500.0 DEGREES F.	310.85 MILLIONS BTU	11 COOL SHIFT FEED GAS TO 60CF, RAISE STEAM.
13	600.0 DEGREES F.	109.45 MILLIONS BTU	13 COOL INTER-STAGE SHIFT GAS IN WHB; RAISE STEAM.
49	925.0 DEGREES F.	307.25 MILLIONS BTU	46 COOL ANODE EXIT GAS IN WHB - RAISE STEAM.
51	1115.0 DEGREES F.	171.90 MILLIONS BTU	51 COOL CATHODE RECYCLE GAS IN WHB - RAISE STEAM.

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CASE 7

Employ Sulfur-Tolerant Fuel Cell

CASE L. THICEL
POWER LIST
23 FEB 83 10:11:20

L. THICEL SULFUR-TOLERANT CELL, 95% CONV, INT. REFORM, NO PRECOOL, 13.5X CH4 RAW

TABULATION OF SHAFT ENERGY AVAILABLE OR NEEDED.

102	-34601.1	KILOWATTS	AIR COMPRESSOR FOR SUPPLY TO CRYOGENIC OXYGEN PLANT
104	-14264.6	KILOWATTS	CRYOGENIC OXYGEN COMPRESSOR FOR SUPPLY TO COAL GASIFIER
2	-4550.7	KILOWATTS	2 U-GAS GASIFIER NORMAL OPERATION. STREAM #2 = NET MAKE.
5	60261.5	KILOWATTS	5 EXPAND RAW GAS TO FUEL CELL PRESSURE; COLLECT KILOWATTS.
18	-1563.7	KILOWATTS	18 COMPRESS COAL INJECTION SIDESTREAM; RECYCLE TO GASIFIER.
22	-148563.8	KILOWATTS	22 COMPRESS COMBUSTION AIR TO FUEL CELL PRESSURE.
29	566517.0	KILOWATTS	29 THE FUEL CELL, MAXIMUM CONVERSION, REFORM WITHIN ANODE.
33	-5568.2	KILOWATTS	33 ANODE EXIT GAS BOOSTER FAN, TO FEED CATALYTIC BURNER.
36	247015.7	KILOWATTS	36 EXPAND STACK GAS FROM GASIFIER PRESSURE; COLLECT KW'S.
41	-50579.8	KILOWATTS	41 CATHODE RECYCLE GAS CIRCULATION FAN.

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L. THIGEL SULFUR-TOLERANT CELL, 85% CONV, INT. REFORM, NO PRECOOL, 13.5% CH4 RAW

CASE L. THIGEL
STEAM LIST
23 FEB 83 10:11:20

TABULATION OF STEAM NEEDED IN PROCESS.

2 550.0 PSIA 20904. MOLS STEAM 2 U-GAS GASIFIER NORMAL OPERATION. STREAM #2 = NET MAKE.

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L. THICEL SULFUR-TOLERANT CELL, 35% CONV, INT. REFORM, NO PRECOOL, 13.5% CH4 RAW

CASE L. THICEL
WASTE HEAT OUT
23 FEB 83 10:11:20

TABULATION OF WASTE HEAT AVAILABLE FROM PROCESS BOILERS.

8	1200.0 DEGREES F.	2.64 MILLIONS BTU	8 COOL RAW GAS IN WHB; MAKE STEAM.
32	900.0 DEGREES F.	321.60 MILLIONS BTU	32 COOL ANODL EXIT GAS IN WHB - RAISE STEAM.
38	1070.0 DEGREES F.	753.11 MILLIONS BTU	38 COOL CATHODE EXIT RECYCLE GAS IN WHB; RAISE STEAM.

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CASE 8

Employ High-Temperature CO₂ Transfer

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TABULATION OF SHAFT ENERGY AVAILABLE OR NEEDED.

102	-36441.0 KILOWATTS	AIR COMPRESSOR FOR SUPPLY TO CRYOGENIC OXYGEN PLANT
104	-15023.1 KILOWATTS	CRYOGENIC OXYGEN COMPRESSOR FOR SUPPLY TO COAL GASIFIER
2	-4792.7 KILOWATTS	2 U-GAS GASIFIER NORMAL OPERATION. STREAM #2 = NET MAKE.
27	-159451.0 KILOWATTS	27 COMPRESS COMBUSTION AIR TO FUEL CELL PRESSURE.
40	26633.0 KILOWATTS	40 EXPAND SHIFT (I) EXIT GAS THROUGH POWER RECOVERY TURBINE.
49	-2827.0 KILOWATTS	49 REMOVE SULFUR FROM RAW GAS BY SELEXOL SYSTEM.
63	709.7 KILOWATTS	63 ADD STEAM VIA TURBINE; PREVENT CARBON DROP IN FUEL CELL.
69	647939.1 KILOWATTS	NORMAL FUEL CELL, 1200 F INPUT SETS VOLTAGE, 1260 F EXIT.
76	-17155.8 KILOWATTS	76 CATHODE RECYCLE GAS CIRCULATION FAN.
81	-7451.0 KILOWATTS	31 ANODE EXIT GAS BOOSTER FAN, TO FEED CATALYTIC BURNER.
82	224662.9 KILOWATTS	82 EXPAND CATHODE EXIT STACK GAS TO RECOVER POWER.
96	-1371.3 KILOWATTS	96 COMPRESS COAL FEED INJECTION GAS, A SMALL RECYCLE.

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L.RUN.D2 HIGH TEMP. CO2 PASS ADDED TO RASE CASE, COMPARE L.AIR.4E
CASE L.RUN.D2
STEAM LIST
1 MAR 93 14:03:06

TABULATION OF STEAM NEEDED IN PROCESS.

2	550.0 PSIA	22016.	MOLS STEAM	2	U-GAS GASIFIR NORMAL OPERATION. STREAM #2 = NET MAKE.
36	1200.0 PSIA	4797.	MOLS STEAM	36	HIGH-TEMPERATURE CO2 REMOVAL FROM SHIFT I EXIT GAS.
37	1200.0 PSIA	1439.	MOLS STEAM	37	REGENERATION OF MGO PUTS CO2 IN CATHODE FEED AIR.
49	65.0 PSIA	9119.	MOLS STEAM	49	REMOVE SULFUR FROM RAW GAS BY SELEXOL SYSTEM.
63	1500.0 PSIA	536.	MOLS STEAM	63	ADD STEAM VIA TURBINE; PREVENT CARBON DROP IN FUEL CELL.

CASE L.RUN.D2
WASTE HEAT OUT
1 MAR 93 14:03:06

L.RUN.D2 HIGH TEMP. CO2 PASS ADDED TO BASE CASE, COMPARE L.AIR.4E

TABULATION OF WASTE HEAT AVAILABLE FROM PROCESS BOILERS.

B	1420.0 DEGREES F.	231.42 MILLIONS BTU	8 COOL HOT SOUR RAW GAS IN WHB; RAISE STEAM.
15	690.0 DEGREES F.	217.21 MILLIONS BTU	15 COOL QUENCHED RAW GAS IN WHB; MAKE STEAM.
20	370.0 DEGREES F.	33.51 MILLIONS BTU	20 COOL RAW GAS; PRE-HEAT BOILER FEED WATER.
36	855.0 DEGREES F.	256.75 MILLIONS BTU	36 HIGH-TEMPERATURE CO2 REMOVAL FROM SHIFT I EXIT GAS.
41	600.0 DEGREES F.	19.83 MILLIONS BTU	41 COOL INTER-STAGE SHIFT GAS IN WHB; RAISE STEAM.
75	950.0 DEGREES F.	226.38 MILLIONS BTU	75 COOL CATHODE RECYCLE GAS IN WHB - RAISE STEAM.
80	890.0 DEGREES F.	112.96 MILLIONS BTU	80 COOL ANODE EXIT GAS IN WHB; RAISE STEAM.

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CASE 9

Employ High-Temperature Sulfur Removal

TAPULATION OF SHAFT ENERGY AVAILABLE OR NEEDED.

102	-34573.0	KILOWATTS	AIR COMPRESSOR FOR SUPPLY TO CRYOGENIC OXYGEN PLANT
104	-14287.0	KILOWATTS	CRYOGENIC OXYGEN COMPRESSOR FOR SUPPLY TO COAL GASIFIER
2	-4544.0	KILOWATTS	U-GAS GASIFIER NORMAL OPERATION. STREAM #2 = NET MAKE.
112	-59.0	KILOWATTS	OXYGEN COMPRESSOR FOR REGENERATING SPINELS FOR HOT H2S DELETE
113	-1775.9	KILOWATTS	SUPPLY OF #3. T/O EXTRA OXYGEN FOR HOT H2S REMOVAL.
114	-567.1	KILOWATTS	COMPRESSOR FOR RECYCLING GILBERT GAS FOR HOT H2S REMOVAL.
16	41814.2	KILOWATTS	16 INWARD SULFUR-LEAK RAW GAS TO FUEL CELL PRESSURE.
27	-14055.0	KILOWATTS	27 COMPRESS COMBUSTION AIR TO FUEL CELL PRESSURE.
35	56525.2	KILOWATTS	NORMAL FUEL CELL, 1200 F INPUT SETS VOLTAGE, 1260 F EXIT.
42	-25010.1	KILOWATTS	42 CATHODE RECYCLE GAS CIRCULATION FAN.
46	-6617.3	KILOWATTS	46 ANODE EXIT GAS BOOSTER FAN, TO FEED CATALYTIC BURNER.
43	243042.0	KILOWATTS	43 EXPAND CATHODE EXIT STACK GAS TO RECOVER POWER.
55	-14170.6	KILOWATTS	55 COMPRESS COAL FEED INJECTION GAS, A SMALL RECYCLE.

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CASE L.PUN.C3
STEAM LIST
O MAP 93 13:51:32

L.RUY.C3 HIGH TEMP. H2S : DUST REMOVAL, MAKE MOLTON SULFUR FROM OXYGEN.

TABULATION OF STEAM NEEDED IN PROCESS.

2	550.0 PSIA	2000.0	WOLS STEAM	2	U-GAS GASIFIER NORMAL OPERATION. STREAM #2 = NET MAKE.
14	1200.0 PSIA	127.0	WOLS STEAM	14	ADSORB. H2S ONTO ZINC SPINEL; A NEW IGT DEVELOPMENT.
05	1200.0 PSIA	50.0	WOLS STEAM	15	RECLAIM Zn SPINEL (OXIDE) BY OXYGEN; MAKE SULFUR.

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CASE L.PUN.C3
 WASTE HEAT OUT
 9 MAR 83 13:31:33

L.PUN.C3 HIGH TEMP. HIS 3 DUST REMOVAL, MAKE WOLTON SULFUR FROM OXYGEN.

 TABULATION OF WASTE HEAT AVAILABLE FROM PROCESS BOILERS.

2	1503.0	DEGREES F.	192.13	MILLIONS BTU	P COOL HOT SOUP RAW GAS IN WMB; RAISE STEAM.
14	1200.0	DEGREES F.	-27.77	MILLIONS BTU	14 REFORM 425 ONTO ZINC SPINEL; A NEW IGT DEVELOPMENT.
15	1400.0	DEGREES F.	-71.81	MILLIONS BTU	COOLING COIL IN FLUID BED SPINEL REGENERATOR.
15	2600.0	DEGREES F.	11.23	MILLIONS BTU	SULFUR CONDENSER WHICH GENERATES 15 # STEAM.
19	750.0	DEGREES F.	32.93	MILLIONS BTU	19 COOL SULFUR-CLEAN RAW GAS; MAKE STEAM.
41	310.0	DEGREES F.	608.42	MILLIONS BTU	41 COOL CATHODE RECYCLE GAS IN WMB; RAISE STEAM.
45	350.0	DEGREES F.	254.31	MILLIONS BTU	45 COOL ANODE EXIT GAS IN WMB; RAISE STEAM.
53	200.0	DEGREES F.	13.24	MILLIONS BTU	53 COOL SMALL COAL INJECTION RECYCLE STREAM #20; MAKE STEAM.

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APPENDIX Ca. Original Efficiency Comparison

ORIGINAL EFFICIENCY COMPARISON

During the course of the program, 66 separate cases were examined for the integration of coal gasification for molten carbonate fuel cells for power generation. Of these, 20 cases were analyzed for variations in system efficiency by modification of process configurations. These cases were undertaken early in the program, on the basis of a constant coal feedstock to the system with varying power production, and are discussed in this appendix of the report. Additionally, nine of these 20 cases were revised to a constant power output basis, with slight modifications in system parameters to make them more truly comparable, and analyzed on an economic basis in the main body of this report. Twenty additional cases were evaluated to indicate the effects of primary assumptions upon the total system efficiency; these cases are presented in Appendix Da. The remaining cases are not specifically presented. Many of these represent minor variations in a specific analysis or involved short optimization studies by varying a single process parameter. In these cases, the results are discussed below but not specifically analyzed in depth.

Many of the cases were reanalyzed periodically during the program. As the output from later systems was assessed, it became apparent that the overall system was extremely sensitive to minor variations in specific inputs. Therefore, earlier cases were reassessed, with particular attention paid to these inputs, so that the results would be more comparable. An example of the complete printout for a single case is included as Appendix Cd in this document, to illustrate the types of calculations that were made to integrate the entire system. Complete printouts of the major systems analyzed were placed on file with the prime contractor; however, only summary sheets are included (Appendixes Cc, Db, and B) with this document because of the bulk that would result.

Efficiency Comparison of Alternative Systems

The primary charge of the project was to evaluate alternative systems that might be conceived for integration of coal gasification with power generation by molten carbonate fuel cells. Emphasis was to be placed on identifying development effort that may be cost-effective in reducing the overall cost of future power generation. Twenty systems, incorporating various features of this integration, are summarized in Table Ca-1.

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Table Ca-1. INITIAL EFFICIENCY COMPARISONS
(Coal Feed - 4991 Tons/Day)

Case	System	Power Output		Eff, %	Comments		
		Net Mech	Fuel Cell				
		MW					
		Net	Steam	Net			
I	Base Case	7	537	111	656	42.1	
Id	Heat Recuperation	12	534	105	650	41.8	
Ie	Add Quench, Shift	18	538	64	620	39.8	
If	Move Turboexpander	40	542	56	637	40.9	
II	Fluidized-Bed Gasifier	24	490	71	585	37.6	
III	Fluidized-Bed, Reform	49	635	11	695	44.7	3.7% CH ₄ in raw gas
IV	Fluidized-Bed, High CH ₄ , Reform	51	651	4	706	45.4	6.1% CH ₄ in raw gas
IVa	Conceptual CH ₄ , Reform	67	698	-20	746	47.9	9.7% CH ₄ in raw gas
IVb	70% Conversion	75	543	33	651	41.8	
IVc	80% Conversion	60	615	13	688	44.2	
IVd	90% Conversion	46	684	-7	723	46.5	
IVe	Optimized IV	60	651	1	712	45.8	
IVf	Low Quench	49	658	12	720	46.2	
IVg	No Quench	45	656	28	729	46.8	
V	Dust-Tolerant Shift Reactor	56	653	30	739	47.5	5 Cases
VI	High-Temperature Dust Removal	78	655	3	736	47.3	3 Cases
VII	Sulfur-Tolerant Cell	54	639	94	767	49.3	
VIII	High-Temperature CO ₂ -Transfer	7	693	50	750	48.2	2 Cases
IX	High-Temperature Desulfurization	63	639	78	780	50.1	3 Cases
X	Anode Tail Gas Recycle	13	805	-143	675	43.4	

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Additional systems were also analyzed, but are not presented in depth, either because they showed little potential or were primarily optimization studies on the systems included.

This appendix presents the results of the initial analysis (after several iterations for comparability) of the twenty systems. The cases all present the power output obtained from a constant 4991 tons/day of coal — the amount required for 675 MW of power from an early iteration of the base case. The analysis reflects the learning experience that was encountered as the project progressed. The cases are closely comparable, based upon knowledge gained on the system sensitivity. However, the final iteration, presented in the body of the report, shows slight discrepancies from the analysis below.

Case I — The Base Case

The prime contractor prepared a reference design, derived from evaluations by others, as a starting point for this systems analysis effort. That contractor's report, together with its economic analysis, is included as Appendixes Aa and Ab. The discussion of the system in that document is detailed; only the major features of the system are described below, together with deviations and evaluations used in the current analysis.

The basic configuration of the base case is presented in Figure Ca-1. In that base case, an entrained-bed gasifier produces a high-temperature raw gas. The gas is cooled by waste heat recovery to high-pressure steam. The gas is further cooled by "regenerative" heat recovery (to the purified gas). After scrubbing for dust removal and condensation, the gas is treated for selective H₂S removal at ambient or reduced temperature. The sweetened gas is reheated by the raw gas in the heat exchange described above. Because not all of the heat available in the raw gas is required for reheat to the purified gas, additional high-pressure steam is recovered by parallel cooling of the raw gas. The hot, purified gas is turboexpanded to the fuel cell operating pressure, humidified for carbon control, and fed to the fuel cell. After 85% conversion of the hydrogen and carbon monoxide within the fuel cell, the spent anode exhaust is catalytically combusted with hot, compressed air and fed to the fuel cell cathode. An additional cathode feed is a recirculating stream derived from the cathode exhaust and cooled (to high-pressure, superheated steam production) for fuel cell temperature control. The remainder of the spent cathode gas is exhausted through a turboexpander and heat recovery to

the incoming air. Isentropic compressor efficiencies are 88% and expander efficiencies are 92%. Other assumptions used in the analysis were --

1. The fuel cell operating pressure was taken as 150 psia so that the output from the process simulator would closely approximate fuel cell data. (See "Assumptions" section of main text).
2. The spent cathode exhaust was discharged from the stack at approximately 500°F to avoid potential plume formation.
3. The gas purification by a selective Selexol system was specified for 19% H₂S in the acid-gas, a value that has been found to be a satisfactory feed for split-flow Claus sulfur plants. No effort was made to recover COS in the acid-gas removal system, because of the high steam and power demands of the Selexol system when applied for this duty.
4. Steam addition for carbon control, feeding the fuel cell, was selected as midway between the theoretical requirements for inhibition of graphite and amorphous carbon deposition.
5. The power production from the high-pressure steam system, with reheat, was taken as 35.28% efficiency to include factors for boiler feedwater pumping and the cooling water system.
6. The precise estimation of temperature levels available for high-pressure steam production and reheat was not attempted because of the additional energy production and demands within the overall system (sulfur plant, tailgas system, water stripping, ammonia recovery, and water treatment system) have not been quantified.

Evaluations of the system include --

1. The overall system, with the conservative approach indicated above, calculates as 42.1% efficient, compared with 46.7% calculated by others for this system. Power generated in the fuel cell stack in the current design is much greater, proportionally, than that shown by others; the recovery of waste heat is reduced because of the more conservative design employed.
2. The entrained-bed-gasifier design appears optimistic. The design assumes the capability to pump slurry at 70% feed solids concentration, and the oxygen demand is relatively low, particularly for the 100% carbon utilization claimed.
3. The raw gas leaving the gasifier is thermodynamically unstable for carbon deposition when cooled to the temperatures of waste heat recovery.
4. The temperature profile of the "regenerative" heat exchange system indicates a degree of countercurrency not often encountered in chemical engineering practice, particularly when operating with hazardous, toxic gases at elevated temperature and pressure.

5. The heat balance indicates major energy losses to latent and sensible heat of the system exhaust, to steam cycle losses, and to fuel cell air compression.
6. The gasification efficiency of the coal in the entrained-bed gasifier is relatively low. Although 100% conversion of the carbon is assumed, the heating value of the gas is only 77.2% of the coal fed to the system.

Base Case Re-engineering

A number of cases were considered to ensure system operability for the base case. In particular, carbon stability of the raw gas was examined, the countercurrent heat exchange was converted to a safer operation, and the efficiency of rotating equipment was reviewed. Additionally, subsystems were added or moved within the process to achieve greater recovery of energy.

The first evaluation was the relative efficiency of the compressors and expanders within the system. The base case was reevaluated at three levels of assumed compressor and expander efficiency. It was found that an assumed 5 percentage point reduction in the efficiency of rotating equipment caused a decrease of 1.5 percentage points in the system efficiency. This efficiency loss was approximately constant for all efficiencies greater than 75%. The primary reason for the loss was the compressor. In multistage compressors, interstage cooling is used to condition the gas to an acceptable temperature prior to each stage of operation. With lower efficiencies of compression, more of the energy supplied as work of compression is removed as waste heat in the interstage coolers. With expanders, on the other hand, efficiency loss is not as critical. Energy that is not recovered to the shaft still exists as thermal energy in the expanded gas, to be recovered downstream in the steam cycle.

After consultation with experts in the field, it was determined that efficiencies approaching those assumed in the base case design would be achieved in the field, although guarantees offered by equipment vendors may be about 5 percentage points lower. Therefore, it was decided to employ the efficiencies of the base case for rotating turbo equipment throughout the remainder of the program.

Case Id evaluates more conventional heat exchange equipment, as applied to the base case design.

The reference design employs the use of highly efficient, apparently countercurrent, "regenerative" heat exchange. In the chemical process industries, regenerative heat exchange is defined as the alternating heating and cooling of inert solids. For example, refractory checkerwork may be heated by contact with a hot gas. After an interval of time, the hot gas is valved off, cold gas is valved in, and the heat stored within the checkerwork is transferred to the colder gas. Conventional nomenclature defines energy recovery by heat transfer through metal walls as "recuperative."

In the chemical process industries, conventional heat exchange is not countercurrent, but rather crossflow. The cost of double-pipe heat exchangers, required to approach countercurrency, is excessive, primarily because of the mechanical problems that arise from the thermal stresses encountered. The differential expansion of the shell and tubes must be considered and usually packing glands are employed. Such a design would be questionable with toxic, flammable, hazardous, high-pressure, high-temperature gases.

In crossflow heat exchange, normal engineering practice is to assign a temperature differential between the temperature to which the hot gas is cooled and the temperature to which the cold gas is heated. Thus, in the general case, crossflow heat exchange cannot approach the efficiency of the countercurrent heat exchange.

An exception for improved temperature recovery is the use of several crossflow heat exchangers in series. If the quantity of heat being transferred is large, and the surface area required for the heat exchangers is therefore also large, a number of heat exchangers (or shells) may be required. Rather than arrange these shells in parallel, a series configuration can tend to approach countercurrency in the heat exchange.

In the current system, because the quantities of heat being transferred are quite large, the series heat exchange approach is reasonable. Engineering judgment was used to assign a practical number of shells at each point in the heating and cooling of the fuel gas stream, and the system was evaluated with specific heat exchangers identified.

This approach necessitated an energy loss in the system, relative to the countercurrent exchangers; approximately 100°F of energy will be lost from the

gas stream to air cooling and/or cooling water for each heat exchanger train included. The overall system efficiency drops by approximately 0.3 percentage points when this type of heat exchange is included.

Case 1e was developed to evaluate the performance penalties encountered if conservative engineering were employed in the design to counteract potential carbon deposition from the gasifier raw gas.

Raw gas from the entrained-bed gasifier has significant thermodynamic potential to deposit carbon. When analyzed on the triangular C-H-O diagram, the raw gas composition contains relatively more carbon atoms that can theoretically exist in a thermodynamically stable gas. (See Appendix Cb for further discussion.) Although mechanisms may exist to inhibit carbon deposition in such a system, current prudent engineering practice would dictate that the gas should be treated to be stable with respect to carbon deposition before it is cooled to the temperature where this effect would be expected.

For this case, sufficient water was added to the raw gas to yield an overall gas composition that would be thermodynamically stable. The water was added as liquid. The quantity of water added to the quench was somewhat greater than actually required to control carbon deposition. The theoretical quantity of water was determined and a short optimization study was performed to find the preferred quantity of water for the most efficient overall system.

Rather than lose the hydrogen content of the water that was added for quenching the raw gas, this case also included a water-gas shift reactor on the raw gas to convert much of the water to hydrogen before condensation. For this effort, two stages of water-gas shift reaction were employed for a reasonable degree of water conversion.

The net effect of these counter-acting system changes was a decrease in the overall system efficiency by 2 percentage points, to 39.8%. The addition of the quench water removes heat from the high-temperature gas — heat that was recovered to the steam system in the base case. For this case, much of that heat was not recoverable to power. The use of the shift reactor, however, converts a portion of the water to hydrogen for eventual consumption in the fuel cell and reduces the quantity of steam that must be added just prior to feeding the fuel cell. As discussed in Appendix Cb, the addition of

steam at that point is a second-order effect, but higher hydrogen content to the fuel cell is a preferred option.

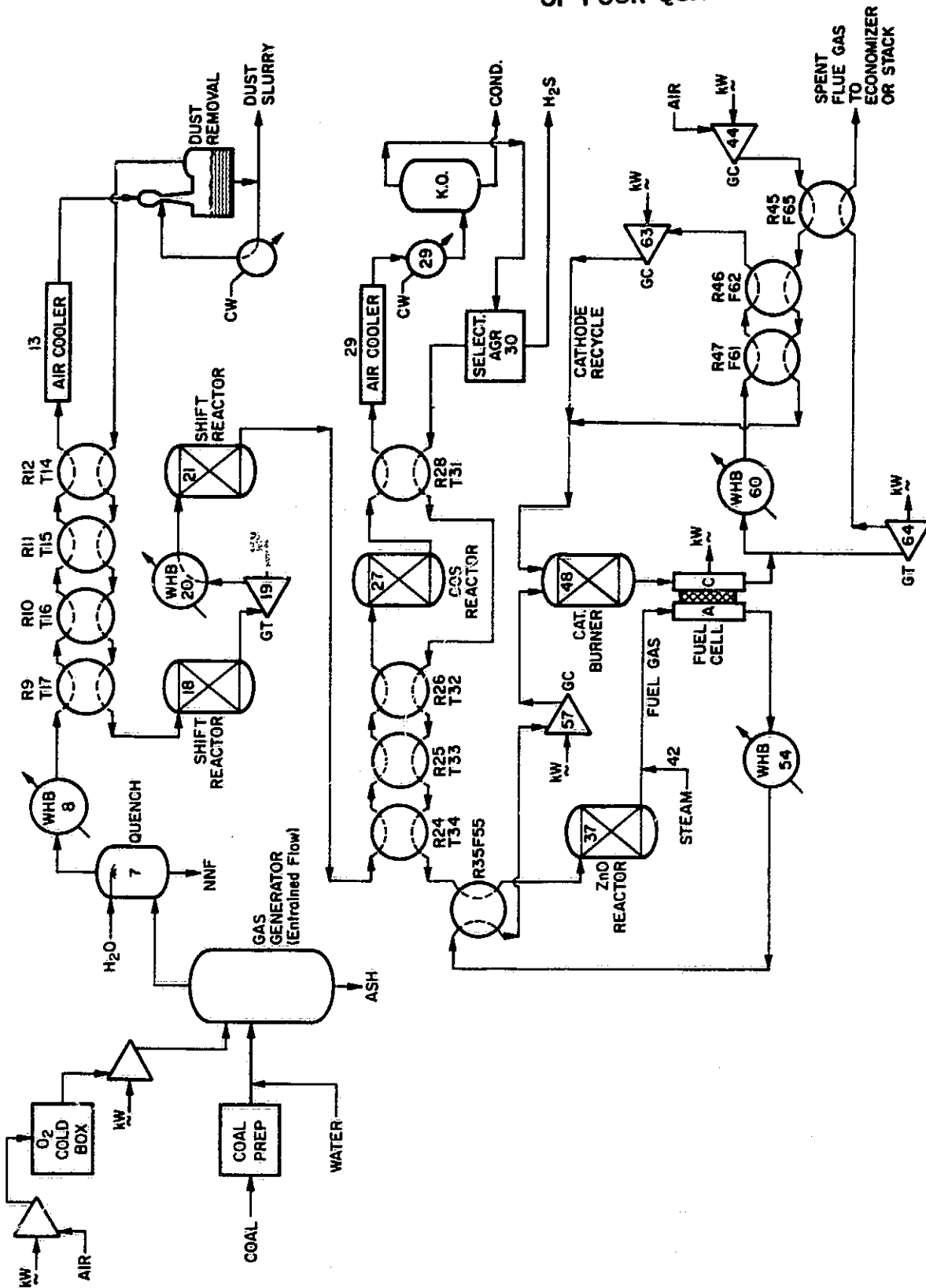
Case If was developed to indicate the impact of a preferred location for the turboexpander employed on the raw gas. In the initial base case design, the expansion turbine was located on the purified gas. With the inclusion of the shift reactor, as in Case Ie, the temperature of the gas leaving the first shift reactor is much greater than the gas temperature feeding the turboexpander. Additionally, this gas contains water vapor and carbon dioxide, at temperature and pressure, that are unavailable to the turboexpander in Case Ie. Consequently, Case If was designed to evaluate the system with the turboexpander moved to the discharge of the first shift reactor. Figure Ca-2 illustrates the configuration of the system.

With the improved temperature and gas quantity for the fuel gas expansion turbine, the system efficiency increased to 40.9%, approximately 1.2 percentage points lower than the original base case design. However, this case contains several system modifications for improved process operability, safety, and system control. Consequently, the use of raw gas quench, water-gas shift reaction, and improved location of the expansion turbine were used in further systems for analysis.

Case II — Fluidized-Bed Operation

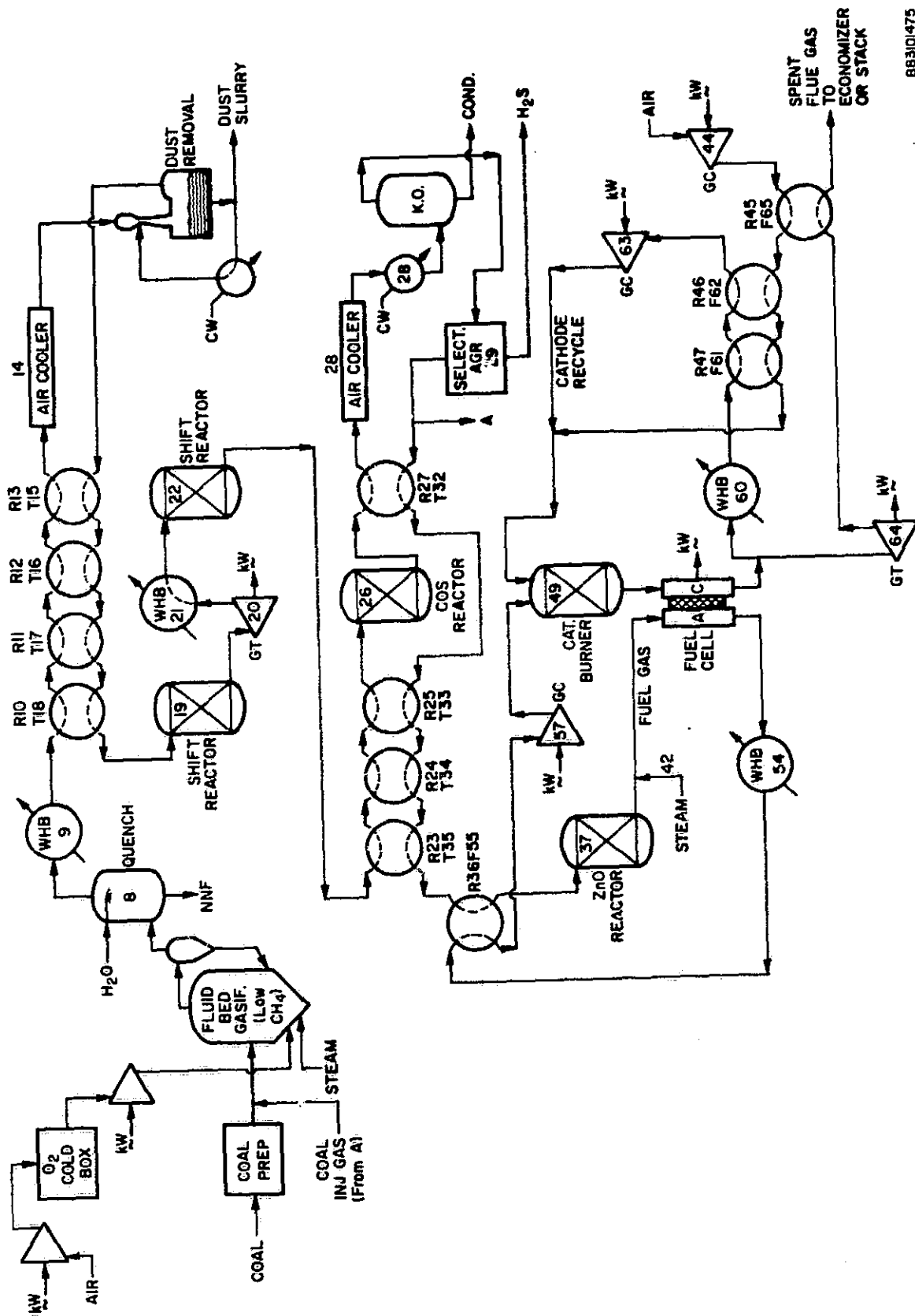
Case II was developed to evaluate the performance of a fluidized-bed gasifier within the system, rather than the entrained-bed gasifier used above. The overall processing scheme for the fluidized-bed gasifier is presented in Figure Ca-3; the scheme is essentially identical to that used in Case If, with the substitution of the fluidized-bed gasifier for the entrained-bed system.

The entrained-bed gasifier requires relatively high oxygen demands because of the high temperature of the product gas. Therefore, on the overall systems basis, more of the carbon in the coal is converted to carbon dioxide, rather than fuel gas species. The single-stage fluidized-bed gasifier should provide operating advantages because it requires less oxygen for operation. A fluidized-bed gasifier was specified to operate with relatively conservative design conditions. The operating pressure of the gasifier was selected as 500 psi to give a margin of safety relative to current limits of lockhopper



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Figure Ca-2. PROCESS FLOW DIAGRAM OF BASE CASE AFTER REDESIGN FOR OPERABILITY



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Figure Ca-3. PROCESS FLOW DIAGRAM OF INITIAL EVALUATION WITH FLUIDIZED-BED GASIFICATION

technology. Coal injection gas requirements were identified, as well as dust losses and carbon content in the ash, to assure a realistic design.

The chemistry for a single-stage coal gasifier, operating at 1875°F at 500 psi, indicates that approximately 8.3% of the carbon fed to the gasifier will be present in the off-gas as methane, if the gasifier is designed to minimize methane production. This methane content translates to 3.7% of the raw gasifier effluent. The summary printout of the gasifier design is included in Appendix Ce.

The gasifier, as expected, was somewhat more efficient than the entrained-bed gasifier. The steam/carbon ratio in the gasifier was assumed at 1.0 molar and the oxygen required, for the system to be in mass and energy balance, was 0.35 moles/mole of carbon in the coal. Approximately 3.1% of the carbon in coal was lost to the dust and ash losses; the higher heating value of the product gas is 82.4% of the heating value of the coal.

The downstream processing was similar to Case If and optimization studies showed that the preferred amount of water quenching of the raw gas, on a total system basis, should be minimized. For the conservative design, the quantity of water added at that point was taken to be midway between the water demand for carbon control to graphitic carbon thermodynamic equilibrium and amorphous carbon equilibrium.

Case II was over 3 percentage points lower in system efficiency, compared with Case If, which employed the same system components but with the entrained-bed gasifier. The fuel cell power production was considerably lower, because the methane fuel content of the gas was not available as power from the fuel cell, but rather was burned to the steam cycle. Even the steam cycle power output was not considerably higher than Case If, because of the steam demands of the fluidized-bed gasifier.

Alternative gasifier configurations have been developed that minimize the steam demand for fluidizing the gasifier. For example, some systems have employed recycle of product gas as a primary fluidization medium with only minimal steam addition for reaction control. However, in other systems analysis efforts, the gas composition in these configurations is much more unstable with respect to carbon deposition; therefore, with the greater energy penalty for additional quench, the overall system is not more efficient.

In the economic analysis work that followed, and is reported in the body of the text, the Case II system offers promise. Although the system with the fluidized-bed gasifier shows much lower efficiency than the entrained gasifier, enough capital savings was realized that the two systems are essentially identical in the cost of electricity. Therefore, the use of fluidized-bed gasifiers in conjunction with molten carbonate fuel cell power production appears to warrant further investigation.

Case III — Internally Reforming Fuel Cell

Case III was developed to evaluate the concept of a fuel cell that incorporates internal reforming of the methane to hydrogen and carbon monoxide, with thermal and chemical integration of the reforming with the fuel cell unit. Although the concept of an internally reforming fuel cell is relatively old, attention has again been focused on the operation of fuel cells that incorporate this feature. Active catalysts have been identified and evaluation is under way at several fuel cell laboratories.

The chemistry of the overall system is discussed in detail in Appendix Cb. That chemistry is summarized briefly below.

The steam-methane reforming reaction is the chemical recombination of methane and water vapor to produce carbon monoxide and hydrogen. Thermodynamically, high conversions are favored at the fuel cell operating temperature and pressure. When incorporated within the anode compartment of the fuel cell, the reforming reaction utilizes water vapor that is generated within the fuel cell, thus decreasing the back-pressure of water vapor in the Nernst concentration polarization effect. Additionally, the quantity of hydrogen in the cell is increased, further improving the polarization. The steam-methane reforming reaction is endothermic — 1000 Btu of methane heating value is converted to approximately 1300 Btu heating value in the product gases, requiring an input to the reaction of approximately 300 Btu. The excess heat released in the fuel cell can be used to drive this reaction, rather than be discharged to a steam cycle. On a total systems basis, the reduced heat load in the gasifier (with decreased carbon and oxygen consumption) results from the production of methane in the gasifier, but this methane can be converted to hydrogen fuel within the fuel cell, using excess heat from that cell.

As illustrated in Figure Ca-4, the configuration of Case III is essentially identical to Case II, with the incorporation of the reforming reaction within the fuel cell.

Thermal and chemical integration of methane reforming, within the anode cavity of the fuel cell, produces exceptional advantages when combined with a methane producing gasifier. Case III exhibited a full 7 percentage points greater efficiency than Case II, with most of the advantage occurring in the fuel cell output. The most inefficient form of power generation — the steam cycle — was reduced to a low level. It is concluded that the concept of an internally reforming, molten carbonate fuel cell, deserves increased attention.

Case IV — Operation of the Gasifier With Higher Methane Production

Case III, above, indicated high promise for a fluidized-bed gasifier integrated with an internally reforming molten carbonate fuel cell. The gasifier operating characteristics had been specified for relatively low methane production. By proper mechanical construction of the gasifier, the methane production can be increased. Such a gasifier is more efficient, has lower oxygen demand, and higher product gas heating value.

Case IV was developed to determine if the higher methane content from the product in the product gas would be desirable in the overall system. New gasifier material and energy balances (Appendix Ce) were developed for the candidate coal and the gasifier design that would convert 13.5% of the carbon in the coal to methane (6.1% methane in the product gas). This gasifier design was similar to that used by M. W. Kellogg in its evaluation of fluidized-bed gasifiers for SNG production.⁶ For the current case, conservatism in design was maintained: the steam requirements of the gasifier were maintained constant at 1 mole of steam/mole of carbon fed to the gasifier. Oxygen requirements of the gasifier decreased to 0.31 mole/mole of carbon, as required by the gasifier heat balance. Carbon lost to dust and ash was maintained at 3.1% of the feed and the heating value of the resulting product gas was 82.7% of that in the coal.

At this time, it is of value to review the gasifier operating parameters for those systems which were used in this study. Although two other gasifiers are used in later analyses, they are either conceptual or showed no real

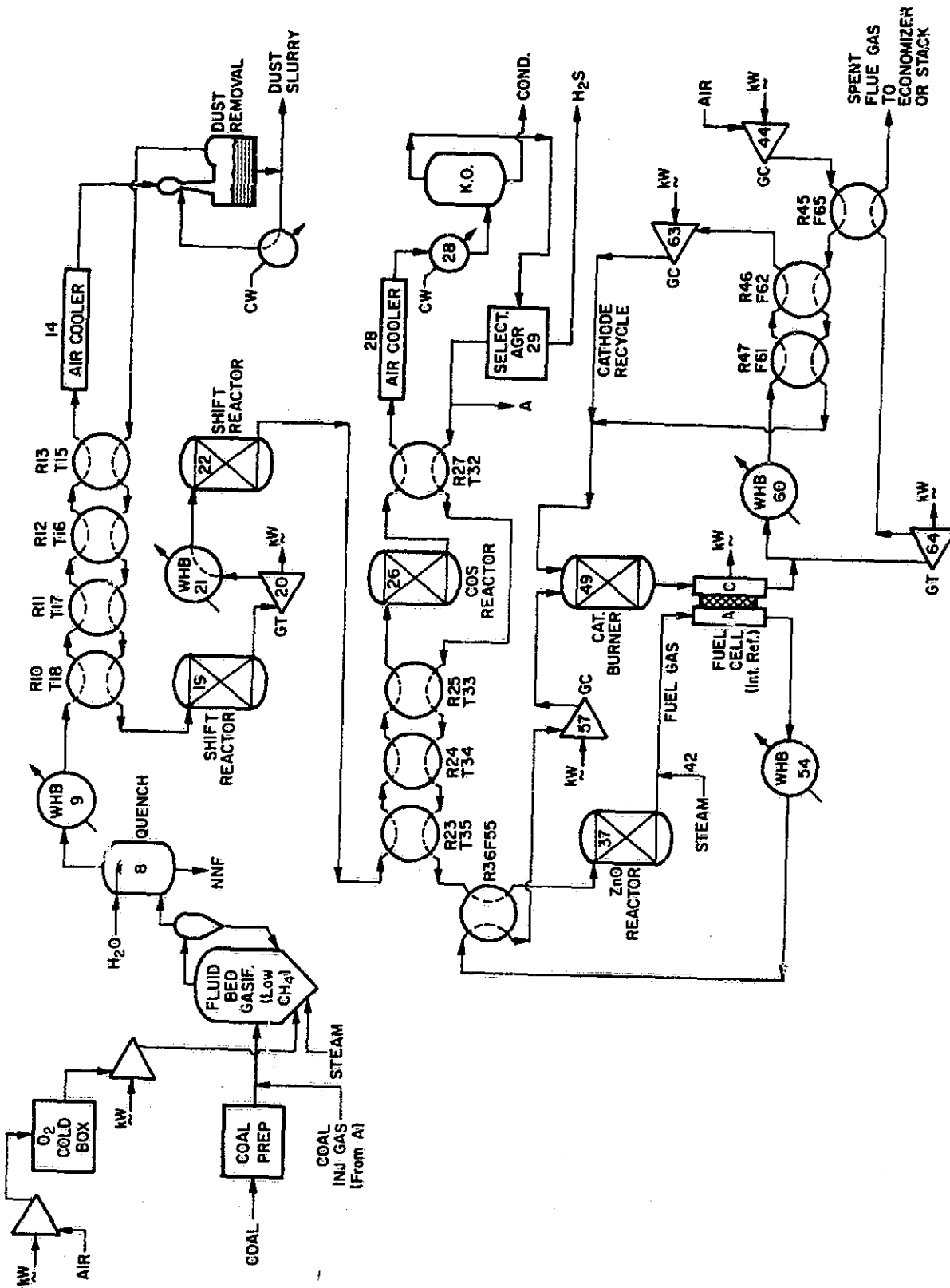


Figure Ca-4. PROCESS FLOW DIAGRAM INCORPORATING FLUIDIZED-BED GASIFICATION AND STEAM-METHANE REFORMING INTEGRATED WITH THE FUEL CELL

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operating advantage. Therefore, the three gasifiers discussed at this point are summarized below in Table Ca-2.

The fluidized-bed gasifiers have lower oxygen demand and convert a greater fraction of the coal heating value into the product gas that may be employed in the fuel cell. When one considers the overall chemistry of the system (Appendix Cb), the carbon in the coal is converted to hydrogen for consumption in the fuel cell through either the water-gas shift or the internal reforming reactions. The equivalent hydrogen produced per unit of coal is considerably higher for the fluidized-bed gasifiers and the gasifier producing increased methane appears to be preferred from this analysis. From the triangular C-H-O diagram of Appendix Cb, one notes that the fluidized-bed gasifier requires much less water addition for stabilizing carbon deposition; the gasifier with the increased methane production requires greater water addition than the low-methane fluidized-bed gasifier. This effect will counterbalance part of the advantage of the increased methane production.

The overall system considered for the analysis is identical to Cases I, II, and III, but incorporating the improved gasifier with the internally reforming fuel cell; it is illustrated in Figure Ca-5.

The increased methane content of the gasifier, combined with the decreased oxygen requirement, increased the overall system efficiency by about 0.7 percentage points. Thus, this system is apparently preferred, but not with the degree of improvement anticipated.

System Modifications

Several cases were evaluated to optimize the system of Case IV and to indicate the impact of operating variables. In an unreported analysis, the current density of the fuel cell was varied over a narrow range and the overall impact upon system efficiency was found to be minor. From this effort, a fuel cell current density of 150 mA/cm^2 was used in the remainder of the work.

In another optimization study, the air flow rate to the cathode was varied and found to be an important variable upon the system output. From this effort, the air flow to the cathode was set at 160% of stoichiometric requirements.

Table Ca-2. GASIFIER SUMMARIES

Product Gas	Entrained Bed	Fluidized Bed	
		Low Methane	High Methane
Temperature, °F	2450	1875	1875
Pressure, psia	600	500	500
Composition, mol %			
H ₂	29.4	29.3	25.8
CH ₄	0.1	3.7	6.1
CO	43.3	24.9	24.8
CO ₂	8.9	14.3	15.1
H ₂ O	18.3	27.8	28.2
HHV/HHV Coal	0.772	0.824	0.827
O ₂ Demand, mols/mol C	0.47	0.35	0.31
Equivalent H ₂ * HHV/HHV Coal	0.776	0.868	0.895

* Equivalent H₂ = (HHV H₂) (H₂ + CO + 4CH₄)/HHV Coal.

Case IVa was developed to evaluate the impact of a conceptual gasifier that is anticipated to produce greater methane contents than systems currently under development. IGT has initiated patent action on a modified gasifier configuration that, through control of operating conditions within the single fluidized bed, is anticipated to enhance methane production within the gasifier. The operating conditions have been identified and the methane content of the product gas has been estimated. Gasifier heat material balances were developed for the candidate coal and the gasifier was integrated into the system to indicate the impact. With the exception of the modified gasifier, the system was identical to that of Case IV, as illustrated in Figure Ca-5. The use of the conceptual gasifier improved the system performance by 2.5 percentage points, significantly higher than the earlier cases. Because the gasifier is conceptual, however, the system was not employed in later design evaluations.

Cases IVb, c, and d were developed to illustrate the impact of relative fuel conversion, taken within the fuel cell, upon the overall system efficiency. Apparently the system is not optimized at the 85% conversion that had been employed as a basis for this study — even greater efficiency can be achieved at 90% conversion efficiency across the fuel cell. Careful study of this case, however, indicates that the effect might be a result of the mathematical fuel cell simulator employed in the analysis, and these effects might not necessarily occur in practice. Therefore, the value of 85% fuel gas conversion efficiency (hydrogen, carbon monoxide, and methane) was maintained constant for the work in this effort. Also, this conversion efficiency was maintained to improve comparability to other analyses.

Case IVe is essentially identical to Case IV, but includes the results of several small optimization studies. It includes the preferred air flow to the cathode, as discussed above. It also incorporates a two-stage compressor on the cathode air supply. At an overall air compression ratio of 10, conventional practice would suggest that a three-stage compressor be utilized. However, by technology transfer from other compressor applications, it is known that outboard bearings and labyrinth seals may be used to permit higher gas discharge temperatures. With this modification, a two-stage compressor could be employed. This option permits increased heat recovery from the compression energy into the cathode air, permitting a slightly increased system efficiency of 0.4 percentage points.

Cases IVf and g, taken together with Case IV, indicate the impact of the amount of quench water used in the system. Case IV is based upon a conservative utilization of quench water, IVf employs minimum theoretical amount of quench water, and IVg assumes that quenching will not be required, perhaps because of the inhibition of carbon deposition by the presence of other species in the raw fuel gas. Note that if the quench operation could be completely eliminated, the process efficiency improves by 1.4 percentage points. Therefore, an in-depth evaluation of the carbon deposition phenomena, and the possible inhibition of carbon deposition, should be undertaken.

Case V -- Dust-Tolerant Water-Gas Shift Reaction

Case V was developed to indicate the potential for process improvement if the water-gas shift reaction could be operated on a dusty gas, rather than one that had been previously washed for dust removal. Five systems were evaluated incorporating a dust-tolerant shift reactor; the system presented as Case V in Table Ca-1 is the preferred variation. Catalyst configurations for such an operation have been identified and the penalty for inclusion of such a system is a moderate decrease in the space velocity required to achieve a specified degree of reaction.

Because this process modification was not selected for system costing and economic analysis, a separate drawing has not been prepared to illustrate the system. The system may be described as follows:

The raw gas from a fluidized-bed gasifier (taken from Case IV) is quenched for carbon stability and heat is recovered through a waste heat boiler to the required inlet temperature for the water-gas shift reaction. Although the gas has undergone primary solids removal through the cyclone associated with the gasifier, it still contains a dust loading that would have been washed out in prior systems. For this case, however, the dusty gas is fed directly to the water-gas shift reaction system. The shifted gas is cooled through a heat exchanger train; scrubbed for dust, ammonia, and excess water removal, and selectively desulfurized in an acid gas removal system. The purified gas is reheated in the heat exchanger train, expanded to fuel cell operating pressure, and humidified for carbon control before feeding the internally-reforming fuel cell. This system permits elimination of one gas washing train prior to the shift reactor, resulting in an energy saving equivalent to a 100°F temperature rise in the gas. The turbine operates on

the desulfurized gas; therefore, the mass of water vapor and CO₂ present in the inlet to the gas turbine has been reduced. However, the heat of the shift reaction is still available to the turbine through the recuperative heat exchange train.

The overall efficiency of the modified system is 47.5%, 1.7 to 2.1 percentage points greater than systems without the dust-tolerant water-gas shift reactor. The overall system is somewhat simpler than Case IV because of the elimination of one heat exchange train. Therefore, it is suggested that the concept of the dust-tolerant shift reactor should be investigated more thoroughly.

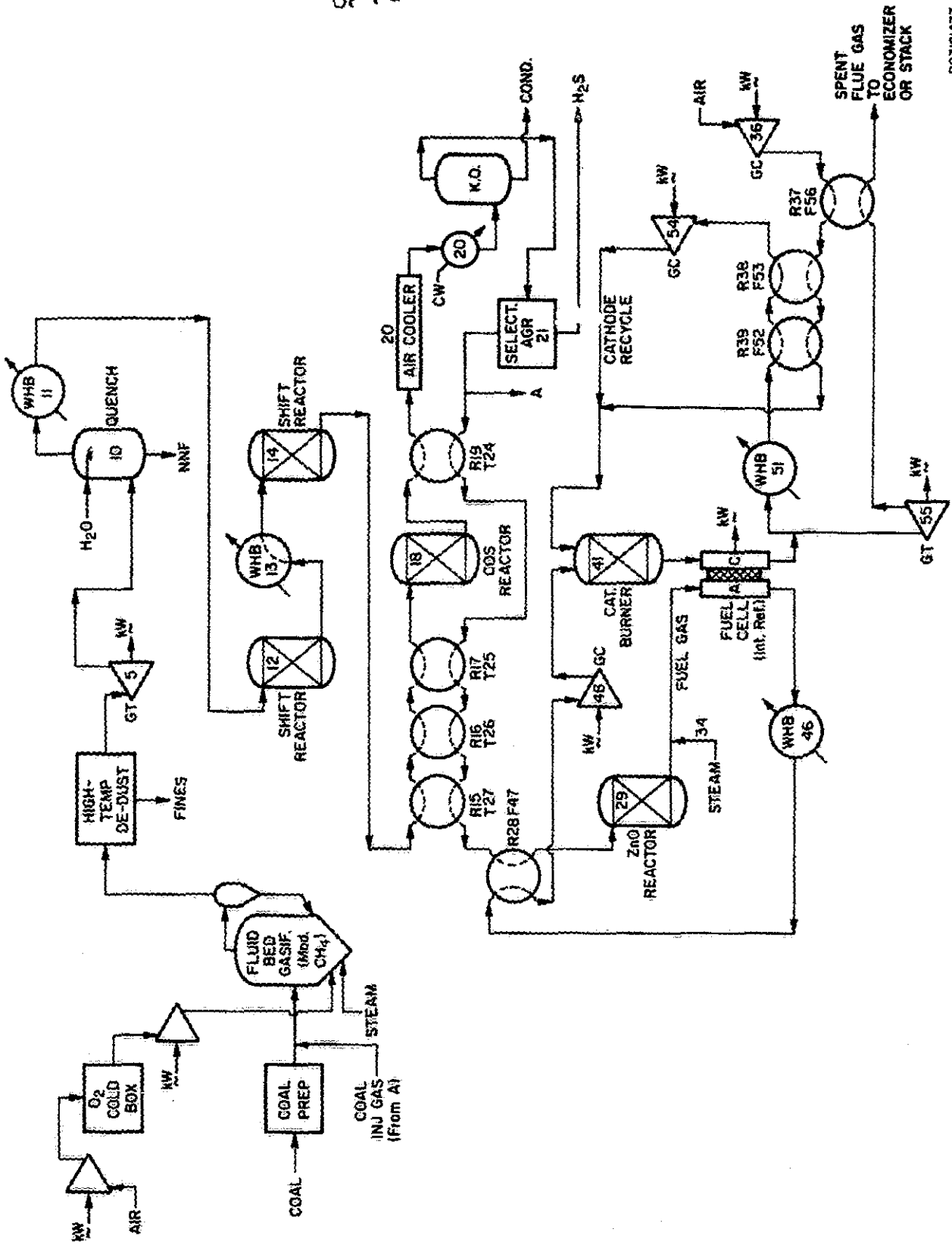
This system was not selected for economic analysis for the main body of the report. Similar system simplification and efficiency were determined for Case VI (below) and the economic evaluation of Case VI should be indicative of the economic advantages of the dust-tolerant shift reaction system.

Case VI - High-Temperature Dust Removal

Case VI was developed to indicate the potential for high-temperature dust removal in the coal gasification/fuel cell power generation system. Several techniques for high-temperature dust removal are being developed in various laboratories; the techniques are primarily being developed because of the need for this subsystem with pressurized fluidized-bed combustors when integrated with combined cycle power generation. The results of this analysis indicate that high-temperature dust removal is also applicable in the integration of coal gasification with molten carbonate fuel cell power generation.

Figure Ca-6 indicates that the high temperature dust removal subsystem considerably simplifies the overall process. Three cases were examined employing high-temperature dust removal; the preferred embodiment is shown in Figure Ca-6. In this system, the raw off-gas from the fluidized-bed gasifier (producing the moderate methane content of Case IV) is treated for high temperature dust removal, immediately after the cyclones that return most of the gasifier dust for regasification. The gas is then expanded. At this point, the gas temperature is high and the temperature drop across the turbine is not so great that carbon deposition is to be expected. Because of the higher temperature, the power recovered from this turbine is greater than in earlier systems. The gas then passes through quench, heat recovery, cooling,

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Figure Ca-6. PROCESS FLOW DIAGRAM INCORPORATING HIGH-TEMPERATURE DUST REMOVAL

and acid-gas removal before reheating through the single heat exchanger train. The fuel cell incorporates the internal reforming concept.

Note that the turbine could also be employed after the initial quench. In this case, the mass of gas that is passing through the turbine is increased, but the inlet temperature is reduced. After analysis, the placement of the turbine before quench appears to be a preferred embodiment of the system.

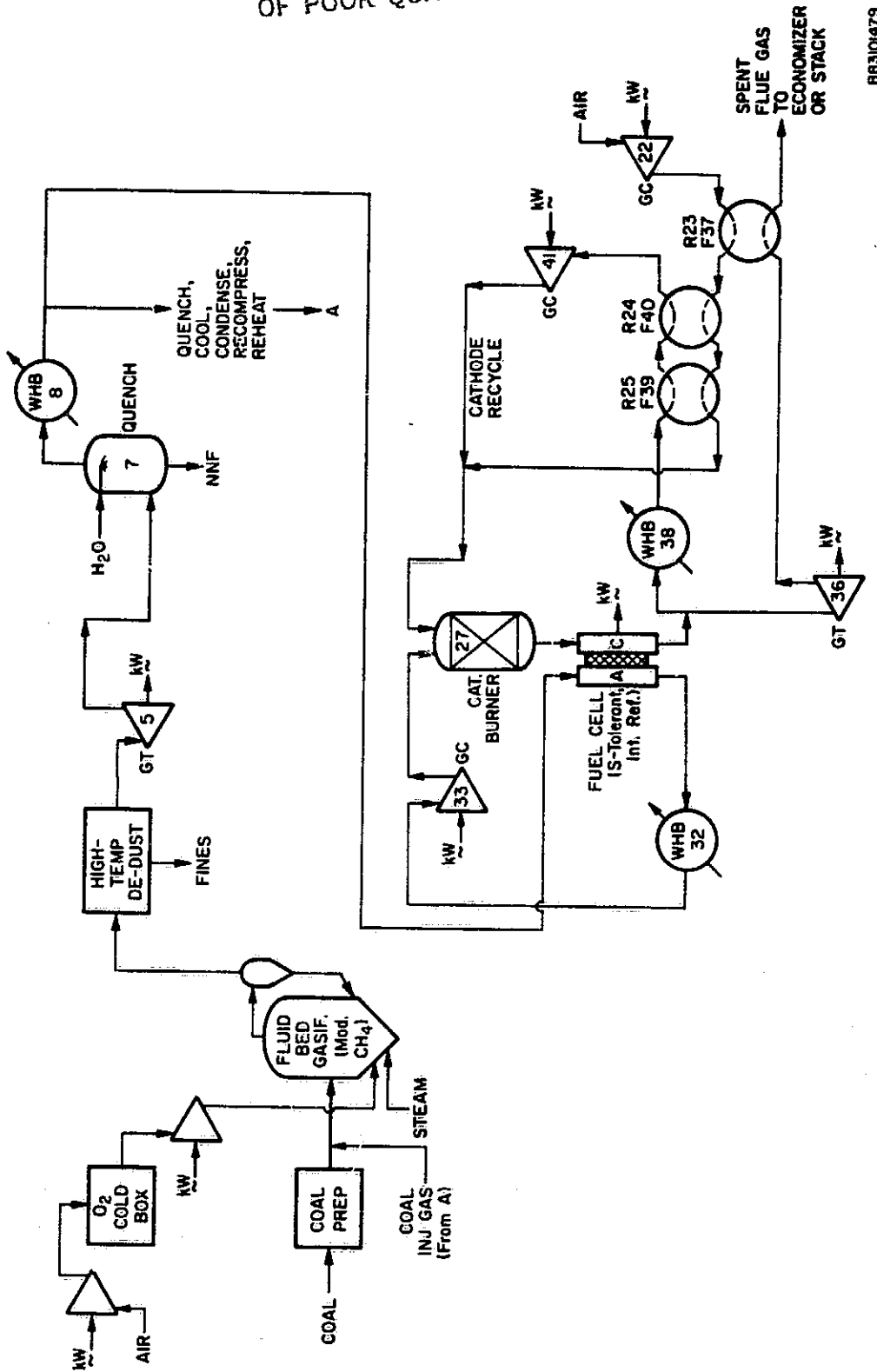
The overall efficiency of this system calculates at 47.3%, nearly identical to the dust-tolerant shift catalysis considered in Case V. On the overall systems basis, the concepts of high-temperature dust removal and dust-tolerant shift reaction appear to be complementary -- either system offers approximately the same value with approximately the same degree of system complexity. Both systems should be developed because each becomes a backup system for the other, improving the probability of success in the overall research effort.

Case VII -- Sulfur-Tolerant Fuel Cell

The overall system for integration of coal gasification with molten carbonate fuel cell power generation would have the maximum simplicity if the fuel cell could be modified to operate directly on the fuel gas after quenching and cooling to the fuel cell operating temperature, without sulfur treatment. Conceptually, this problem appears solvable -- sulfur-tolerant electrodes for this duty have been preliminarily identified and the electrolyte composition issue appears capable of resolution, based upon thermodynamic considerations.

The overall system (Figure Ca-7), employing a sulfur-tolerant fuel cell stack, is quite simple -- minimal heat exchange, no water-gas shift reactor, and no condensation of water from the gas. The fluidized-bed gasifier with the moderate methane make is employed, as is high-temperature dust removal and the internal-reforming capability within the fuel cell. The relatively high concentrations of carbon dioxide and water vapor in the fuel cell feed tend to suppress the fuel cell voltage and hence its power output. However, the overall system shows an excellent overall efficiency (49.3%), about 3.5 to 4 percentage points better than the systems with low-temperature sulfur removal.

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Figure Ca-7. PROCESS FLOW DIAGRAM INCORPORATING SULFUR-TOLERANT FUEL CELL

One note of caution should be considered in the energy analysis of this system: The debits and credits for SO₂ removal from the final stack were not considered (similar to the assumptions in the other cases that a final sulfur recovery from the acid-gas was not considered). Such stack desulfurization processes are relatively energy-intensive and may modify the relative ranking of this concept when analyzed in depth.

Final stack gas desulfurization, however, might not be required. Case IX (below) presents an analysis of a system that includes high-temperature desulfurization to the level required for current fuel cells. Such desulfurization capability has now been proved at only the laboratory scale. However, high-temperature desulfurization that is sufficient for environmental considerations has now been tested on larger equipment. Should that process be incorporated with a sulfur-tolerant fuel cell, the economic and system advantages would still be impressive. Therefore, it is recommended that efforts to develop sulfur-tolerant fuel cells should be undertaken.

Case VIII — High-Temperature Carbon Dioxide Transfer

A system has been conceived for pressure swing absorption of carbon dioxide from high-temperature, high-pressure gases. The sorbent would be regenerated by contacting with the compressed cathode air (which has a low concentration of carbon dioxide prior to being merged with other streams). Sorbents have been identified for this duty and the energy debits and credits for the sorption and regeneration phenomena have been calculated. The sorption of carbon dioxide takes place as shown in Figure Ca-8, following the water-gas shift reactor, where the partial pressure of carbon dioxide is the highest. Approximately 50% to 60% of the carbon dioxide is removed from this gas stream. The sorbent is regenerated by the cathode air; significant heat exchange is required to heat this air and provide a thermal balance for the system. Purge steam is passed through the sorbent bed on both sides of the sorption and regeneration cycles to avoid mixing of oxidizing and reducing species.

This approach permits transfer of sufficient carbon dioxide from the hot fuel gas such that humidification of the feed to the fuel cell is not required. Thus, the gas feeding the fuel cell has higher purity and the cell exhibits excellent voltage. Similarly, the carbon dioxide is transferred directly to the cathode where it is required for the chemical reactions.

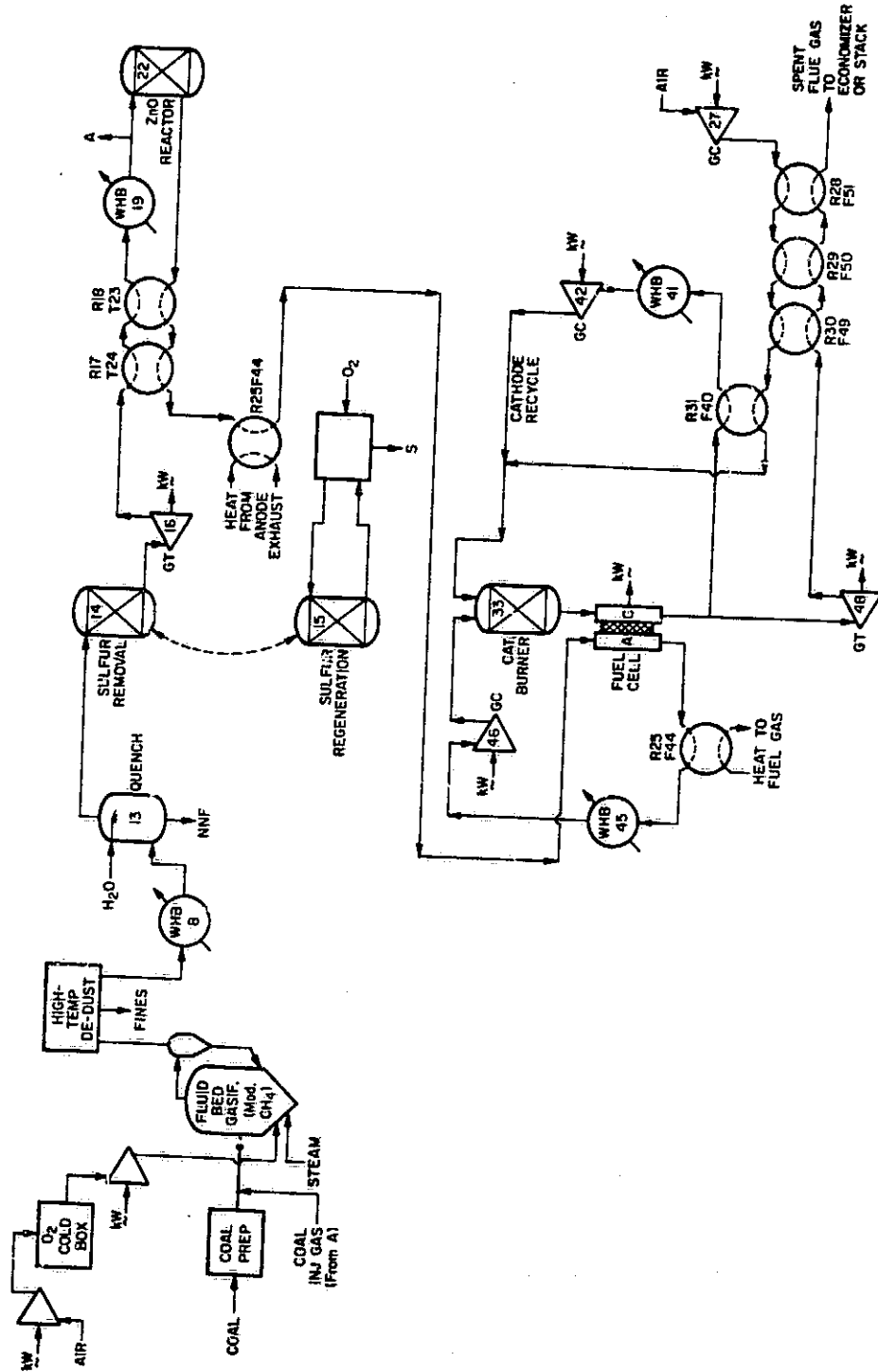
The system with CO₂ transfer is compared to Case IVe, which also uses a fluidized-bed gasifier of moderate methane production and the internally-reforming fuel cell. The impact of transferring the carbon dioxide is about 2.5 percentage points gain in system efficiency. As noted in the economics discussion in the body of the report, however, this efficiency gain approximately offsets the increased capital requirement for the processing subsystem. Nevertheless, it is believed that the concept merits additional study; the approach offers significant advantages in certain recycle systems that could not be evaluated within the constraints of the present program.

Case IX -- High-Temperature Desulfurization

Several concepts are under development at laboratories throughout the country for removal of sulfur from raw fuel gases at elevated temperature. The concept considered here is the mixed metal oxide system being developed at the Institute of Gas Technology. This system has particular attraction because the sorbent, when regenerated, produces elemental sulfur directly, rather than a dilute gas containing SO₂ that must then be concentrated and reduced. Three cases were considered in the evaluation of this process, varying the technique employed for regeneration of the sorbent and recovery of the sulfur values in the gas. In the system as shown in Figure Ca-9, the raw gas first passed through high-temperature dust removal, quench, and waste heat recovery to the operating temperature of the sorbent. The sorbent is anticipated to purify the gas to the degree required for the molten carbonate fuel cell. If the gas is not purified to this degree, however, sulfur guards are under development that will operate at elevated temperature. Nevertheless, the gas was then cooled through turbo expansion and a recuperative heat exchanger train to 750°F before zinc oxide treatment. After reheating, the gas passed to the internally reforming fuel cell to consume the fuel values that were produced in the methane-producing fluidized-bed gasifier.

As mentioned above, three systems were considered for regeneration of the loaded sorbent. The systems were engineered sufficiently to feel confident of the operability, and energy balances were developed. The system selected for Table Ca-1 produced elemental sulfur directly with essentially no tail gas stream. Note, therefore, that this system is not strictly comparable to the earlier systems -- the energy balance has included all the systems necessary for sulfur removal, recovery, and environmental control, in contrast to the other systems analyzed.

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Figure Ca-9. PROCESS FLOW DIAGRAM INCORPORATING HIGH-TEMPERATURE SULFUR REMOVAL

Although the gas feeding the fuel cell has high concentration of water vapor and CO_2 , the overall system efficiency is high — about 50%. In the economic analysis of the body of the report, this system shows a preferred cost-of-electricity and the concept appears to warrant additional research effort.

Case X — Anode Tail Gas Recycle

Early in the program, it was recognized that 15% of the fuel values in the raw gas were being exhausted to the steam cycle because the fuel cell could not be driven to 100% conversion of the fuel. A system was conceived for recycling the spent anode exhaust back through the condensation and acid gas removal systems, permitting recovery of this fuel to the fuel cell. In this case, the acid-gas removal was total — carbon dioxide was removed from the gas as well as hydrogen sulfide. The carbon dioxide was removed selectively and recompressed into the cathode to supply the necessary fuel cell reactants. In this manner, the feed to the fuel cell could be dry and nearly CO_2 -free, thus providing the maximum voltage from the fuel cell. Also, all of the hydrogen value in the fuel is recovered electrochemically in the fuel cell, except for a small purge stream taken to eliminate nitrogen from the recirculating gas loop. This purge stream was burned into the cathode exhaust to provide the maximum temperature for the expansion turbine.

Significant effort was expended in trying to develop an operable system on the simulator. The primary problem was the excellent performance of the fuel cell — insufficient heat was discharged from the fuel cell at high temperatures to bootstrap the system to the fuel cell operating temperature. Thus, extreme heat exchange was incorporated into the system to provide proper energy management. Nevertheless, this system showed relatively poor efficiency and was not included in the economic analysis. No illustration of this system is provided; because the system could not be improved sufficiently to merit economic analysis, an illustration was not developed. The overall concept of recycle systems, however, deserves additional attention. Several other recycle systems have been conceived that may offer greater merit than the one attempted here.

APPENDIX Cb.

Discussion of System Elements
and Interrelated Effects

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DISCUSSION OF SYSTEM ELEMENTS
AND INTERRELATED EFFECTS

General

The work effort described in this document is an evaluation of alternative means for production of electricity from coal using molten carbonate fuel cells.

The overall system involves 1) gasification of the coal, 2) purification of the fuel gas produced, 3) fuel cell conversion of the fuel gas to electricity, and 4) various bottoming cycles to recover heat from the system and convert that heat to power. Discussion of the major system elements are presented below.

The approach used in this program is systems analysis. In this approach, the operation of each element of the process is described mathematically and the subprocess units are integrated by machine computation to define the operating performance of the total system. With this approach, the individual subelements are not optimized; rather, their operating regimes are varied so that the overall system performs in a preferred mode.

This approach is contrasted to that which has been historically followed. For example, a gasifier manufacturer might define his preferred operating mode and the fuel cell developer might specify his preferred feed; however, the required interfacing might be so complex that the overall system is less desirable than alternatives. In the preferred case, compromises might be taken in the gasifier and/or fuel cell operating conditions to minimize the complications of interfacing.

As is shown in Appendix Da, the system is extremely sensitive to assumed changes in operating variables. Basically, because large quantities of energy are being exchanged, transferred, or transformed in the various steps within the overall system, slight changes in the efficiency of any subsystem can severely impact the overall output efficiency that, marginally, becomes the difference between two large numbers.

Therefore, to better understand the overall system and its interrelated effects, one must not only comprehend the first principles involved in each process element, but second-order effects are also important.

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The Fuel Cell

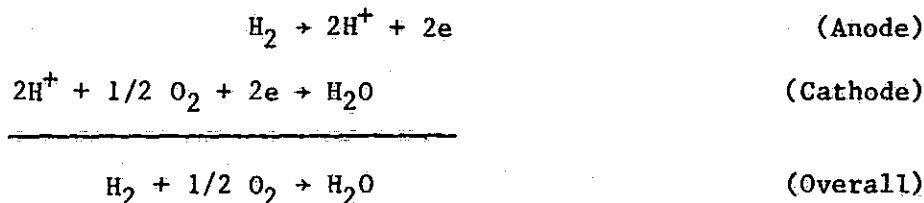
A fuel cell is a device that electrochemically converts the free energy of a chemical oxidation-reduction reaction directly into electricity. As such, it is similar to an electrical storage battery, except that reactants are continuously supplied to the fuel cell and the reaction products are removed from the system.

Because electricity is thermodynamically equivalent to work, the fuel cell effectively produces work directly from the chemical process, without the temperature limitations of a Carnot cycle.

The free energy of formation of water vapor in a fuel cell is about 79% of the enthalpy of formation. Therefore, an ideal fuel cell can generate 79 kWh of power from 341,300 Btu of hydrogen fuel. An ideal heat engine operating between 1200° and 95°F can only generate 69 kWh from the same amount of fuel. Further, real heat engines are considerably less efficient than their conceptually ideal models. Real heat engines, with all their associated parasitic auxiliaries, typically yield a net power output of 35% to 40% of fuel input.

The fuel cell operates by physically separating the two halves of a chemical redox reaction, forcing the necessary electron transfer to take an external path and do useful work. The two locations for the half-reactions are electrically interconnected by some intermediate of the chemical reaction, usually an ion in an electrolyte, thus closing the electrical circuit.

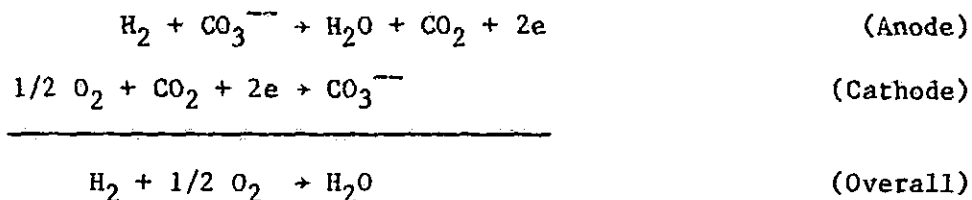
The general chemical reaction employed in most fuel cells is the reaction of hydrogen with oxygen to make water. In the simplest fuel cell, called the low temperature, acid fuel cell, molecular hydrogen reacts at the anode to produce hydrogen ions and electrons. At the cathode, the hydrogen ions react with oxygen to make water, removing electrons from the external circuit:



The two electrodes are separated by an acidic medium containing H^+ ions and the necessary electron transfer takes an external path between the electrodes.

The free energy of this reaction is approximately -55 kcal/g-mole (to produce water vapor), resulting in a theoretical open circuit voltage of 1.185 V/cell when conversion factors are included.

At higher temperatures (about 1200° to 1300°F) molten carbonate ion is the electrolytic transfer medium:



At elevated temperature, the free energy of the reaction is reduced to about 47.3 kcal/g-mole or 1.027 V theoretical open circuit voltage. Note that carbon dioxide must be transferred between the electrodes in order to maintain the system. This is the fuel cell reaction that is the primary subject of the current project.

The molten carbonate cell, as mentioned above, requires the recycle of carbon dioxide around the system. Heat management at elevated temperatures is more complex (and is the primary subject of this document); yet, waste heat from the system may be recovered to valuable use.

The fuel cell converts, at most, the free energy of the chemical reaction to electricity. The total enthalpy change of the hydrogen-oxygen reaction in the molten carbonate cell is 59.12 kcal/g-mole. If the fuel cell could operate without losses, about 79% of the heat of reaction could be withdrawn as electricity (equivalent to the free energy of the reaction of 47.3 kcal/g-mole). The remainder of the fuel heating value is released as heat. Note that any cell losses, as reflected by a decrease in cell voltage, also result in additional heat production.

The theoretical open circuit voltage of the molten carbonate fuel cell — 1.027 V/cell at 1200°F — is based upon free energy calculations and assumes that all reactants and products of the reaction occur at one atmosphere partial pressure. In real cells, however, these gases are usually present at different concentrations (or pressures) and this concentration changes as the gas is consumed (or produced) in the cell. In the anode compartment, the proper partial pressure ratio factor to be used is $(\text{H}_2)/(\text{H}_2\text{O}) \cdot (\text{CO}_2)$. A change in this ratio by a factor of 10 causes a change in the Nernst (or

concentration) potential of 91.5 mV at 1200°F. At the cathode, where no reaction products are formed, the partial pressure factor is $(CO_2)(O_2)^{0.5}$. In both cases, reducing the factor causes a reduction in cell voltage and an increase in the heat release.

In the fuel cell simulator used in this project, the log mean average concentrations (i.e., the log mean average of the concentration at the inlet and at the outlet of the gas compartment) are used for calculating the concentration polarization effects within the cell. This is a reasonable approximation to the losses, depending upon the direction of gas flow within the system. Concentration effects at the cathode were generally minimal in this program; the reduced concentration of reacting species was counter-balanced by the 10 atmosphere cell operating pressure. At the anode, however, the Nernst polarizations were typically 70 to 75 mV at the 85% fuel utilization.

Other losses exist within the fuel cell. Notable among these are the so-called electrode polarizations and internal resistance of the cell. The internal resistance of the cell is the resistivity to current flow through the cell and the migration of the electrolyte ions. The total losses here are proportional to cell current. The primary aspect of cell construction that affects the resistivity is the thickness of the electrolyte used within the cell. With the available fuel cell model, the internal resistance of the cell caused a voltage loss of about 49 mV at 160 mA/cm².

The electrode polarizations are a combination of factors. The true site for the electrochemical reaction is projected to be the area of those electrolyte pores that are wall-wetted (but not filled) with electrolyte. The reactant gases must dissolve into the electrolyte (perhaps as governed by Henry's law) and diffuse to the electrode surface for reaction, with counter-diffusion and volatilization of reaction products. These effects are controlled by the partial pressures of the gases within the electrode pore and the current density of the cell. Some investigators claim very low power losses for this effect; the fuel cell simulator used in this project reported losses that were comparable in magnitude to the anode gas polarization.

In the project effort, the fuel cell was simulated at a power density of about 150 to 160 mA/cm² and 85% fuel utilization, factors that had been previously selected as preferred from earlier investigations. Optimization

studies in this effort did not always agree with these previous studies; however, these parameters were held constant for comparability. The operating pressure of 10 atm was selected because of the reasonable agreement of the model to experimental data at that pressure.

The overall fuel cell potential, as estimated by the fuel cell simulator, varied from 954 mV for a fuel feed gas with low CO₂ and water concentrations to 815 mV for a system with relatively high contents of these species.

Currently, sulfur species are poisonous to molten carbonate fuel cells. When sulfur is present in the fuel gas, the nickel electrode material is deactivated, apparently by the formation of nickel sulfide. If sulfur is present in the anode gas, not only is the anode polarized but, because the anode exhaust gas is recycled after combustion into the cathode chamber, the cathode is also polarized. Additionally, sulfate ions form in the electrolyte, further degrading the overall cell performance. One of the major functions of the overall work effort is to find acceptable means for removal of sulfur from coal gas with minimum performance penalties.

For the total system, the maximum energy of the original fuel should be recovered in the fuel cell, because of its relatively higher efficiency, compared to the recovery of waste heat through Carnot-limited bottoming cycles. Directionally, certain effects should be expected in the fuel cell, based upon the fuel cell performance characteristics discussed above:

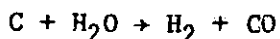
1. Higher-pressure operation is slightly preferred.
2. The fuel should be relatively dry and CO₂-free.
3. The presence of CO in the fuel gas has a mixed effect because it impacts (through the water-gas shift reaction discussed later) the relative quantities of the terms in both the numerator and the denominator of the Nernst effect ratio. Generally, it was found that lower CO concentrations are preferred.
4. If internal reforming of the methane present in the fuel can be accomplished, the Nernst effect is doubly adjusted to higher voltage.

The above effects are based upon operation of the fuel cell alone. When the fuel cell is integrated into the total system, counterbalancing effects may occur and the optimal operation of the fuel cell may not result in an optimal overall system.

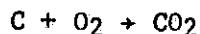
The Coal Gasifier

The overall system of conversion of coal to electricity, using molten carbonate fuel cells, requires a coal gasifier to convert the natural reductant — coal — into the species that is normally consumed by the fuel cell — hydrogen. As will be shown later, this function should be expanded to include the production of all gaseous species that may be readily converted into hydrogen for use in the fuel cell.

The primary chemical reaction in a coal gasifier is the steam-carbon reaction:



The above reaction is highly endothermic and occurs, at reasonable rates, at temperatures greater than about 1700°F. In most gasifier configurations, the endothermic heat of the reaction is supplied by burning a portion of the feed coal with air or oxygen:



The combination of the two primary reactions in the gasifier — the steam-carbon reaction and the carbon combustion reaction — yields an off-gas that contains primarily hydrogen, carbon monoxide, carbon dioxide, and water vapor. The water vapor is present because an excess of steam was used to drive the gasification reaction towards completion. The exact composition of the gas will be a function of the relative quantities of steam and coal that were used and the amount of oxygen required to sustain the reaction. Similarly, the proportions of the relative constituents will be governed by the chemical thermodynamics of the system at the operating temperature and pressure.

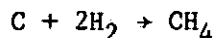
Based on heat balance, if carbon were the only constituent gasified, approximately 23% of the feed must be consumed with oxygen to supply the endothermic heat of the remaining 77% of the carbon that reacts with steam to form hydrogen and carbon monoxide. This carbon consumed by combustion is not a heat loss from the system but, rather, supplies the additional heat that is available in the fuel gases as heat of combustion.

Most gasifiers require more oxygen than the theoretical minimum oxygen/carbon molar ratio of 0.23 required to support the steam-carbon

reaction. The increased oxygen is required to raise the temperature of the gasifier system to the level required for satisfactory kinetics. The additional combustion of carbon results in sensible heat in the off-gases; heat that can only be recovered to Carnot-limiting thermal cycles, rather than as fuel hydrogen for the fuel cell.

The entrained-flow gasifier of this study employs an oxygen/carbon molar ratio of about 0.47 at the operating temperature of 2450°F. Thus, only slightly more than half of the feed carbon reports to the product gas as hydrogen and carbon monoxide. The fluidized-bed gasifier, operating at 1875°F, requires an oxygen/carbon molar ratio of about 0.35, converting much more of the coal into fuel gas species.

An alternative chemical reaction, termed hydrogasification, may also take place in the gasifier. In this reaction, carbon reacts directly with hydrogen to form methane.

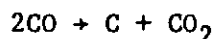


This reaction is exothermic. If it is promoted in the gasifier, it can help supply the heat for the steam-carbon reaction and displace part of the need for the carbon combustion reaction. In one conceptual gasifier evaluated in this study, the hydrogasification reaction is enhanced and the oxygen demand is markedly reduced, resulting in more efficient conversion of the feed coal into fuel gas species.

In the conventional molten carbonate fuel cell system, the methane produced by the hydrogasification reaction is not an acceptable fuel. Rather, the heating value of this material is only recovered to a thermal cycle. However, as discussed in detail later, the fuel cell may be integrated with steam-reforming of the methane to permit its utilization in the fuel cell for a more efficient overall system.

Gasifiers that promote the hydrogasification reaction are inherently more efficient because of reduced oxygen consumption (and burning of feed carbon to CO_2). Developmental gasifiers such as HYGAS[®], Exxon Catalytic, and Hydrane require little or no oxygen for supplemental heat, because of the very high methane production. Although conceptually advantageous in this system, these gasifiers were not assessed in the program because they generally add system complexity which, with the high leverage of plant capital on cost-of-electricity, would probably not be cost-effective at the plant scale used.

Note that it is possible to write the chemical reaction between carbon and oxygen to produce carbon monoxide rather than carbon dioxide. With this approach, all of the carbon value in the coal appears as fuel gas species. However, as will be shown later, any such gas produced is thermodynamically unstable at lower temperatures, tending to deposit carbon from the gas according to the Boudouard reaction:

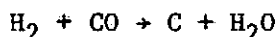


This effect will be discussed in more detail in the next subsection.

Other Chemical Reactions

Carbon Deposition

Gas compositions that may be thermodynamically stable at elevated temperatures may, upon cooling, become unstable and have high potential for carbon deposition. The two primary reactions that are responsible for carbon deposition from the coal-gas compositions considered are --



Note that both of these reactions are the reverse of gasification reactions and are interconnected by the water/gas shift reaction (discussed later).

The thermodynamics of the above chemical reactions have been evaluated; the stability of the gas mixture is a strong function of temperature, pressure, and the allotropic form of the carbon that is deposited. Figure Cb-1 is a representation of the carbon stability limits on the triangular C-H-O diagram for the deposition of amorphous carbon at 1200°F at the operating pressures encountered in this program. The raw off-gas compositions from typical gasifiers encountered in this program are indicated, illustrating instability of these gases at the operating conditions specified. In general, the maximum instability will occur at a temperature of about 1050° to 1200°F, depending upon the precise gas composition. The carbon deposition reaction is catalyzed, particularly by nickel (as found in stainless steel or fuel cell electrodes), and is also self-catalytic: Once carbon has been formed from an unstable gas mixture, the remaining unstable carbon falls out quickly.

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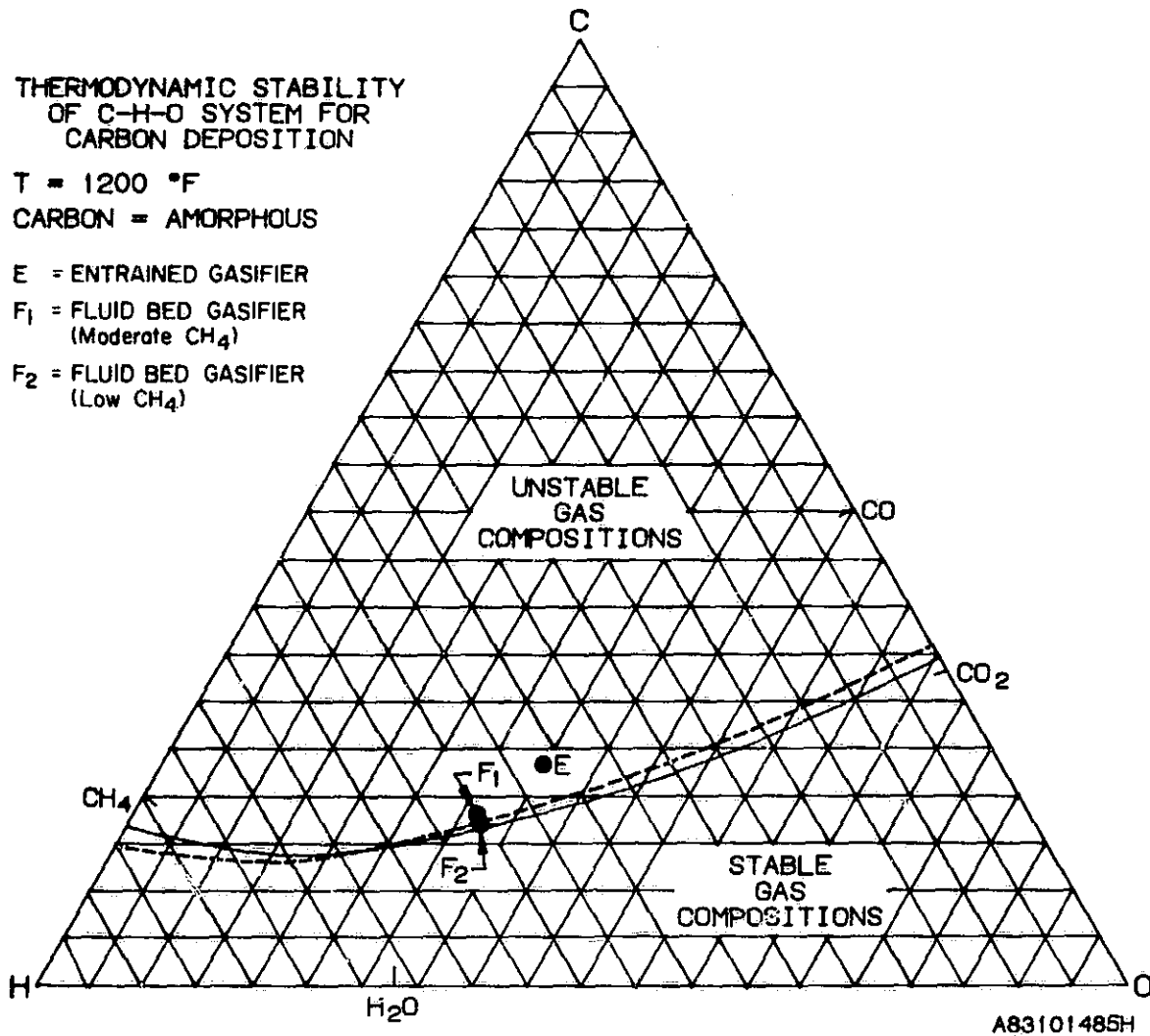


Figure Cb-1. THERMODYNAMIC STABILITY OF FUEL GAS MIXTURES

The usual technique to avoid carbon deposition is to add sufficient water vapor to the gas so that the second of the carbon deposition reactions above is driven to the left. Similarly, the first of the reactions does not occur because of the interdependence of the water-gas shift reaction in converting carbon monoxide to hydrogen. Figure Cb-2 illustrates the effect of steam addition on carbon deposition and further illustrates the impact of temperature on carbon deposition.

Most of the systems encountered in the current study include a water quench to add water vapor to the unstable gas and avoid the thermodynamic potential for carbon fallout. The water may have been added directly after the gasifier so that the gas, when cooled, would remain stable. Similarly, steam is added to the intermediate gas (which has been dehydrated in the course of sweetening) to avoid carbon deposition within the fuel cell.

When water is added to the hot gas after the gasifier, the system efficiency is impaired because, somewhere in the system, heat is added to liquid water to convert it to steam and the latent heat of vaporization is lost. Also, the steam absorbs some of the heat of the gasifier exhaust and part of this sensible heat is lost to cooling water.

When steam is added to the fuel cell feed, a similar effect takes place — the latent heat of vaporization is lost from the system, as is a portion of the sensible heat of the water vapor. Additionally, the water vapor present in the fuel cell feed depresses the hydrogen concentration in the feed and causes a Nernst concentration polarization.

Note, however, these effects are minimal if significant excess heat must be removed from the system as steam power. As discussed later in "turbine operations," the heat converted to steam power still loses the latent heat of vaporization of the steam. The expansion of the water vapor (in the fuel gas) through the various gas turbines in the system approximates the performance of the steam turbine, although it does not exhaust the heat at as low a temperature as the steam turbine. Generally, water vapor added to the fuel gas is a second-order effect for efficiency loss, when compared to a steam turbine, unless that water is condensed from the system under pressure and except for Nernst effects within the fuel cell.

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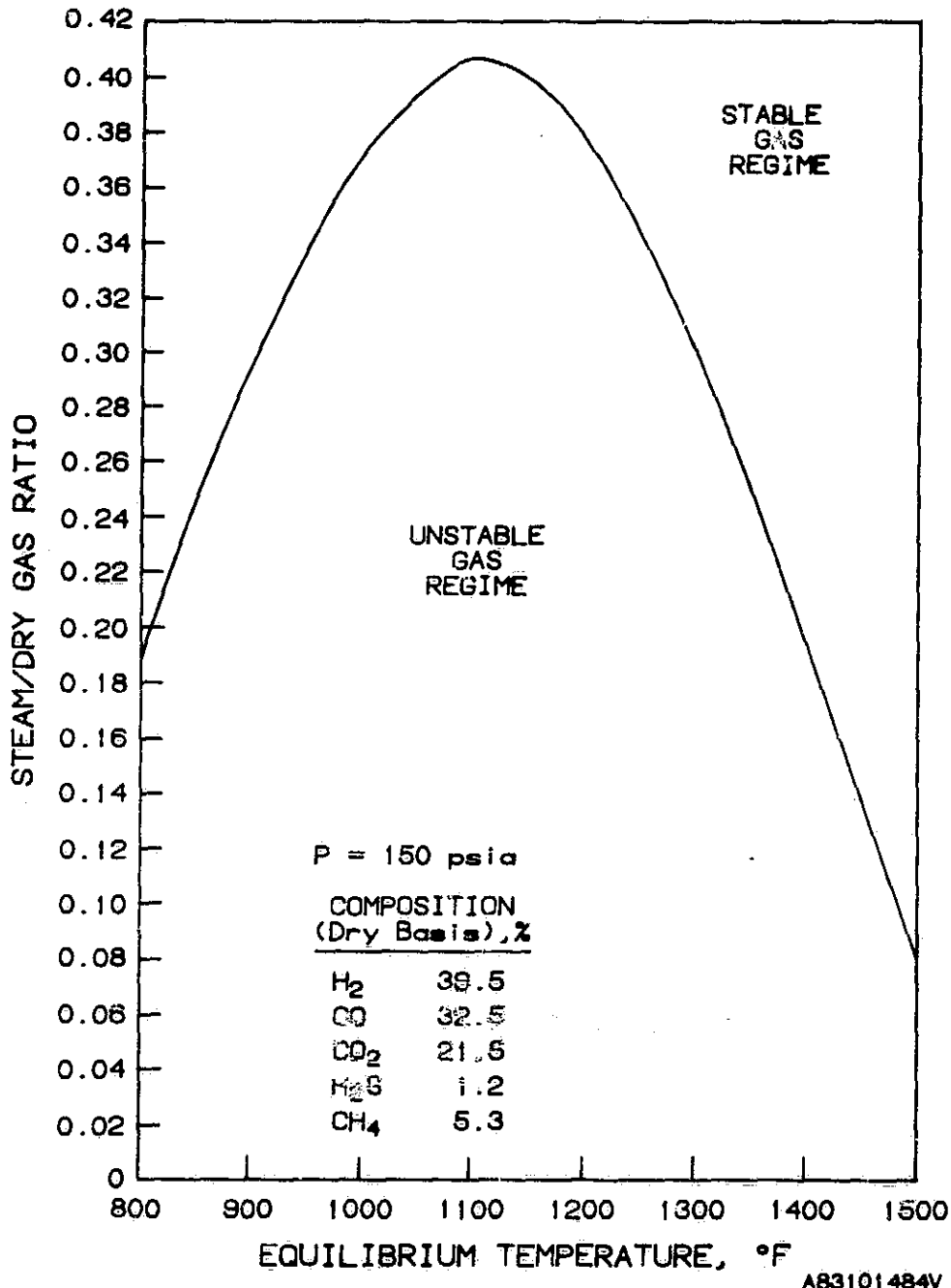
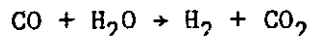


Figure Cb-2. EFFECT OF WATER ADDITION ON STABILITY OF A FUEL GAS MIXTURE

Water-Gas Shift Reaction

Most of the carbon monoxide produced by the gasifier is eventually converted to hydrogen for utilization within the fuel cell by the water-gas shift reaction:



This reaction is mildly exothermic. It may take place within the fuel cell, where the carbon monoxide is reacted with water vapor present in the feed or generated by the fuel cell reaction. If it takes place within the fuel cell, it alters both the numerator and the denominator of the concentration polarization equation. Generally, carbon monoxide is not preferred in the fuel cell feed because it dilutes the hydrogen.

If the water-gas shift reaction is included in the system prior to the fuel cell, a separate catalytic reactor is employed. To use available sulfur-tolerant water-gas shift catalysts, the gas must be heated to a light-off temperature of approximately 550° to 600°F. The heat of the reaction is recovered at approximately 600° to 800°F, if a separate reactor is employed, compared with the 1250° to 1300°F if the reaction takes place within the fuel cell. If a separate water-gas shift reaction step is included in the overall system, the water added for quenching the raw gas is partially converted to hydrogen, minimizing the effects of heat loss in the initial quench. The additional hydrogen in the sweetened gas improves the fuel cell voltage and reduces the need for steam in the fuel cell feed (as discussed above, a second-order effect). However, the higher hydrogen content of the feed gas changes the relative quantities of water vapor and carbon dioxide in the fuel cell exhaust and, hence, the final stack in the system. Thus, higher hydrogen content to the cell increases the quantity of latent heat that is lost to the stack. Overall, several counterbalancing effects occur and the impact of an external water-gas shift reactor is not obvious. In general, the results of the project indicate that the inclusion of the water-gas shift reactor improves the overall system performance, if the quench water is condensed in the system.

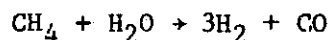
Carbonyl Sulfide Hydrolysis

The conversion of carbonyl sulfide to hydrogen sulfide by reaction with water vapor does not have a direct effect upon the overall system efficiency.

However, the relative quantity of carbonyl sulfide in the raw gas has an important bearing upon the performance of the acid-gas removal system and the quantity of zinc oxide that is required for final gas purification. This reaction system was therefore included in most systems to enhance the practicality of the simulation and reduce the power and steam demands of the acid-gas removal process.

Methane-Steam Reforming

Methane reacts with steam endothermically over nickel catalysts to produce hydrogen and carbon monoxide:



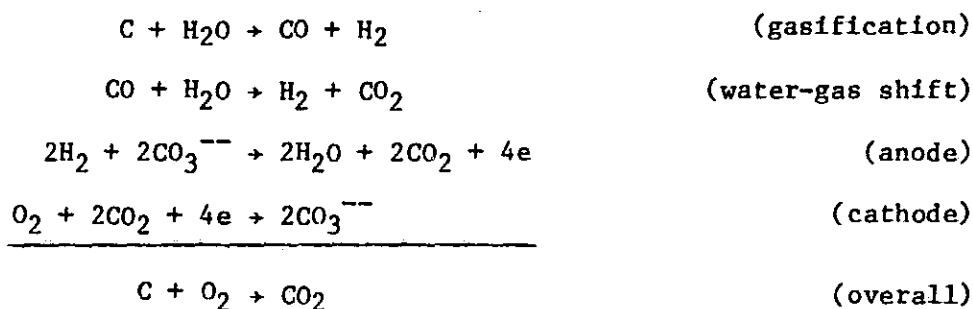
Significant heat is absorbed when this reaction occurs. At lower pressures, such as found in these system, the thermodynamics favor methane decomposition at temperatures greater than about 1200°F.

Thermodynamically, the steam-reforming reaction is ideally suited to be chemically and thermally integrated within anode compartment of the fuel cell. The fuel cell anode reaction produces the steam required to drive the reaction, and the cell provides the heat necessary to sustain the reaction. Several laboratories have been evaluating this integration: Catalysts to be incorporated within the anode chamber have been identified and tested. The overall system appears technically feasible at any condition where the fuel cell operating voltage is less than approximately 1.0 V, thus providing sufficient heat (total ΔH) to sustain the reaction. Consequently, in this effort, the reforming reaction was generally incorporated, chemically and thermally, within the anode compartment of the fuel cell.

Several other techniques are available for incorporation of the steam-reforming reaction within the system. The waste heat in the anode exhaust can be used to reform that exhaust, down to an equilibrium temperature level. Also, the reforming reaction can be incorporated thermally (but not chemically) within the stack, using non-active sections of the stack. These approaches were not employed in the current study because of the high promise of the integrated internally reforming system.

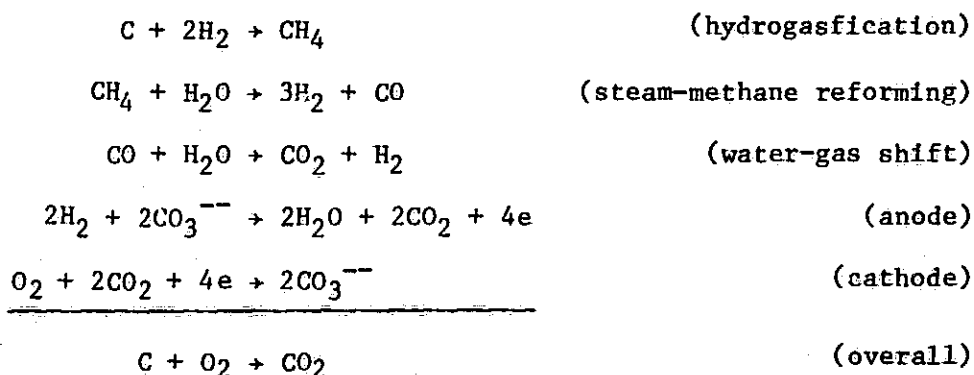
Summary of Chemistry

The total chemistry of the overall system can be summarized according to the reactions that occur. If the steam-carbon reaction is the only gasification mechanism, as encountered in entrained-flow gasifiers:



The overall reaction is the oxidation of carbon with oxygen to produce CO_2 . Intermediates include hydrogen, which is the material oxidized in the fuel cell. At perfect utilization, all the carbon monoxide is reacted by the water-gas shift reaction. Carbon dioxide must be recycled from the anode to the cathode and carbonate ion is an additional intermediate. Note that exactly as much water is produced in the system as is required for the chemical reaction. Note, however, that in the real systems considered, steam is raised and water vapor is discharged to the stack (including significant heat of vaporization). The techniques conceived for recycle of this vapor within the system could not be included within the constraints of the program.

If the hydrogasification reaction can be promoted within the gasifier, the reaction sequence changes to the following:



In this system, the sum of the first two reactions equals the first reaction of the previous set. However, the heat demand of the second reaction is supplied by the fuel cell, rather than by burning oxygen within the gasifier.

The individual species of H_2O and H_2 must either be recycled within the system or else the species must be provided externally. For example, the H_2 of the first reaction could be supplied by the first reaction of the earlier set of equations presented, if both reactions occur simultaneously.

Power Recovery Turbines

Although, on first examination, gas turbines appear to provide a means for recovery of pressure values of a gas into energy, they actually can be considered to operate on temperature. The value of the pressure is merely to define the temperature difference that can be recovered into power across the turbine.

In general, the highest temperature feed to the turbine will result in the highest total power output because of the Carnot laws of heat engines. The steam turbine is generally limited to approximately $1000^\circ F$ temperature on the input because of materials problems. However, the gas turbine concurrently operates at much higher temperatures. The combination of the gas turbine and the steam turbine — known as a combined cycle — is capable of much higher efficiency than the steam turbine alone because the heat discharged from the gas turbine can be recovered into steam and hence into the relatively low-temperature exhaust that corresponds to 2 to 4 inches of mercury absolute pressure in the condenser.

Consequently, on the overall system, it is preferred to operate the gas turbine at the maximum inlet temperature possible. The outlet temperature will be determined by the pressure ratio and the efficiency of the turbine. Note that the enthalpy difference extracted from the gas is directly translated into power, with only a small depreciation for mechanical efficiency, because the turbine inefficiency appears as heat in the outlet gas. Most of the heat in the outlet gas, eventually, is recovered into a steam cycle.

The steam cycle is depreciated in efficiency by the latent heat of vaporization of the water that must be converted into steam in the system. This inefficiency will exist, whether the steam cycle is used directly or the steam is injected into the process gas for eventual recovery in the system.

Consider the case of the steam that is injected into the sweetener gas, upstream of the fuel cell. Under the assumptions of the system, this steam is

raised at relatively high pressure and 1000°F. This steam is expanded across a power recovery turbine and injected into the gas stream to inhibit carbon deposition. In passing through the fuel cell, the mass of the steam is heated to fuel cell operating temperature and, eventually, exhausts with the pressurized cathode outlet gas. At this point, the mass of the water vapor is further expanded to atmospheric pressure, recovering power from part of the heat in the gas. Finally, the cathode gas, containing water vapor, is cooled through heat exchange and perhaps economizers before exhausting to the atmosphere. The final outlet temperature of the gas is 250° to 500°F, depending upon whether economizers are used or not. Relative to the same mass of steam passing through the standard steam cycle, the heat content of the anode gas injection steam has been recovered into power, or to reheat more water, to nearly the degree that would have been realized had the steam operated in a steam turbine alone. Thus, the injection of steam into the anode gas does not cause a direct loss of the power equivalent of that steam; rather, only a fraction of that power is lost because of the higher exhaust gas temperature. In other words, the effect of injecting steam into the anode feed gas is a second-order effect on efficiency as it impacts the overall system.

Summary

In general, one can often evaluate the impacts of a change in the gasifier/fuel cell system on first principles. For example, the use of a methane-producing gasifier, in conjunction with an internally reforming fuel cell, is a relatively straightforward application of first principles. The gasifier operates more efficiently, converting a greater fraction of the feed carbon into fuel gas species and consumes less parasitic power for the oxygen plant. The fuel cell converts its waste heat into reforming methane and maintains a higher concentration of hydrogen within the cell. Less heat is withdrawn from the fuel cell to the steam cycle; rather, that heat is used more effectively by producing hydrogen from methane.

In other systems, second-order effects occur and the solution is not so straightforward. For example, the incorporation of a separate, external shift reactor has several positive and negative effects. The summation of these impacts becomes minor in importance in affecting power outputs. Directionally, a separate water-gas shift reactor is a preferred alternative,

particularly if prudent design is followed, the raw gasifier off-gas is quenched for carbon control, and that quench water is condensed from the system.

Of particular interest was one case evaluated with a less conservative gasifier design. Less steam was fed to the gasifier and the overall gasifier performance was significantly improved. However, the altered gas composition required more severe quenching and the overall system did not benefit. In other words, optimization of the gasifier operating conditions did not optimize the overall system. Similarly, cases that gave extreme purification of the fuel cell feed, at a very high fuel cell voltage output, required sufficient parasitic power demand that the overall system did not benefit.

In the more detailed discussions of the individual systems presented in the body of the document and in Appendixes Ca and Da, the relative importance of each element in the system becomes more apparent.

APPENDIX Cc. Summary Printouts for Cases
in Appendix Ca

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CASE I

The Base Case

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TABULATION OF SHAFT ENERGY AVAILABLE CP NEEDED.

102	-55524.6	KILOWATTS	AIR COMPRESSOR FOR SUPPLY TO CRYOGENIC OXYGEN PLANT
104	-24225.4	KILOWATTS	CRYOGENIC OXYGEN COMPRESSOR FOR SUPPLY TO COAL GASIFIER
2	-5737.5	KILOWATTS	2 TEXACO GASIFIER FOR JET PROP. LASS. BASE CASE.
13	-1106.4	KILOWATTS	13 REMOVE SULFUR FROM RAW GAS BY SELEXOL SYSTEM.
17	22454.1	KILOWATTS	17 EXPAND SWEETENED RAW GAS TO FUEL CELL PRESSURE.
22	21431.5	KILOWATTS	22 FLOW STEAM DOWN POWER TURBINE. ADD TO PROCESS GAS (# 17).
3	-14774.5	KILOWATTS	3 COMPRESS COMBUSTION AIR TO FUEL CELL PRESSURE.
10	54845.4	KILOWATTS	10 THE FUEL CELL, 95% CONVERSION, SHIFT ONLY WITHIN ANODE.
15	-933.6	KILOWATTS	15 ANODE EXIT GAS BOOSTER FAN
20	-32577.2	KILOWATTS	20 CATHODE RECYCLE GAS CIRCULATION FAN.
21	245109.6	KILOWATTS	21 EXPAND STACK GAS THROUGH TURBINE TO RECOVER POWER.

1 JPL-1 BASE CASE 35% CONV., NO BURNCH, NO PRE-SHIFT, ISO EFF: COMP 92% T 92%
CASE 1 JPL-1
STEAM LIST
18 MAR 93 9:45:11

TABULATION OF STEAM NEEDED IN PROCESS.

13	65.0 PSIA	4747.	WOLS STEAM	17	REMOVE SULFUR FROM PAW GAS BY SELEXOL SYSTEM.
22	1500.0 PSIA	16351.	WOLS STEAM	22	FLOW STEAM DOWN POWER TURBINE. ADD TO PROCESS GAS (# 17).

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CASE 1 JPL.1
WASTE HEAT OUT
18 MAR 81 9:06:11

1 JPL.1 BASE CASE 85% CONV., NO QUELCH, NO REF-SHIFT, ISO EFF: COMP 80% T 92%

TABULATION OF WASTE HEAT AVAILABLE FROM PROCESS BOILERS.

4	300.0 DEGREES F.	442.64 MILLIONS BTU	4 COOL RAW GAS TO 800 DEGREES FARENHEIT.
7	600.0 DEGREES F.	19.72 MILLIONS BTU	7 WASTE HEAT BOILER PARALLEL TO RAW GAS RECUP. EXCH. TRAIN.
14	250.0 DEGREES F.	146.42 MILLIONS BTU	14 COOL ANODE EXIT GAS IN WH9 - RAISE STEAM.
13	1050.0 DEGREES F.	546.21 MILLIONS BTU	13 COOL CATHODE RECYCLE GAS IN WH8 - RAISE STEAM.

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CASE Id

Conservative Heat Recuperation

TABULATION OF SHAFT ENERGY AVAILABLE OR NEEDED.

- 97 -55524.6 KILOWATTS AIR COMPRESSOR FOR SUPPLY TO CRYOGENIC OXYGEN PLANT
- 99 -24225.4 KILOWATTS CRYOGENIC OXYGEN COMPRESSOR FOR SUPPLY TO COAL GASIFIER
- 2 -5237.6 KILOWATTS 2 TEKACO GASIFIER FOR JET PROP. LABS. BASE CASE.
- 11 -3746.3 KILOWATTS 11 REMOVE SULFUR FROM RAW GAS BY SELEXOL SYSTEM.
- 17 23662.2 KILOWATTS 17 EXPAND SWEETENED RAW GAS TO FUEL CELL PRESSURE.
- 22 21431.6 KILOWATTS 22 BLOW STEAM DOWN POWER TURBINE. ADD TO PROCESS GAS (# 17).
- 3 -147714.3 KILOWATTS 3 COMPRESS COMBUSTION AIR TO FUEL CELL PRESSURE.
- 10 542731.2 KILOWATTS 10 THE FUEL CELL, 85% CONVERSION, SHIFT ONLY WITHIN ANODE.
- 15 -8939.6 KILOWATTS 15 ANODE EXIT GAS BOOSTER FAN
- 20 -39237.8 KILOWATTS 20 CATHODE RECYCLE GAS CIRCULATION FAN.
- 21 251910.7 KILOWATTS 21 EXPAND STACK GAS THROUGH TURBINE TO RECOVER POWER.

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JPL-TX-10 MODIFIED BASE CASE 95% CONV, NO QUENCH, NO PRE-SHIFT, TRUE HEAT RECU. CASE JPL-TX-10
STEAM LIST
5 JAN 83 16:35:31

TABULATION OF STEAM NEEDED IN PROCESS:

11 65.0 PSIA 4714. MOLS STEAM 11 REMOVE SULFUR FROM RAW GAS BY SELEXOL SYSTEM.
22 1500.0 PSIA 16351. MOLS STEAM 22 BLOW STEAM DOWN POWER TURBINE. ADD TO PROCESS GAS (# 17).

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JPL.TX.1D MODIFIED BASE CASE 85% CONV, NO QUENCH, NO PRE-SHIFT, TRUE HEAT RECUP. CASE JPL.TX.1D WASTE HEAT OUT 5 JAN 83 16:35:31

TABULATION OF WASTE HEAT AVAILABLE FROM PROCESS BOILERS.

4	900.0 DEGREES F.	613.07 MILLIONS BTU	4 COOL RAW GAS TO 900 DEGREES FAHRENHEIT, MAKE STEAM.
14	950.0 DEGREES F.	195.43 MILLION BTU	14 COOL ANODE EXIT GAS IN WMB - RAISE STEAM.
18	1050.0 DEGREES F.	669.98 MILLIONS BTU	18 COOL CATHODE RECYCLE GAS IN WMB - RAISE STEAM.

ORIGINAL PAGE 19 OF POOR QUALITY

CASE Ie

Carbon Control and Shift

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TABULATION OF SHAFT ENERGY AVAILABLE OR NEEDED:

97	-55524.6 KILOWATTS	AIR COMPRESSOR FOR SUPPLY TO CRYOGENIC OXYGEN PLANT
99	-24225.4 KILOWATTS	CRYOGENIC OXYGEN COMPRESSOR FOR SUPPLY TO COAL GASIFIER
2	-5737.6 KILOWATTS	2 TEXACO GASIFIER FOR JET PROP. LABS. BASE CASE.
27	-6604.9 KILOWATTS	27 REMOVE SULFUR FROM RAW GAS BY SELEXOL SYSTEM.
36	4616.1 KILOWATTS	36 ADD STEAM VIA TURB; PREVENT CARBON DROP IN FUEL CELL.
38	33209.4 KILOWATTS	38 EXPAND SWEETENED RAW GAS TO FUEL CELL PRESSURE.
3	-147714.8 KILOWATTS	3 COMPRESS COMBUSTION AIR TO FUEL CELL PRESSURE.
10	549074.1 KILOWATTS	10 THE FUEL CELL, 85% CONVERSION, SHIFT ONLY WITHIN ANODE.
15	-9304.7 KILOWATTS	15 ANODE EXIT GAS BOOSTER FAN
20	-22213.2 KILOWATTS	20 CATHODE RECYCLE GAS CIRCULATION FAN.
21	251442.8 KILOWATTS	21 EXPAND STACK GAS THROUGH TURBINE TO RECOVER POWER.

JPL-TX-1E MODIFIED BASE CASE 25% CONV, WITH QUENCH, PRE-SHIFT, 2 COS DESTRUCTION

TABULATION OF STEAM NEEDED IN PROCESS.

27 6568. MOLS STEAM 27 REMOVE SULFUR FROM RAW GAS BY SELEXOL SYSTEM.
36 1500.0 PSIA 6700. MOLS STEAM 36 ADD STEAM VIA TURB PREVENT CARBON DROP IN FUEL CELL.

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JPL-TX-1E MODIFIED BASE CASE 85% CONV, WITH QUENCH, PRE-SHIFT, & COS DESTRUCTION CASE JPL-TX-1E
WASTE HEAT OUT WASTE HEAT OUT
5 JAN 83 16:59:02

TABULATION OF WASTE HEAT AVAILABLE FROM PROCESS BOILERS.

9	900.0 DEGREES F.	193.21 MILLIONS BTU	8 COOL RAW GAS TO 900 DEGREES FAHRENHEIT, MAKE STEAM.
20	600.0 DEGREES F.	169.55 MILLIONS BTU	20 COOL SWIFTED EXIT GAS; RAISE STEAM.
14	850.C DEGREES F.	192.92 MILLIONS BTU	14 COOL ANODE EXIT GAS IN WHB - RAISE STEAM.
18	1050.0 DEGREES F.	356.04 MILLIONS BTU	18 COOL CATHODE RECYCLE GAS IN WHB - RAISE STEAM.

CASE I f
Optimized by Moving Expander

JPL.TX.IF MODIFIED BASE CASE 85% CONV. QUENCH, PRE-SHIFT, COSA, TURBO-EXP MOVED.

TABULATION OF SHAFT ENERGY AVAILABLE ON NEEDED.

97	-55542.6	KILOWATTS	AIR COMPRESSOR FOR SUPPLY TO CRYOGENIC OXYGEN PLANT
99	-24233.3	KILOWATTS	CRYOGENIC OXYGEN COMPRESSOR FOR SUPPLY TO COAL GASIFIER
2	-5739.4	KILOWATTS	2 TEXACO GASIFIER FWH JET PROP. LABS. BASE CASE.
19	-43603.7	KILOWATTS	19 EXPAND SHIFT (1) EXIT GAS THROUGH POWER RECOVERY TURBINE.
30	-7225.4	KILOWATTS	30 REMOVE SULFUR FROM RAW GAS BY SELEXOL SYSTEM.
42	6468.0	KILOWATTS	42 ADD STEAM VIA TURBINE; PREVENT CARBON DROP IN FUEL CELL.
44	-147425.5	KILOWATTS	44 COMPRESS COMBUSTION AIR TO FUEL CELL PRESSURE.
51	553267.7	KILOWATTS	51 THE FUEL CELL; 85% CONVERSION; SHIFT ONLY WITHIN ANODE.
57	-8412.1	KILOWATTS	57 ANODE EXIT GAS BOOSTER FAN, TO FEED CATALYTIC BURNER.
63	-15251.3	KILOWATTS	63 CATHODE RECYCLE GAS CIRCULATION FAN.
64	252302.9	KILOWATTS	64 EXPAND STACK GAS THROUGH TURBINE TO RECOVER POWER.

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JPL.TX.IF MODIFIED BASE CASE 5% CONV, QUENCH, PRE-SHIFT, COSX, TURBU-EXP MOVED.
CASE JPL.TX.IF
STEAM LIST
5 JAN 83 16157151

TABULATION OF STEAM NEEDED IN PROCESS.

30	65.0 PSIA	12168. MOLS STEAM	30 REMOVE SULFUR FROM RAW GAS BY SELEXOL SYSTEM.
42	1500.0 PSIA	4859. MOLS STEAM	42 ADD STEAM VIA TURBINE; PREVENT CARBON DROP IN FUEL CELL.

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JPL TX IF
WASTE HEAT OUT
5 JAN 83 16157151

JPL TX IF MODIFIED BASE CASE 85% CONV, BLENCH, PRE-SHIFT, COSA, TUMBU-EXP MOVED.

TABULATION OF WASTE HEAT AVAILABLE FROM PROCESS BUILENS.

8	700.0 DEGREES F.	264.96 MILLIONS BTU	8 COOL RAN GAS TO 700 DEGREES FAHRENHEIT, MAKE STEAM.
20	600.0 DEGREES F.	49.27 MILLIONS BTU	20 COOL INTER-STAGE SHIFT GAS IN WHB1 RAISE STEAM
54	420.0 DEGREES F.	324.07 MILLIONS BTU	54 COOL ANODE EXIT GAS IN WHB - RAISE STEAM.
60	1050.0 DEGREES F.	274.00 MILLIONS BTU	60 COOL CATHODE RECYCLE GAS IN WHB - RAISE STEAM.

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CASE II

Fluidized-Bed Gasifier

TABULATION OF SHIFT ENERGY AVAILABLE OR NEEDED:

97 -4295.3 KILOWATTS AIR COMPRESSOR FOR SUPPLY TO CRYOGENIC OXYGEN PLANT
99 -1725.1 KILOWATTS CRYOGENIC OXYGEN COMPRESSOR FOR SUPPLY TO COAL GASIFIER
2 -5240.3 KILOWATTS U-GAS GASIFIER NORMAL OPERATION, STEAM #2 = NET MAKE.
20 32755.2 KILOWATTS 20 EXPAND SHIFT (I) ENY GAS THROUGH POWER RECOVERY TURBINE.
24 -7447.1 KILOWATTS 20 REMOVE SULFUR FROM RAW GAS BY SELEXOL SYSTEM.
42 4064.9 KILOWATTS 42 ADD STEAM VIA TURBINE; PREVENT CARBON DROP IN FUEL CELL.
44 -17187.4 KILOWATTS 44 COMPRESS COMBUSTION AIR TO FUEL CELL PRESSURE.
51 50909.1 KILOWATTS 51 THE FUEL CELL, 85% CONVERSION, SHIFT ONLY WITHIN ANODE.
57 -7025.2 KILOWATTS 57 ANODE EXIT GAS BOOSTER FAN TO FEED CATALYTIC BURNER.
63 -52324.3 KILOWATTS 63 CATHODE RECYCLE GAS CIRCULATION FAN.
64 285910.2 KILOWATTS 64 EXPAND STACK GAS THROUGH TURBINE TO RECOVER POWER.
70 -1932.3 KILOWATTS 70 COMPRESS RECYCLE GAS TO INJECT FEED COAL TO GASIFIER.

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JPL 00.2 FLUID BED FACE CASE F5V CONV, QUENCH, PRE-SHIRT, COSK, 8.33X CM4 PAR
CASE JPL 00.2
STEAM LIST
4 JAN 83 13:36:00

TABULATION OF STEAM NEEDED BY PROCESS:

	24072. MOLS STEAM	2 U-GAS GASIFIED	NORMAL OPERATION.	STREAM #2 = NET MAKE.
2	500.0 PSIA			
37	650.0 PSIA			
42	1500.0 PSIA			
		1352. MOLS STEAM	29 REMOVE SULFUR FROM RAW GAS BY SELEXOL SYSTEM.	
		2811. MOLS STEAM	42 ADD STEAM VIA TURBINE; PREVENT CARBON DROP IN FUEL CELL.	

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CASE JPL-US-2
WASTE HEAT OUT
4 JAN 92 13:35:00

JPL-US-2 FLUID TOU RATE CASE 524 CONV, QUENCH, PRE-SHIFT, CCSX, 2.13X C14 RAM

.....
INSULATION OF WASTE HEAT AVAILABLE FOR PROCESS COILS.

0	700.0 DEGREES F.	45.21 MILLIONS BTU	2 COIL RAN GAS TO 700 DEGREE FAMPENHEIT, MAKE STEAM.
21	500.0 DEGREES F.	3.22 MILLIONS BTU	21 COOL INTER-STAGE SHIFT GAS IN WH9, RAISE STEAM
54	920.0 DEGREES F.	33.71 MILLIONS BTU	54 COOL ANODE EXIT GAS IN WH9 - RAISE STEAM.
60	1000.0 DEGREES F.	700.34 MILLIONS BTU	60 COOL CATHODE RECYCLE GAS IN WH5 - RAISE STEAM.

CASE III

Internally Reforming Fuel Cell

JPLUGS FLUID BED, REFORM, 55% CONVA, QUENCH, PRE-SHIFT, CCRK, 2.11% CH4 RAW
CASE JPLUGS
POWER LIST
7 JAN 92 14:04:31

TABULATION OF SHAFI ENERGY AVAILABLE OR NEEDED.

97	-42995.3	KILOWATTS	AIR COMPRESSOR FOR SUPPLY TO CRYOGENIC OXYGEN PLANT
99	-17725.1	KILOWATTS	CRYOGENIC OXYGEN COMPRESSOR FOR SUPPLY TO COAL GASIFIER
2	-5240.3	KILOWATTS	2 U-GAS GASIFIER NORMAL OPERATION. STREAM 42 = NET MAKE.
20	38751.4	KILOWATTS	20 EXPAND SHIFTER (1) EXIT GAS THROUGH POWER RECOVERY TURBINE.
24	-7487.1	KILOWATTS	20 REMOVE SULFUR FROM RAW GAS BY SELENOL SYSTEM.
42	9064.0	KILOWATTS	42 AEO STEAM VIA TURBINE; PREVENT CARBON DROP IN FUEL CELL.
44	-1702-2.1	KILOWATTS	44 COMPRESS COMBUSTION AIR TO FUEL CELL PRESSURE.
51	64795.3	KILOWATTS	51 FMC FUEL CELL, MAXIMUM CONVERSION, REFORM WITHIN ANODE.
57	-8162.4	KILOWATTS	57 ANODE EXIT GAS BOOSTER FAN, TO FEED CATALYTIC BURNER.
63	-19451.2	KILOWATTS	63 CATHODE RECYCLE GAS CIRCULATION FAN.
64	210255.1	KILOWATTS	64 EXPAND STACK GAS THROUGH TURBINE TO RECOVER POWER.
70	-1921.5	KILOWATTS	70 COMPRESS RECYCLE GAS TO INJECT FEED COAL TO GASIFIER.

JPL.UG.3
CASE
STEAM LIST
6 JAN 83 14:04:33

JPL.UG.3 FLUID BED, REFORM, 35% CONV, SUENCH2 PRE-SHIFT, COCV, 8.23X CH4 92W

TABULATION OF STEAM NEEDED IN PROCESS

2	500.0 PSIA	24272. MOLS STEAM	3 U-GAS GASIFIER NORMAL OPERATION, STEAM P2 = NET MAKE.
29	55.0 PSIA	12052. MOLS STEAM	29 REMOVE SULFUR FROM PAW GAS BY SELEMOL SYSTEM.
42	1500.0 PSIA	4811. MOLS STEAM	42 ADD STEAM VIA TURBINE; PREVENT CREEP DROP IN FUEL CELL.

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TABULATION OF WASTE HEAT AVAILABLE FROM PROCESS BOILERS.

9	700.0 DEGREES F.	490.21 MILLIONS BTU	9 COOL RAW GAS TO 700 DEGREES FARENHEIT, MAKE STEAM.
24	600.0 DEGREES F.	1773 MILLIONS BTU	24 COOL INTER-STAGE SHIFT GAS IN WH5, RAISE STEAM.
54	525.0 DEGREES F.	327.70 MILLIONS BTU	54 COOL ANODE EKIT GAS IN WH3 - RAISE STEAM.
60	1050.0 DEGREES F.	213.92 MILLIONS BTU	60 COOL CATHODE RECYCLE GAS IN WH6 - RAISE STEAM.

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CASE IV

High-Methane Gasifier

JPL-UG-4 FLUID DEL. REFORMS 15% CONV. QUENCH. PRE-SHIFT, COSA, 13:50 CH4 RAW
 CASE POWER LIST
 4 JAN 83 14:5:19

 TABULATION OF SHAFI ENERGY AVAILABLE OR NEEDED.

37	-39044.7 KILOWATTS	AIR COMPRESSOR FAN SUPPLY TO CRYOGENIC OXYGEN PLANT
39	-16426.3 KILOWATTS	CRYOGENIC OXYGEN COMPRESSOR FAN SUPPLY TO COAL GASIFIER
4	-5640.3 KILOWATTS	2 U-GAS GASIFIER NORMAL OPERATION. STREAM #2 = NET MAKE.
20	31756.2 KILOWATTS	20 EXHAUST SHIFTS (17) EXIT GAS THROUGH POWER RECOVERY TURBINE.
24	-8057.1 KILOWATTS	24 REMOVE SULFUR FROM RAW GAS BY SLEAZOL SYSTEM.
42	10667.5 KILOWATTS	42 ADD STEAM VIA TURBINE TO PREVENT CARBON DROOP IN FUEL CELL.
44	-170942.3 KILOWATTS	44 COMPRESS COMBUSTION AIR TO FUEL CELL PRESSURE.
51	603892.0 KILOWATTS	51 FUE FUEL CELL. MAXIMUM CONVERSION. REFORM WITHIN ANODE.
57	-8324.6 KILOWATTS	57 ANODE EXIT GAS HOOPER FANS TO FUEL CATALYTIC BURNER.
63	-19496.7 KILOWATTS	63 CATHODE RECYCLE GAS CIRCULATION FAN.
64	270032.7 KILOWATTS	64 EXHAUST STACK GAS THROUGH TURBINE TO RECOVER POWER.
70	-1797.3 KILOWATTS	70 COMPRESS RECYCLE GAS TO INJECT FEED COAL TO GASIFIER.

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JPL 0004 FLETO 0005 REF 0006 P30 TUNV 0007 BUENGRS PRE-SHIP 1. COSA, 13.56 CH4 RAW
CASE JPL 0004
STEAM 11ST
4 JAN 83 14135:19

REGULATION OF STEAM SUPPLY IN FUELS

2 20000 PSIG 24072. MOLS STEAM 2 U-GAS WASTIF IEM NORMAL OPERATION. STEAM #2 = WFI MAKE.
29 6520 PSIG 11789. MOLS STEAM 29 REMOVE SULFUR FROM RAW GAS BY SFLXUL SYSTEM.
42 150000 PSIG 80119. MOLS STEAM 42 ADD STEAM VIA TURBINE 1 PREVENT CARBON DROP IN FUEL CELL.

CASE: JPL-06-4
WASTE HEAT OUT
9 JAN 83 14:35:14

JPL-06-4 PLD10 BEG, RECURSIVE HX CONV, GULMUM, PRL-SMFT, CUSA, 13.5% CH4 HAW

TABLETION OF WASTE HEAT AVAILABLE FROM PROCESS UTILERS.

9	700.0 DEGREES F.	420.00 MILLIONS BTU	9 COOL HAW GAS TO 700 DEGREES F	PARMENHELI, WAST STEAM.
21	600.0 DEGREES F.	10.00 MILLIONS BTU	21 COOL INFR-STAGE SHFT GAS IN WMB	RAISE STEAM.
54	925.0 DEGREES F.	319.00 MILLIONS BTU	54 COOL ANODE CATH GAS IN WMB	RAISE STEAM.
60	1090.0 DEGREES F.	199.02 MILLIONS BTU	60 COOL CATHODE RECYCLE GAS IN WMB	RAISE STEAM.

CASE IVa

Conceptual Gasifier

INSULATION OF SHAFT ENERGY AVAILABLE IS NEEDED.

97	-11753.6 KILOWATTS	AIR COMPRESSOR FOR SUPPLY TO CRYOGENIC OXYGEN PLANT
98	-13730.1 KILOWATTS	CRYOGENIC OXYGEN COMPRESSOR FOR SUPPLY TO COAL GASIFIER
2	-5290.7 KILOWATTS	U-245 GASIFIER NORMAL OPERATION, STREAM #2 = NET MAKE.
20	30747.8 KILOWATTS	20 EXPAND SPIST (1) EXIT GAS THROUGH POWER RECOVERY TURBINE.
21	-6471.0 KILOWATTS	21 REMOVE SULFUR FROM RAW GAS BY SELKAL SYSTEM.
42	12333.7 KILOWATTS	42 AND STEAM VAL TURBINE; PREVENT CARBON DUMP IN FUEL CELL.
44	-17072.3 KILOWATTS	44 COMPRESS COMBUSTION AIR TO FUEL CELL PRESSURE.
51	71233.2 KILOWATTS	51 THE FUEL CELL, MAXIMUM CONVERSION, PERFORM WITHIN ANODE.
57	-5283.4 KILOWATTS	57 ANODE EXIT GAS BOOSTER FAN, TO FEED CATALYTIC BURNER.
58	23167.7 KILOWATTS	58 CATHODE RECYCLE GAS CIRCULATION FAN.
64	23167.1 KILOWATTS	64 EXPAND STACK GAS THROUGH TURBINE TO RECOVER POWER.
70	-12993.3 KILOWATTS	70 COMPRESS RECYCLE GAS TO INJECT FEED COAL TO GASIFIER.

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CASE JPLUGS44

STEAM LIST

5 JUN 83 085452

JPLUGS44 FLUIDS, REFORMER, TRX COLTS, BLINCH, REFORMER, COSY, ROCK CH4 RAW

TABULATION OF STEAM NEEDED IN PROCESS.

2	500.0 PSIA	24072. MOLES STEAM	2 U-345 GASIFIED NORMAL OPERATION, STREAM #2 = NET MAKE.
27	65.0 PSIA	11278. MOLES STEAM	29 SENSITIVE SULFUR FROM RAW GAS BY SELENOL SYSTEM.
42	1500.0 PSIA	32455. MOLES STEAM	42 ADD STEAM VIA TURBINE; PREVENT CARBON DROP IN FUEL CELL.

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CASE JPLUG.4A
WASTE HEAT OUT
5 JAN 83 3:55:52

JPLUG.4A FLUID BED, REFIN, 2ND CONV, BURNCH, 2ND-SHIFT, COSM, 20.02 CH4 RAW

TABULATION OF WASTE HEAT AVAILABLE FROM PROCESS BOILERS.

	7	21	54	80
700.0 DEGREES F.	247.42 MILLIONS BTU	0 COOL RAW GAS TO 700 DEGREES FARENHEIT, MAKE STEAM.		
600.0 DEGREES F.	0.74 MILLIONS BTU	21 COOL INTER-STAGE SHIFT GAS IN WHR? RAISE STEAM.		
525.0 DEGREES F.	350.15 MILLIONS BTU	54 COOL ANODE EXIT GAS IN WHR - RAISE STEAM.		
475.0 DEGREES F.	185.0 MILLIONS BTU	80 COOL CATHODE RECYCLE GAS IN WHR - RAISE STEAM.		

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CASES IVb, c, and d

Vary Fuel Cell Conversion

CASE CONV70
WASTE HEAT OUT
5 JAN 83 11:34:14

CONV70 FLUID BED, REFORM, 85% CONV, QUENCH, PRE-SHIFT, COSN, 13.5X CH4 RAW

TABULATION OF WASTE HEAT AVAILABLE FROM PROCESS BOILERS.

9	700.0 DEGREES F.	458.66 MILLIONS BTU	9 COOL RAW GAS TO 700 DEGREES FAHRENHEIT, MAKE STEAM.
21	600.0 DEGREES F.	10.06 MILLIONS BTU	21 COOL INTER-STAGE SHIFT GAS IN WHB; RAISE STEAM.
54	925.0 DEGREES F.	434.58 MILLIONS BTU	54 COOL ANODE EXIT GAS IN WHB - RAISE STEAM.
60	1115.0 DEGREES F.	372.08 MILLIONS BTU	60 COOL CATHODE RECYCLE GAS IN WHB - RAISE STEAM.

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CONV70 FLUID BED, REFORM, BSX CONV, QUENCH, PRE-SHIFT, COSX, 13.5X CH4 RAW
CASE CONV70
STEAM LIST
5 JAN 83 11:36:14

TABULATION OF STEAM NEEDED IN PROCESS.

2	500.0 PSIA	24072. MOLS STEAM	2 U-GAS GASIFIER NORMAL OPERATION. STREAM #2 = NET MAKE.
29	65.0 PSIA	11889. MOLS STEAM	29 REMOVE SULFUR FROM RAW GAS BY SELEXOL SYSTEM.
42	1500.0 PSIA	8015. MOLS STEAM	42 ADD STEAM VIA TURBINE; PREVENT CARBON DROP IN FUEL CELL.

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TABULATION OF SHAFT ENERGY AVAILABLE OR NEEDED.

97	-39844.7	KILOWATTS	AIR COMPRESSOR FOR SUPPLY TO CRYOGENIC OXYGEN PLANT
99	-16426.3	KILOWATTS	CRYOGENIC OXYGEN COMPRESSOR FOR SUPPLY TO COAL GASIFIER
2	-5240.3	KILOWATTS	2 U-GAS GASIFIER NORMAL OPERATION. STREAM #2 = NET MAKE.
20	31756.2	KILOWATTS	20 EXPAND SHIFT (I) EXIT GAS THROUGH POWER RECOVERY TURBINE.
29	-6958.0	KILOWATTS	29 REMOVE SULFUR FROM RAW GAS BY SELEXOL SYSTEM.
42	10667.7	KILOWATTS	42 ADD STEAM VIA TURBINE; PREVENT CARBON DROP IN FUEL CELL.
44	-170942.3	KILOWATTS	44 COMPRESS COMBUSTION AIR TO FUEL CELL PRESSURE.
51	533732.9	KILOWATTS	51 THE FUEL CELL, MAXIMUM CONVERSION, REFORM WITHIN ANODE.
57	-7765.7	KILOWATTS	57 ANODE EXIT GAS BOOSTER FAN, TO FEED CATALYTIC BURNER.
63	-21907.8	KILOWATTS	63 CATHODE RECYCLE GAS CIRCULATION FAN.
64	303549.5	KILOWATTS	64 EXPAND STACK GAS THROUGH TURBINE TO RECOVER POWER.
70	-1797.3	KILOWATTS	70 COMPRESS RECYCLE GAS TO INJECT FEED COAL TO GASIFIER.

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CONV80 FLUID BED, REFORM, 85% CONV, QUENCH, PRE-SHIFT, COSX, 13.5% CH4 RAW

CASE CONV80
STEAM LIST

5 JAN 83 11:40:08

TABULATION OF STEAM NEEDED IN PROCESS.

2	500.0 PSIA	24072.	MOLS STEAM	2	U-GAS GASIFIER NORMAL OPERATION. STREAM #2 = NET MAKE.
29	65.0 PSIA	11889.	MOLS STEAM	29	REMOVE SULFUR FROM RAW GAS BY SELEXOL SYSTEM.
42	1500.0 PSIA	8015.	MOLS STEAM	42	ADD STEAM VIA TURBINE; PREVENT CARBON DROP IN FUEL CELL.

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CONV80 FLUID BED, REFORM, 85% CONV, QUENCH, PRE-SHIFT, COSX, 13.5X CH4 9AW CASE CONV80
WASTE HEAT OUT
5 JAN 83 11:40:08

TABULATION OF WASTE HEAT AVAILABLE FROM PROCESS BOILERS.

9	700.0 DEGREES F.	458.66 MILLIONS BTU	9 COOL RAW GAS TO 700 DEGREES FAHRENHEIT, MAKE STEAM.
21	600.0 DEGREES F.	10.06 MILLIONS BTU	21 COOL INTER-STAGE SHIFT GAS IN WHB; RAISE STEAM.
54	925.0 DEGREES F.	364.68 MILLIONS BTU	54 COOL ANODE EXIT GAS IN WHB - RAISE STEAM.
60	1115.0 DEGREES F.	240.53 MILLIONS BTU	60 COOL CATHODE RECYCLE GAS IN WHB - RAISE STEAM.

TABULATION OF SHAFT ENERGY AVAILABLE OR NEEDED.

97	-39844.7	KILOWATTS	AIR COMPRESSOR FOR SUPPLY TO CRYOGENIC OXYGEN PLANT
99	-16426.3	KILOWATTS	CRYOGENIC OXYGEN COMPRESSOR FOR SUPPLY TO COAL GASIFIER
2	-5240.3	KILOWATTS	2 U-GAS GASIFIER NORMAL OPERATION, STREAM #2 = NET MAKE.
20	31756.2	KILOWATTS	20 EXPAND SHIFT (1) EXIT GAS THROUGH POWER RECOVERY TURBINE.
29	-6958.0	KILOWATTS	29 REMOVE SULFUR FROM RAW GAS BY SELENOL SYSTEM.
42	10667.7	KILOWATTS	42 ADD STEAM VIA TURBINE; PREVENT CARBON DROP IN FUEL CELL.
44	-170942.3	KILOWATTS	44 COMPRESS COMBUSTION AIR TO FUEL CELL PRESSURE.
51	628019.5	KILOWATTS	51 THE FUEL CELL, MAXIMUM CONVERSION, REFORM WITHIN ANODE.
57	-8145.0	KILOWATTS	57 ANODE EXIT GAS BOOSTER FAN, TO FEED CATALYTIC BURNER.
63	-20514.4	KILOWATTS	63 CATHODE RECYCLE GAS CIRCULATION FAN.
64	287622.6	KILOWATTS	64 EXPAND STACK GAS THROUGH TURBINE TO RECOVER POWER.
70	-1797.3	KILOWATTS	70 COMPRESS RECYCLE GAS TO INJECT FEED COAL TO GASIFIER.

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CONV90 FLUID BED, REFORM, 85% CONV, QUENCH, PRE-SHIFT, COSX, 13.5% CH4 RAW

TABULATION OF SHAFT ENERGY AVAILABLE OR NEEDED.

97	-39844.7	KILOWATTS	AIR COMPRESSOR FOR SUPPLY TO CRYOGENIC OXYGEN PLANT
99	-16426.3	KILOWATTS	CRYOGENIC OXYGEN COMPRESSOR FOR SUPPLY TO COAL GASIFIER
2	-5240.3	KILOWATTS	2 U-GAS GASIFIER NORMAL OPERATION. STREAM #2 = NET MAKE.
20	31756.2	KILOWATTS	20 EXPAND SHIFT (1) EXIT GAS THROUGH POWER RECOVERY TURBINE.
29	-6958.0	KILOWATTS	29 REMOVE SULFUR FROM RAW GAS BY SELEXOL SYSTEM.
42	10667.7	KILOWATTS	42 ADD STEAM VIA TURBINE; PREVENT CARBON DROP IN FUEL CELL.
44	-170942.3	KILOWATTS	44 COMPRESS COMBUSTION AIR TO FUEL CELL PRESSURE.
51	698380.3	KILOWATTS	51 THE FUEL CELL, MAXIMUM CONVERSION, REFORM WITHIN ANODE.
57	-8501.4	KILOWATTS	57 ANODE EXIT GAS BOOSTER FAN, TO FEED CATALYTIC BURNER.
63	-19240.8	KILOWATTS	63 CATHODE RECYCLE GAS CIRCULATION FAN.
64	272989.1	KILOWATTS	64 EXPAND STACK GAS THROUGH TURBINE TO RECOVER POWER.
70	-17977.3	KILOWATTS	70 COMPRESS RECYCLE GAS TO INJECT FEED COAL TO GASIFIER.

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CONV90 FLUID BED, REFORM, 95X CONV, QUENCH, PRE-SHIFT, COSK, 13.5X CH4 RAW
CASE CONV90
STEAM LIST
5 JAN 83 11:44:57

TABULATION OF STEAM NEEDED IN PROCESS.

2	500.0 PSIA	24072. MOLS STEAM	2 U-GAS GASIFIER NORMAL OPERATION. STREAM #2 = NET MAKE.
29	65.0 PSIA	11889. MOLS STEAM	29 REMOVE SULFUR FROM RAW GAS BY SELEXOL SYSTEM.
42	1500.0 PSIA	8015. MOLS STEAM	42 ADD STEAM VIA TURBINE; PREVENT CARBON DROP IN FUEL CELL.

CONV90 FLUID BED, REFORM, 85% CONV, QUENCH, PRE-SHIFT, COSX, 13.5X CH4 RAW CASE CONV90
WASTE HEAT OUT WASTE HEAT OUT
5 JAN 83 11:44:57

TABULATION OF WASTE HEAT AVAILABLE FROM PROCESS BOILERS.

9	700.0 DEGREES F.	458.66 MILLIONS BTU	9 COOL RAW GAS TO 700 DEGREES FAHRENHEIT, MAKE STEAM.
21	600.0 DEGREES F.	10.06 MILLIONS BTU	21 COOL INTER-STAGE SHIFT GAS IN WHB; RAISE STEAM.
54	925.0 DEGREES F.	292.05 MILLIONS BTU	54 COOL ANODE EXIT GAS IN WH9 - RAISE STEAM.
60	1115.0 DEGREES F.	120.91 MILLIONS BTU	60 COOL CATHODE RECYCLE GAS IN WHB - RAISE STEAM.

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CASE IVe

Optimized Fluidized-Bed System

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TABULATION OF SHAFT ENERGY AVAILABLE OR NEEDED.

97	-39844.7	KILOWATTS	AIR COMPRESSOR FOR SUPPLY TO CRYOGENIC OXYGEN PLANT
99	-16426.3	KILOWATTS	CRYOGENIC OXYGEN COMPRESSOR FOR SUPPLY TO COAL GASIFIER
2	-5240.3	KILOWATTS	2 U-GAS GASIFIER NORMAL OPERATION, STREAM #2 = NET MAKE.
20	31756.2	KILOWATTS	20 EXPAND SHIFT (1) EXIT GAS THROUGH POWER RECOVERY TURBINE.
29	-6958.0	KILOWATTS	29 REMOVE SULFUR FROM RAW GAS BY SELEXOL SYSTEM.
42	10667.7	KILOWATTS	42 ADD STEAM VIA TURBINE1 PREVENT CARBON DROPP IN FUEL CELL.
44	-174015.4	KILOWATTS	44 COMPRESS COMBUSTION AIR TO FUEL CELL PRESSURE.
51	664628.2	KILOWATTS	51 THE FUEL CELL, MAXIMUM CONVERSION, REFORM WITHIN ANODE.
57	-8324.9	KILOWATTS	57 ANODE EXIT GAS HOUSTER FAN, TO FEED CATALYTIC BURNER.
63	-15467.8	KILOWATTS	63 CATHODE RECYCLE GAS CIRCULATION FAN.
64	285486.4	KILOWATTS	64 EXPAND STACK GAS THROUGH TURBINE TO RECOVER POWER.
70	-1797.3	KILOWATTS	70 COMPRESS RECYCLE GAS TO INJECT FEED COAL TO GASIFIER.

TABULATION OF STEAM NEEDED IN PROCESS.

2	500.0 PSIA	24072. MOLS STEAM	2 U-GAS GASIFIER NORMAL OPERATION, STREAM #2 = NET MAKE.
29	65.0 PSIA	11889. MOLS STEAM	29 REMOVE SULFUR FROM RAW GAS BY SELEXOL SYSTEM.
42	1500.0 PSIA	8015. MOLS STEAM	42 ADD STEAM VIA TURBINE TO PREVENT CARBON DROP IN FUEL CELL.

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AIR4.160 JPL4, 2-STAGE AIR COMPRESSOR, WITH AIR FEED @ 160.0% OF STOICHIOMETRIC
CASE AIR4.160
WASTE HEAT OUT
14 JAN 83 9:21:14

TABULATION OF WASTE HEAT AVAILABLE FROM PROCESS BUILDERS.

9	700.0	DEGREES F.	458.66	MILLIONS BTU	9	COOL RAW GAS TO 700 DEGREES FAHRENHEIT, MAKE STEAM.
21	600.0	DEGREES F.	10.06	MILLIONS BTU	21	COOL INTER-STAGE SHIFT GAS IN WHB1 RAISE STEAM.
54	425.0	DEGREES F.	338.74	MILLIONS BTU	54	COOL ANODE EXIT GAS IN WHB - RAISE STEAM.
60	1115.0	DEGREES F.	157.28	MILLIONS BTU	60	COOL CATHODE RECYCLE GAS IN WHB - RAISE STEAM.

CASE IVF

Minimum Humidification

TABLATION OF SHAFT ENERGY AVAILABLE OR NEEDED.

97	-39844.7 KILOWATTS	AIR COMPRESSOR FOR SUPPLY TO CRYOGENIC OXYGEN PLANT
99	-16426.3 KILOWATTS	CRYOGENIC OXYGEN COMPRESSOR FOR SUPPLY TO COAL GASIFIER
2	-5240.3 KILOWATTS	2 U-GAS GASIFIER NORMAL OPERATION, STREAM #2 .5 NET MAKE.
20	30639.4 KILOWATTS	20 EXPAND SHIFT (I) EXIT GAS THROUGH POWER RECOVERY TURBINE.
29	-6488.6 KILOWATTS	29 REMOVE SULFUR FROM RAW GAS BY SELEXOL SYSTEM.
42	8907.3 KILOWATTS	42 ADD STEAM VIA TURBINE PREVENT CARBON DROP IN FUEL CELL.
44	-171083.5 KILOWATTS	44 COMPRESS COMBUSTION AIR TO FUEL CELL PRESSURE.
51	671199.0 KILOWATTS	51 THE FUEL CELL, MAXIMUM CONVERSION, REFORM WITHIN ANODE.
57	-8174.4 KILOWATTS	57 ANODE EXIT GAS BOOSTER FAN, TO FEED CATALYTIC BURNER.
63	-20080.5 KILOWATTS	63 CATHODE RECYCLE GAS CIRCULATION FAN.
64	278927.9 KILOWATTS	64 EXPAND STACK GAS THROUGH TURBINE TO RECOVER POWER.
70	-1784.0 KILOWATTS	70 COMPRESS RECYCLE GAS TO INJECT FEED COAL TO GASIFIER.

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JPL.2S.4 FLUID BED, REFORM, 85% CONV, QUENCH, PRE-SHIFT, COSX, 13.5% CH4 RAW
CASE JPL.2S.4
STEAM LIST
28 JAN 83 14143100

TABULATION OF STEAM NEEDED IN PROCESS,

2	550.0 PSIA	24072.	MULS STEAM	2	U-GAS GASIFIER NORMAL OPERATION.	STREAM #2 = NET MAKE.
29	65.0 PSIA	11804.	MULS STEAM	29	REMOVE SULFUR FROM RAW GAS BY SELEXOL SYSTEM.	
42	1500.0 PSIA	6692.	MULS STEAM	42	ADD STEAM VIA TURBINE TO PREVENT CARBON DROP IN FUEL CELL.	

JPL.2S.4 JPL.2S.4
WASTE HEAT OUT
28 JAN 83 14143100

JPL.2S.4 FLUID BED, REFORM, 85% CONV, QUENCH, PRE-SHIFT, COSX, 13.5% CM4 RAM

LABULATION OF WASTE HEAT AVAILABLE FROM PROCESS BUILDERS.

9	700.0 DEGREES F.	516.63 MILLIONS BTU	9 COOL RAW GAS TO 700 DEGREES FAHRENHEIT, MAKE STEAM.
21	600.0 DEGREES F.	5.66 MILLIONS BTU	21 COOL INTER-STAGE SHIFT GAS IN WHB1 RAISE STEAM.
54	925.0 DEGREES F.	331.55 MILLIONS BTU	54 COOL ANODE EXIT GAS IN WHB - RAISE STEAM.
60	1115.0 DEGREES F.	188.01 MILLIONS BTU	60 COOL CATHODE RECYCLE GAS IN WHB - RAISE STEAM.

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CASE IVg

No Quench

TABULATION OF SHAFT ENERGY AVAILABLE OR NEEDED.

97 -39844.7 KILOWATTS AIR COMPRESSOR FOR SUPPLY TO CRYOGENIC OXYGEN PLANT
99 -16426.3 KILOWATTS CRYOGENIC OXYGEN COMPRESSOR FOR SUPPLY TO COAL GASIFIER
2 -5240.3 KILOWATTS 2 U-GAS GASIFIER NORMAL OPERATION. STREAM #2 = NET MAKE.
20 28573.4 KILOWATTS 20 EXPAND SHIFT (1) EXIT GAS THROUGH POWER RECOVERY TURBINE.
29 -6630.1 KILOWATTS 29 REMOVE SULFUR FROM RAW GAS BY SELEXOL SYSTEM.
42 7583.8 KILOWATTS 42 ADD STEAM VIA TURBINE; PREVENT CARBON DROP IN FUEL CELL.
44 -171083.5 KILOWATTS 44 COMPRESS COMBUSTION AIR TO FUEL CELL PRESSURE.
51 669336.2 KILOWATTS 51 THE FUEL CELL, MAXIMUM CONVERSION, REFORM WITHIN ANODE.
57 -8014.1 KILOWATTS 57 ANODE EXIT GAS BOOSTER FAN, TO FEED CATALYTIC BURNER.
63 -19956.5 KILOWATTS 63 CATHODE RECYCLE GAS CIRCULATION FAN.
64 278054.1 KILOWATTS 64 EXPAND STACK GAS THROUGH TURBINE TO RECOVER POWER.
70 -1753.3 KILOWATTS 70 COMPRESS RECYCLE GAS TO INJECT FEED COAL TO GASIFIER.

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CASE DKF.4
STEAM LISI
12 JAN 83 15:18:23

DKF.4 COMPETITIVE, REFORM, 85% CONV, NO QUENCH, PRE-SHIFT, COSX, 13.5% CH4

TABULATION OF STEAM NEEDED IN PROCESS.

2	500.0 PSIA	24072.	MOLS STEAM	2	U-GAS GASIFIER NORMAL OPERATION. STREAM #2 = NET MAKE.
29	65.0 PSIA	11608.	MOLS STEAM	29	REMOVE SULFUR FROM RAW GAS BY SELEXOL SYSTEM.
42	1500.0 PSIA	5698.	MOLS STEAM	42	ADD STEAM VIA TURBINE; PREVENT CARBON DROP IN FUEL CELL.

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DKF.4 CASE DKF.4
COMPETITIVE, REFORM, 85% CONV, NO QUENCH, PRE-SHIFT, COSX, 13.5% CH4 WASTE HEAT OUT
12 JAN 83 15:18:23

TABULATION OF WASTE HEAT AVAILABLE FROM PROCESS BOILERS.

8	1507.8 DEGREES F.	207.82 MILLIONS BTU	8 QUENCH STEP OMITTED; GAS WILL PLUG WASTE HEAT BOILER.
9	700.0 DEGREES F.	422.50 MILLIONS BTU	9 COOL RAW GAS TO 700 DEGREES FAHRENHEIT, MAKE STEAM.
21	600.0 DEGREES F.	-3.97 MILLIONS BTU	21 COOL INTER-STAGE SHIFT GAS IN WH9; RAISE STEAM.
54	925.0 DEGREES F.	337.21 MILLIONS BTU	54 COOL ANODE EXIT GAS IN WMB - RAISE STEAM.
60	1115.0 DEGREES F.	200.80 MILLIONS BTU	60 COOL CATHODE RECYCLE GAS IN WMB - RAISE STEAM.

CASE V

Dust-Tolerant Shift Reactor

TABULATION OF SHAFT ENERGY AVAILABLE OR NEEDED.

97	-3984.7	KILOWATTS	AIR COMPRESSOR FOR SUPPLY TO CRYOGENIC OXYGEN PLANT
99	-16426.3	KILOWATTS	CRYOGENIC OXYGEN COMPRESSOR FOR SUPPLY TO COAL GASIFIER
2	-5240.3	KILOWATTS	2 U-GAS GASIFIER NORMAL OPERATION. STREAM #2 = NET MAKE.
26	-6660.1	KILOWATTS	26 REMOVE SULFUR FROM RAW GAS BY SELEXOL SYSTEM.
37	6282.9	KILOWATTS	37 ADD STEAM VIA TURBINE; PREVENT CARBON DROP IN FUEL CELL.
41	4172.8	KILOWATTS	41 EXPAND SWEET FUEL GAS THROUGH POWER RECOVERY TURBINE.
46	-179574.5	KILOWATTS	46 COMPRESS COMBUSTION AIR TO FUEL CELL PRESSURE.
54	66599.8	KILOWATTS	NORMAL FUEL CELL, 1200 F INPUT SETS VOLTAGE, 1260 F EXIT.
61	-22106.3	KILOWATTS	61 CATHODE RECYCLE GAS CIRCULATION FAN.
63	266304.0	KILOWATTS	63 EXPAND CATHODE EXIT STACK GAS TO RECOVER POWER.
72	-8491.2	KILOWATTS	72 ANODE EXIT GAS BOOSTER FAN, TO FEED CATALYTIC BURNER.
74	-304.9	KILOWATTS	74 COMPRESS COAL FEED INJECTION GAS, A SMALL RECYCLE.

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RUN.B1
STEAM LIST
5 FEB 83 10:20:02

DUSTY SHIFT BASE CASE, FC REFUMK, 13.5% CARB EX COAL TO CH4

TABULATION OF STEAM NEEDED IN PROCESS.

2	550.0 PSIA	24072. MOLS STEAM	2 U-GAS GASIFIER NORMAL OPERATION. STREAM #2 = NET MAKE.
26	65.0 PSIA	7276. MOLS STEAM	26 REMOVE SULFUR FROM RAW GAS BY SELEXOL SYSTEM.
37	1500.0 PSIA	7988. MOLS STEAM	37 ADD STEAM VIA TURBINE! PREVENT CARBON DROP IN FUEL CELL.

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TABULATION OF WASTE HEAT AVAILABLE FROM PROCESS BOILERS.

15	600.0	DEGREES F.	256.28	MILLIONS RTU	15	COOL	RAW GAS IN WHB	MAKE STEAM.
18	600.0	DEGREES F.	65.63	MILLIONS RTU	18	COOL	INTER-STAGE SHIFT GAS IN WHB	RAISE STEAM.
60	880.0	DEGREES F.	432.11	MILLIONS RTU	60	COOL	CATHODE RECYCLE GAS IN WHB	RAISE STEAM.
71	880.0	DEGREES F.	383.62	MILLIONS RTU	71	COOL	ANODE EXIT GAS	RAISE STEAM.

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CASE VI

High-Temperature Dust Removal

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QEX.2 HIGH TEMPERATURE DE-DUSTING, RAW GAS LIKE JPL4, QUENCH AFTER TURBINE.

TABULATION OF SHAFT ENERGY AVAILABLE OR NEEDED.

102	-39544.7	KILOWATTS	AIR COMPRESSOR FOR SUPPLY TO CRYOGENIC OXYGEN PLANT
104	-15426.3	KILOWATTS	CRYOGENIC OXYGEN COMPRESSOR FOR SUPPLY TO COAL GASIFIER
2	-5240.3	KILOWATTS	2 U-GAS GASIFIER NORMAL OPERATION. STRFAM #2 = NET MAKE.
5	55373.3	KILOWATTS	5 EXPAND HOT RAW GAS VIA TURBINE; COLLECT KILOWATTS.
21	-6647.2	KILOWATTS	21 REMOVE SULFUR FROM RAW GAS BY SELEXOL SYSTEM.
34	10426.1	KILOWATTS	34 ADD STEAM VIA TURBINE; PREVENT CARBON DROP IN FUEL CELL.
36	-171083.5	KILOWATTS	36 COMPRESS COMBUSTION AIR TO FUEL CELL PRESSURE.
43	668403.7	KILOWATTS	43 THE FUEL CELL, MAXIMUM CONVERSION, PERFORM WITHIN ANODE.
48	-8325.9	KILOWATTS	48 ANODE EXIT GAS BOOSTER FAN, TO FEED CATALYTIC BURNER.
54	-20313.7	KILOWATTS	54 CATHODE RECYCLE GAS CIRCULATION FAN.
55	231153.2	KILOWATTS	55 EXPAND STACK GAS THROUGH TURBINE TO RECOVER POWER.
63	-1792.6	KILOWATTS	63 COMPRESS RECYCLE GAS TO INJECT FEED COAL TO GASIFIER.

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CASE 9EX.2
STEAM LIST
16 FEB 83 17:10:29

9EX.2 HIGH TEMPERATURE DE-DUSTING, RAW GAS LIKE JPL4, QUENCH AFTER TURBINE.

TABULATION OF STEAM NEEDED IN PROCESS.

2	550.0 PSIA	24072. MOLS STEAM	2 U-GAS GASIFIER NORMAL OPERATION. STREAM #2 = NET MAKE.
21	63.0 PSIA	11997. MOLS STEAM	21 REMOVE SULFUR FROM RAW GAS BY SELEXOL SYSTEM.
34	1500.0 PSIA	7094. MOLS STEAM	34 ADD STEAM VIA TURBINE; PREVENT CARBON DROP IN FUEL CELL.

CASE QEX.2
WASTE HEAT OUT
16 FEB 83 17:10:29

QEX.2 HIGH TEMPERATURE DE-DUSTING, RAW GAS LIKE JPLG, QUENCH AFTER TURBINE.

TABULATION OF WASTE HEAT AVAILABLE FROM PROCESS BOILERS.

	DEGREES F.	MILLIONS BTU	DEGREES F.	MILLIONS BTU	DEGREES F.	MILLIONS BTU
41	600.0	135.94	11	COOL SHIFT FEED GAS TO 600F, RAISE STEAM.		
13	600.0	119.32	13	COOL INTER-STAGE SHIFT GAS IN WHB? RAISE STEAM.		
46	925.0	134.93	46	COOL ANODE EXIT GAS IN WHB - RAISE STEAM.		
51	1115.0	137.43	51	COOL CATHODE RECYCLE GAS IN WHB - RAISE STEAM.		

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CASE VII

Sulfur-Tolerant Fuel Cell

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CASE L. THICEL
POWER LIST
30 MAR 83 8:52:03

L. THICEL SULFUR-TOLERANT CELL, 95% CONV, INT. REFORM, NO PRECOOL, 13.5% CH4 RAW

TABULATION OF SHAFT ENERGY AVAILABLE OR NEEDED.

102	-33844.7 KILOWATTS	AIP COMPRESSOR FOR SUPPLY TO CRYOGENIC OXYGEN PLANT
104	-16426.3 KILOWATTS	CRYOGENIC OXYGEN COMPRESSOR FOR SUPPLY TO COAL GASIFIER
2	-5240.3 KILOWATTS	2 U-GAS GASIFIER NORMAL OPERATION. STREAM #2 = NET MAKE.
5	69393.7 KILOWATTS	5 EXPAND RAW GAS TO FUEL CELL PRESSURE; COLLECT KILOWATTS.
18	-1309.7 KILOWATTS	18 COMPRESS COAL INJECTION SIDESTREAM RECYCLE TO GASIFIER.
22	-171083.5 KILOWATTS	22 COMPRESS COMBUSTION AIR TO FUEL CELL PRESSURE.
29	652392.1 KILOWATTS	29 THE FUEL CELL, MAXIMUM CONVERSION, REFORM WITHIN ANODE.
33	7563.6 KILOWATTS	33 ANODE EXIT GAS BOOSTER FAN, TO FEED CATALYTIC BURNER.
36	284449.3 KILOWATTS	36 EXPAND STACK GAS FROM GASIFIER PRESSURE; COLLECT KW'S.
41	-58244.9 KILOWATTS	41 CATHODE RECYCLE GAS CIRCULATION FAN.

CASE L. THICEL
STEAM LIST
30 MAR 83 8:52:03

L. THICEL SULFUR-TOLERANT CELL, 85% CONV, SWI, #2900W, NO PRECOOL, 13.5% CH4 RAW

TABULATION OF STEAM NEEDED IN PROCESS.

2 550.0 PSIA 24072. WOLS STEAM 2 U-GAS GASIFIER NORMAL OPERATION. STREAM #2 = NET MAKE.

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CASE L. THICEL
WASTE HEAT OUT
30 MAR 83 8:52:03

L. THICEL SULFUR-TOLERANT CELL, 85% CONV, INT. REFORM, NO PRECOOL, 13.5X CH4 RAW

TABULATION OF WASTE HEAT AVAILABLE FROM PROCESS BOILERS.

8	1200.0 DEGREES F.	3.04 MILLIONS BTU	9 COOL RAW GAS IN WHB; MAKE STEAM.
32	900.0 DEGREES F.	170.34 MILLIONS BTU	32 COOL ANODE EXIT GAS IN WH9 - RAISE STEAM.
38	1010.0 DEGREES F.	867.24 MILLIONS BTU	38 COOL CATHODE EXIT RECYCLE GAS IN WHB; RAISE STEAM.



CASE VIII

High-Temperature CO₂ Transfer

CASE RUN.D2
POWER LIST
17 FEB 83 7:18:13

RUN.D2 HIGH TEMP. CO2 PASS ADDED TO BASE CASE, COMPARE AIR.4E

TABULATION OF SHAFT ENERGY AVAILABLE OR NEEDED.

102	-39844.7 KILOWATTS	AIR COMPRESSOR FOR SUPPLY TO CRYOGENIC OXYGEN PLANT
104	-16426.3 KILOWATTS	CRYOGENIC OXYGEN COMPRESSOR FOR SUPPLY TO COAL GASIFIER
2	-5240.3 KILOWATTS	2 U-GAS GASIFIER NORMAL OPERATION. STREAM #2 = NET MAKE.
29	-173461.0 KILOWATTS	29 COMPRESS COMBUSTION AIR TO FUEL CELL PRESSURE.
40	29297.1 KILOWATTS	40 EXPAND SHIFT (1) EXIT GAS THROUGH POWER RECOVERY TURBINE.
49	-3144.6 KILOWATTS	49 REMOVE SULFUR FROM RAW GAS BY SELEXOL SYSTEM.
63	911.2 KILOWATTS	63 ADD STEAM VIA TURBINE; PREVENT CARBON DROP IN FUEL CELL.
69	706989.1 KILOWATTS	NORMAL FUEL CELL, 1200 F INPUT SETS VOLTAGE, 1260 F EXIT.
76	-17225.6 KILOWATTS	76 CATHODE RECYCLE GAS CIRCULATION FAN.
80	-7145.2 KILOWATTS	80 ANODE EXIT GAS BOOSTER FAN, TO FEED CATALYTIC BURNER.
81	241657.8 KILOWATTS	81 EXPAND CATHODE EXIT STACK GAS TO RECOVER POWER.
86	-1502.6 KILOWATTS	86 COMPRESS COAL FEED INJECTION GAS, A SMALL RECYCLE.

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CASE 004.02
STEAM LIST
17 FEB 03 7:10:13

UNID. WITH TEMP. CASE PASS ADDED TO CASE COMPARE AIR.46

TABULATION OF STEAM NEEDED IN PROCESS.

2	550.0 PSIA	24072. MOLS STEAM	2 U-GAS GASIFIER NORMAL OPERATION. STEAM #2 = NET MAKE.
36	1200.0 PSIA	5175. MOLS STEAM	36 HIGH-TEMPERATURE CO2 REMOVAL FROM SHIFT 1 EXIT GAS.
37	1200.0 PSIA	1559. MOLS STEAM	37 REGEN OF H2O PUTS CO2 IN CATH FEED AIR.
49	65.0 PSIA	0000. MOLS STEAM	49 REMOVE SULFUR FROM RAW GAS BY SELEXOL SYSTEM.
63	1500.0 PSIA	589. MOLS STEAM	63 ADD STEAM VIA TURBINE; PREVENT CARBON DROP IN FUEL CELL.

CASE 0UN.D2
WASTE HEAT OUT
17 FEB 83 7:19:13

AIR.6E

RUN.D2 HIGH TEMP. CO2 PASS ADDED TO BASE CASE, COMPARE

TABULATION OF WASTE HEAT AVAILABLE FROM PROCESS BOILERS.

8	1420.0	DEGREES F.	251.07	MILLIONS BTU	8 COOL HOT SOUP RAW GAS IN WHB; RAISE STEAM. (ONCE TO #30)
15	590.3	DEGREES F.	237.34	MILLIONS BTU	15 COOL QUENCHED RAW GAS IN WHB; MAKE STEAM.
20	370.0	DEGREES F.	35.97	MILLIONS BTU	20 COOL RAW GAS; PRE-HEAT BOILER FEED WATER.
35	750.0	DEGREES F.	42.64	MILLIONS BTU	35 COOL SHIFT 1 EXIT GAS IN WH3; RAISE STEAM.
36	853.9	DEGREES F.	273.24	MILLIONS BTU	36 HIGH-TEMPERATURE CO2 REMOVAL FROM SHIFT 1 EXIT GAS.
41	600.0	DEGREES F.	23.23	MILLIONS BTU	41 COOL INTER-STAGE SHIFT GAS IN WHB; RAISE STEAM.
75	940.0	DEGREES F.	237.60	MILLIONS BTU	75 COOL CATHODE RECYCLE GAS IN WHB - RAISE STEAM.
79	890.0	DEGREES F.	221.28	MILLIONS BTU	79 COOL ANODE EXIT GAS IN WHB; RAISE STEAM.

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CASE IX

High-Temperature Sulfur Removal

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CASE RUN-C3
POWER LIST
28 FEB 93 14:11:14

RUN-C3 HIGH TEMP. H2S & DUST REMOVAL, MAKE MOLTEN SULFUR FROM OXYGEN.

TABULATION OF SHAFT ENERGY AVAILABLE OR NEEDED.

102	-39844.7	KILOWATTS	AIR COMPRESSOR FOR SUPPLY TO CRYOGENIC OXYGEN PLANT
104	-16426.3	KILOWATTS	CRYOGENIC OXYGEN COMPRESSOR FOR SUPPLY TO COAL GASIFIER
2	-5240.3	KILOWATTS	2 U-GAS GASIFIER NORMAL OPERATION. STREAM #2 = NET MAKE.
112	-63.7	KILOWATTS	OXYGEN COMPRESSOR FOR REGENERATING SPINEL; FOR HOT H2S DELETE
15	-1240.0	KILOWATTS	SUPPLY OF 96. T/D EXTRA OXYGEN FOR HOT H2S REMOVAL.
113	-768.8	KILOWATTS	COMPRESSOR FOR RECYCLING DILUENT GAS; FOR HOT H2S REMOVAL.
16	48191.4	KILOWATTS	16 EXPAND SULFUR-LEAN RAW GAS TO CELL PRESSURE; GET POWER.
27	-171215.0	KILOWATTS	27 COMPRESS COMBUSTION AIR TO FUEL CELL PRESSURE.
35	651841.6	KILOWATTS	NORMAL FUEL CELL, 1200 F INPUT SETS VOLTAGE, 1260 F EXIT.
42	-26272.5	KILOWATTS	42 CATHODE RECYCLE GAS CIRCULATION FAN.
46	-7620.2	KILOWATTS	46 ANODE EXIT GAS BOOSTER FAN, TO FEED CATALYTIC BURNER.
48	283643.5	KILOWATTS	48 EXPAND CATHODE EXIT STACK GAS TO RECOVER POWER.
55	-1623.7	KILOWATTS	55 COMPRESS COAL FEED INJECTION GAS, A SMALL RECYCLE.

C-4

RUN.C3 HIGH TEMP. H2S & DUST REMOVAL, MAKE MOLTEN SULFUR FROM OXYGEN.

CASE RUN.C3
STEAM LIST
28 FEB 83 14:11:14

TABULATION OF STEAM NEEDED IN PROCESS.

2	550.0 PSIA	24072. MOLS STEAM	2 U-GAS GASIFIER NORMAL OPERATION. STREAM #2 = NET MAKE.
14	1200.0 PSIA	192. MOLS STEAM	14 ADSORB H2S ONTO ZINC SPINEL; A NEW IGT DEVELOPMENT.
15	1200.0 PSIA	59. MOLS STEAM	15 RECLAIM ZN SPINEL (OXIDE) BY OXYGEN; MAKE SULFUR.

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CASE RUN.C3
 WASTE HEAT OUT
 28 FEB 83 14:11:14

RUN.C3 HIGH TEMP. H2S & DUST REMOVAL, MAKE MOLTEN SULFUR FROM OXYGEN.

TABULATION OF WASTE HEAT AVAILABLE FROM PROCESS BOILERS.

8	1500.0	DEGREES F.	209.97	MILLIONS BTU	9 COOL HOT SOUR RAW GAS IN WH9; RAISE STEAM.
14	1200.0	DEGREES F.	-26.24	MILLIONS BTU	14 ADSORB H2S ONTO ZINC SPINEL; A NEW IGT DEVELOPMENT.
15	1400.0	DEGREES F.	-4.39	MILLIONS BTU	COOLING COIL IN FLUID BED SPINEL REGENERATOR.
15	260.0	DEGREES F.	15.32	MILLIONS BTU	SULFUR CONDENSER WHICH GENERATES 15 # STEAM.
19	750.0	DEGREES F.	37.95	MILLIONS BTU	SULFUR-LEAN RAW GAS; MAKE STEAM.
41	880.0	DEGREES F.	642.19	MILLIONS BTU	41 COOL CATHODE RECYCLE GAS IN WHB; RAISE STEAM.
45	880.0	DEGREES F.	309.66	MILLIONS BTU	45 COOL ANODE EXIT GAS IN WHB; RAISE STEAM.
53	360.0	DEGREES F.	15.21	MILLIONS BTU	53 COOL SMALL COAL INJECTION RECYCLE STREAM #20; MAKE STEAM.

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CASE X

Anode Exhaust Recycle

TABULATION OF SHAFI ENERGY AVAILABLE, OR NEEDED.

97	-39844.7	KILOWATTS	AIR COMPRESSOR FOR SUPPLY TO CRYOGENIC OXYGEN PLANT
99	-16426.3	KILOWATTS	CRYOGENIC OXYGEN COMPRESSOR FOR SUPPLY TO COAL GASIFIER
2	-5240.3	KILOWATTS	2 U-GAS GASIFIER NORMAL OPERATION, STREAM #2 - NET MAKE.
22	46127.9	KILOWATTS	22 EXPAND SHIFT (1) EXIT GAS THROUGH POWER RECOVERY TURBINE.
39	-20393.4	KILOWATTS	39 COMPRESS SOUR FUEL GAS FOR FUEL CELL ANODE FEED.
40	-24762.2	KILOWATTS	40 REMOVE SULFUR FROM RAW GAS BY STAGE I SELEXOL SYSTEM.
41	37285.6	KILOWATTS	41 GEI LP STEAM NEEDS OF STEP #40 BY EXPANDING HP STEAM.
45	-46180.2	KILOWATTS	45 REMOVE CARBON DIOXIDE FROM RAW GAS BY SELEXOL STAGE II.
46	84796.0	KILOWATTS	46 EXPAND HP STEAM TO GET POWER & LP STEAM FOR STEP #45.
54	-171818.5	KILOWATTS	54 COMPRESS COMBUSTION AIR TO FUEL CELL PRESSURE.
63	821474.4	KILOWATTS	63 FUEL CELL, 88% CONV./PASS, W. REFORM & ANODE RECYCLE.
70	247452.0	KILOWATTS	70 EXPAND STACK GAS FROM CATHODE EXIT PRESSURE; MAKE KWATTS.
76	-2468.9	KILOWATTS	76 COMPRESS COAL FEEL INJECTION GAS SIDESTREAM RECYCLE #27.
78	-75671.4	KILOWATTS	78 COMPRESS CARBON DIOXIDE FROM SELEXOL II STEP #45.

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TABULATION OF STEAM NEEDED IN PROCESS.

2	500.0 PSIA	24072. MOLS STEAM	2 U-GAS GASIFIER NORMAL OPERATION, SIREAM #2 = NET MAKE.
40	65.0 PSIA	22142. MOLS STEAM	40 REMOVE SULFUR FROM RAW GAS BY STAGE 1 SELEXOL SYSTEM.
40	1500.0 PSIA	0. MOLS STEAM	41 GET LP STEAM NEEDS OF STEP #40 BY EXPANDING HP STEAM.
45	65.0 PSIA	50356. MOLS STEAM	45 REMOVE CARBON DIOXIDE FROM RAW GAS BY SELEXOL STAGE 11.
45	1500.0 PSIA	0. MOLS STEAM	46 EXPAND HP STEAM TO GET POWER & LP STEAM FOR STRY #45.

TABULATION OF WASTE HEAT AVAILABLE FROM PROCESS BOILERS.

26	600.0 DEGREES F.	253.33 MILLIONS BTU	26 COOL INTER-STAGE SHIFT (CYCLE) GAS IN WHBI RAISE STEAM.
28	600.0 DEGREES F.	69.63 MILLIONS BTU	28 COOL INTER-STAGE SHIFT (CYCLE) GAS IN WHBI RAISE STEAM.
33	360.0 DEGREES F.	94.81 MILLIONS BTU	33 COOL SHIFTED CYCLE GAS (SOUR FUEL) PREHEAT BFW.
71	680.0 DEGREES F.	101.60 MILLIONS BTU	71 COOL STACK GAS IN WHBI RAISE STEAM.
75	360.0 DEGREES F.	184.79 MILLIONS BTU	75 STACK GAS ECONOMIZER WARMS BOILER FEED WATER.

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APPENDIX Cd. Complete System Computer Printout
for Case 5 (Similar to Case Ca-1Ve, Da-C)

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TABULATION OF SHAFT ENERGY AVAILABLE ON NEEDED.

97	-39844.7	KILOWATTS	AIR COMPRESSOR FOR SUPPLY TO CRYOGENIC OXYGEN PLANT
99	-16426.3	KILOWATTS	CRYOGENIC OXYGEN COMPRESSOR FOR SUPPLY TO COAL GASIFIER
2	-5240.3	KILOWATTS	U-GAS GASIFIER NORMAL OPERATION. STREAM #2 = NET MAKE.
20	31756.2	KILOWATTS	EXPAND SHIFT (I) EXIT GAS THROUGH POWER RECOVERY TURBINE.
29	-6958.0	KILOWATTS	REMOVE SULFUR FROM RAW GAS BY SELEXOL SYSTEM.
42	10667.7	KILOWATTS	ADD STEAM VIA TURBINE PREVENT CARBON DROP IN FUEL CELL.
44	-174015.4	KILOWATTS	COMPRESS COMBUSTION AIR TO FUEL CELL PRESSURE.
51	664628.2	KILOWATTS	FUEL CELL, MAXIMUM CONVERSION, REFORM WITHIN ANODE.
57	-8324.9	KILOWATTS	ANODE EXIT GAS BOOSTER FAN, TO FEED CATALYTIC BURNER.
63	-15467.8	KILOWATTS	CATHODE RECYCLE GAS CIRCULATION FAN.
64	285486.4	KILOWATTS	EXPAND STACK GAS THROUGH TURBINE TO RECOVER POWER.
70	-1797.3	KILOWATTS	COMPRESS RECYCLE GAS TO INJECT FEED COAL TO GASIFIER.



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AIR4.160
CASE AIR4.160
STEAM LIST
14 JAN 83 912114

AIR4.160 JPL4 2-STAGE AIR COMPRESSOR, WITH AIR FEED @ 160.0% OF STOICHIOMETRIC

TABULATION OF STEAM NEEDED IN PROCESS.

2	500.0 PSIA	24072. MOLS STEAM	2 U-GAS GASIFIER NORMAL OPERATION. STREAM #2 = NET MAKE.
29	65.0 PSIA	11869. MOLS STEAM	29 REMOVE SULFUR FROM RAW GAS BY SELEXOL SYSTEM.
42	1500.0 PSIA	8015. MOLS STEAM	42 ADD STEAM VIA TURBINE; PREVENT CARBON DROP IN FUEL CELL.

CASE AIR4.160
WASTE HEAT OUT
14 JAN 83 9:21:14

AIR4.160 JPL4. 2-STAGE AIR COMPRESSOR, WITH AIR FEED @ 160.0% OF STOICHIOMETRIC

TABULATION OF WASTE HEAT AVAILABLE FROM PROCESS BUILERS.

	9	21	54	60
700.0 DEGREES F.	458.66 MILLIONS BTU	9 COOL RAW GAS TO 700 DEGREES FAHRENHEIT, MAKE STEAM.		
600.0 DEGREES F.	10.06 MILLIONS BTU	21 COOL INTER-STAGE SHIFT GAS IN WHBI RAISE STEAM.		
425.0 DEGREES F.	338.74 MILLIONS BTU	54 COOL ANODE EXIT GAS IN WHB - RAISE STEAM.		
1115.0 DEGREES F.	157.28 MILLIONS BTU	60 COOL CATHODE RECYCLE GAS IN WHB - RAISE STEAM.		

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28)	1	28 COOL TO AMBIENT; DROP CONDENSATE; DUMP HEAT TO AIR & CW.
100.	173.	
29)	3	29 REMOVE SULFUR FROM RAW GAS BY SELEXOL SYSTEM.
100.	168.	
30)	4	0.19 SPLIT RECYCLE GAS #30 FOR COAL FEED; GAS #31 GOES FORWARD.
100.	168.	
32)	5	0.075 SPLIT RECYCLE GAS #30 FOR COAL FEED; GAS #31 GOES FORWARD.
220.	166.	
33)	5	32 H27132 WARM SWEETENED GAS; COOL SHIFT EXIT #27.
340.	164.	
34)	5	33 H25133 WARM SWEETENED GAS; COOL SHIFT EXIT #25.
460.	162.	
35)	5	34 H24134 WARM SWEETENED GAS; COOL SHIFT EXIT #24.
580.	160.	
36)	5	35 H23135 WARM SWEETENED GAS; COOL SHIFT EXIT #23.
750.	158.	
37)	7	36 R36F55 WARM SWEETENED GAS; COOL FUEL CELL ANODE EXIT #55.
750.	153.	
38)	11	37 REMOVE LAST TRACES OF SULFUR BY ZINC OXIDE METATHESIS.
1200.	150.	
39)	11	38 TEST GAS FOR DANGER OF CARBON DEPOSITION.
1400.	150.	
40)	11	39 TEST CARBON DEPOSITION AT ANOTHER TEMPERATURE.
1000.	150.	
41)	11	40 TEST CARBON DEPOSITION AT ANOTHER TEMPERATURE.
900.	150.	
42)	3723	41 TEST CARBON DEPOSITION AT ANOTHER TEMPERATURE.
1000.	1500.	
43)	30	42 ADD STEAM VIA TURBINE; PREVENT CARBON DROP IN FUEL CELL.
90.	14.3	
44)	20	92.0 43 FEED AIR AT 160.0% OF STOICHIOMETRIC. (INCLUDING SCALE).
270.	164.	365.61 44 COMPRESS COMBUSTION AIR TO FUEL CELL PRESSURE.
45)	5	45 H45F65 - WARM COMBUSTION AIR FROM STACK GAS #65.
480.	162.	
46)	5	46 H46F62 - WARM COMBUSTION AIR FROM CATHODE RECYCLE #62.
700.	160.	
47)	5	47 H47F61 - WARM COMBUSTION AIR FROM CATHODE RECYCLE #61.
910.	158.	
48)	3	48 MERGE FRESH AIR #47 WITH RECYCLED CATHODE EXIT GAS #63.
1000.	156.	
49)	5713	49 BURN ANODE EXHAUST GAS #57 WITH DILUTED AIR #48.
1200.	152.	
50)	4203	50 JOIN THE TWO FUEL CELL FEEDS, ANODE #42, CATHODE #49.
1150.	152.	
51)	14	51 THE FUEL CELL, MAXIMUM CONVERSION, REFORM WITHIN ANODE.
1200.	150.	.85
52)	4	SEPARATE ANODE EXIT GAS, #52, FROM CATHODE EXIT GAS, #53.
1200.	145.	
54)	5205	54 COOL ANODE EXIT GAS IN WHB - RAISE STEAM.
925.	143.	
55)	5	55 R36F55 COOL ANODE EXIT GAS TO WARM ANODE FEED GAS #36.
855.	141.	
56)	26	56 THIS STEP DELETED.
100.	140.	
57)	5519	57 ANODE EXIT GAS BOOSTER FAN, TO FEED CATALYTIC BURNER.
885.	156.	
58)	5304	SPLIT CATHODE EXIT, COOLED RECYCLE (#58), STACK EXIT (#59).
1250.	142.	0.396
60)	5805	60 COOL CATHODE RECYCLE GAS IN WHB - RAISE STEAM.
1115.	139.	
61)	5	61 H47F61 COOL CATHODE RECYCLE TO WARM COMBUSTION AIR #47.
465.	137.	-47
62)	5	62 H46F62 COOL CATHODE RECYCLE TO WARM COMBUSTION AIR #46.
880.	135.	-46
63)	20	63 CATHODE RECYCLE GAS CIRCULATION FAN.
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64)	5922	1	64 EXPAND STACK GAS THROUGH TURBINE TO RECOVER POWER.
	580.	17.0	
65)	5	1	65 H45F65 COOL STACK GAS TO WARM COMBUSTION AIR #45.
	548.	15.0	
66)	212	3	66 HEAD COMBUSTION FUEL VALUE OF RAW GAS FROM GASIFIER.
	100.		
67)	3112	3	67 HEAD COMBUSTION FUEL VALUE OF SHIFTED SOUR GAS.
	100.		
68)	3712	3	68 HEAD COMBUSTION FUEL VALUE OF SWEET ANODE FEED GAS.
	100.		
69)	5412	3	69 HEAD COMBUSTION FUEL VALUE OF ANODE EXIT GAS.
	100		
70)	3019	2	70 COMPRESS RECYCLE GAS TO INJECT FEED COAL TO GASIFIER.
	250.	550.	
71)	26	1	71 DUMMY STEP TO PUT RECYCLE GAS AT GASIFIER TEMPERATURE.
	1675.		
72)	12	3	72 HEAD COMBUSTION FUEL VALUE OF RECYCLE COAL INJECTION GAS.
	100.		
73)	305	1	73 HEAD SENSIBLE HEATS OF REACTOR EXIT RAW GAS (INCL. #71)
	1507.8	500.	
74)	25	3	METH-RICH U-GAS WITH ANODE METHANE REFORMING. 13.58 C TO CH4
	30.		

AIR4.160 JPL4. 2-STAGE AIR COMPRESSOR, WITH AIR FEED @ 160.0% OF STOICHIOMETRIC
CASE: AIR4.160
STEP 97
14 JAN 83 9121114

AIR COMPRESSOR FOR SUPPLY TO CHYUGENIC OXYGEN PLANT

THIS STEP DESCRIBES A PROCESS GAS COMPRESSOR OF 2 STAGES.

COMPONENT	STREAM NUMBER 97	MOLS	%
CU2	0.0	0.0	
H2O	1185.1	2.5	
N2	37015.3	77.1	
O2	9778.0	20.4	
TOTAL	47978.3	100.0	

THE COMPRESSION RATIO (PER STAGE) IS 2.63

ENERGY REQUIREMENTS
PER STAGE TOTAL
HORSEPOWER 26725. 53449.
KILOWATTS 19922. 39845.

ISENTROPIC GAS COMPRESSION EFFICIENCY IS 88.0 %. (SET BY USER)
MECHANICAL EFFICIENCY ASSUMED IS 98.0%

AT EACH STAGE SUCTION TEMPERATURE IS: 95.0 DISCHARGE IS: 292.2 DEGREES F.

INTERSTAGE COOLING DUTY IS: 67.99 MILLIONS BTU

PRESSURE (PSIA):	SUCTION DISCHARGE	VOLUME (ACFH):	SUCTION DISCHARGE
STAGE # 1	14.5 38.1	316426.1	163194.7
STAGE # 2	32.3 85.0	141849.4	73157.9

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CASE AIH4.160
STEP 99
14 JAN 83 9121114

AIH4.160 JPL4, 2-STAGE AIR COMPRESSOR, WITH AIR FEED @ 160.0% OF STOICHIOMETRIC

CMBYGENIC OXYGEN COMPRESSOR FOR SUPPLY TO COAL GASIFIER

THIS STEP DESCRIBES A PROCESS GAS COMPRESSOR OF 5 STAGES.

COMPONENT	STREAM NUMBER	MOLS	%
CO2	0.0	0.0	0.0
H2O	0.0	0.0	0.0
N2	39.4	0.5	
O2	7831.8	99.5	
TOTAL	7871.2	100.0	

THE COMPRESSION RATIO (PER STAGE) IS 2.07

ENERGY REQUIREMENTS
PER STAGE TOTAL
HORSEPOWER 4407.
KILOWATTS 3285. 16426.

ISENTROPIC GAS COMPRESSION EFFICIENCY IS 62.0 %. (SET BY USER)
MECHANICAL EFFICIENCY ASSUMED IS 98.0%

AT EACH STAGE SUCTION TEMPERATURE IS: 80.0 DISCHARGE IS: 276.6 DEGREES F.

INTERSTAGE COOLING DUTY IS: 11.21 MILLIONS BTU

STAGE #	SUCTION DISCHARGE	VOLUME (ACFM):	SUCTION DISCHARGE
1	20.0 41.5	36618.4	24099.1
2	35.6 73.9	20555.3	13527.8
3	67.4 139.7	10869.4	7153.3
4	131.9 273.4	5553.4	3654.8
5	262.9 545.0	2785.7	1833.3

NON-LUBRICATED OXYGEN COMPRESSORS MUST BE OPERATED AT LOW COMPRESSION RATIOS AND HAVE LOW ISENTROPIC EFFICIENCIES.

AIR# 160 JPL4, 2-STAGE AIR COMPRESSOR, WITH AIR FEED # 160.0% OF STOICHIOMETRIC

2 U-GAS GASIFIER NORMAL OPERATION. STREAM #2 = NET MAKE.
 THIS STEP DESCRIBES AN OXYGEN-BLOWN COAL GASIFIER OF THE U-GAS TYPE.
 OPERATING TEMPERATURE: 1875.0 F OPERATING PRESSURE: 500.0 PSIA
 COAL FEED RATE: 416. THOUSANDS POUNDS PER HOUR OR 4991. TONS PER DAY
 THE AIR SEPARATION PLANT IS RATED AT 3023. TONS PER DAY.
 THE OXYGEN FEED STREAM IS:

COMPONENT	STREAM NUMBER 99	MOLS	%
CU2	0.0	0.0	0.0
H2O	0.0	0.0	0.0
N2	39.4	0.5	
O2	7831.8	99.5	
TOTAL	7871.2	100.0	

HIGH-PRESSURE STEAM REQUIREMENTS OF THIS COAL GASIFIER ARE: 24072. MOLS.
 OR APPROXIMATELY: 434. THOUSAND POUNDS PER HOUR.

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THIS GASIFIER MAKES THE FOLLOWING RAW GAS (NET):

COMPONENT	STREAM NUMBER 2	MOLS	%
H2	13633.2	25.9	
CH4	3249.8	6.2	
CO	12103.9	23.0	
CU2	7954.5	15.1	
H2O	14906.7	28.3	
NC	185.5	0.4	
C2H6	0.0	0.0	
H2S	477.9	0.9	
NH3	74.4	0.1	
GUS	13.7	0.0	

342

TOTAL 52599.7 100.0
 END OF REITERATIVE CYCLE NUMBER 1
 END OF REITERATIVE CYCLE NUMBER 2
 END OF REITERATIVE CYCLE NUMBER 3
 END OF REITERATIVE CYCLE NUMBER 4
 END OF REITERATIVE CYCLE NUMBER 5
 END OF REITERATIVE CYCLE NUMBER 6
 END OF REITERATIVE CYCLE NUMBER 7

CASE AIR4.160
 STEP 3
 14 JAN 83 9:21:14

AIR4.160 JPL4. 2-STAGE AIR COMPRESSOR, WITH AIR FEED @ 160.0% OF STOICHIOMETRIC

RECYCLE GAS TO INJECT COAL FEED. STREAM #3 = TOTAL EXIT GAS.

THIS STEP ADDS STREAM # 2 & 71 TO PRODUCE STREAM # 3

COMPONENT	FEED STREAM 2 MOLS	FEED STREAM 71 MOLS	FEED STREAM 71 %	EXIT STREAM 3 MOLS	EXIT STREAM 3 %
H2	13633.2	1872.2	51.9	15505.4	27.6
CH4	3249.8	263.3	7.3	3513.2	6.3
CO	12103.9	214.5	5.9	12318.5	21.9
CO2	7954.5	1243.1	34.5	9197.6	16.4
H2O	14906.7	0.0	0.0	14906.7	26.5
N2	185.5	15.0	0.4	200.5	0.4
C2H6	0.0	0.0	0.0	0.0	0.0
H2S	477.9	0.0	0.0	477.9	0.9
NH3	74.4	0.0	0.0	74.4	0.1
CO5	13.7	0.0	0.0	13.7	0.0
TOTAL	52599.7	3608.3	100.0	56208.0	100.0

TEMPERATURE 1875.0 1875.0
 PRESSURE 500.0 500.0

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AIR4.160 JPL4, 2-STAGE AIR COMPRESSOR, WITH AIR FEED @ 160.0% OF STOICHIOMETRIC

 4 TEST CARBON LIMITS TO FIND NECESSARY QUENCH.

THIS IS A TEST WHETHER THE GAS IS STABLE WITH RESPECT TO CARBON DEPOSITION.

GAS COMPOSITION FROM STREAM # 3
 CARBON STABILITY REPORTED AS STREAM # 4 TEMPERATURE: 1200.0 DEGREES F
 PRESSURE: 490.0 PSIA

	GAS STREAM MOLS	GAS STREAM		GRAPHITE		AMORPHOUS	
		MOL FRACTION	MOL FRACTION	LIMIT	MOL FRACTION	LIMIT	MOL FRACTION
H2	15505.4	0.2759	0.1383	0.1367			
CH4	3513.2	0.0625	0.1524	0.1604			
CO	12318.5	0.2192	0.0514	0.0606			
CO2	9197.6	0.1636	0.2857	0.2869			
H2O	14906.7	0.2652	0.3723	0.3554			
N2	200.5	0.0036					
C2H6	0.0	0.0000					
H2S	477.9	0.0085					
NH3	74.4	0.0013					
CO5	13.7	0.0002					
TOTAL	56208.0	1.0000	1.0000	1.0000			1.0000
		H*	0.5145	H*	0.5234	H*	0.5205
		C*	0.1720	C*	0.1571	C*	0.1626
		O*	0.3135	O*	0.3194	O*	0.3169

GAS HUMIDIFICATION REQUIREMENTS TO ACHIEVE STABILITY ARE CALCULATED BELOW

GAS IS UNSTABLE. 3676.54 MOLS STEAM NEEDED.
 STEAM INJECTION CAN DILUTE CARBON CONTENT TO HALFWAY BETWEEN GRAPHITE AND AMORPHOUS CARBON LIMIT.
 AT 1100.0 DEGREES F. AND 490.0 PSIA GAS NEEDS 4622.66 MOLS WATER. 0.1622 = AMORPHOUS CARBON LIMIT.
 AT 1000.0 DEGREES F. AND 490.0 PSIA GAS NEEDS 4399.06 MOLS WATER. 0.1677 = AMORPHOUS CARBON LIMIT.
 AT 900.0 DEGREES F. AND 490.0 PSIA GAS NEEDS 3001.31 MOLS WATER. 0.1801 = AMORPHOUS CARBON LIMIT.

CASE AIR4.160
 STEP 8
 14 JAN 83 9:21:14

AIR4.160 JPL4, 2-STAGE AIR COMPRESSOR, WITH AIR FEED @ 160.0% OF STOICHIOMETRIC

8 QUENCH TO STABILIZE GAS & PREVENT CARBON DEPOSITION.

THIS STEP IS AN ADIABATIC QUENCH OF HOT GAS WITH LIQUID WATER TO AVOID CARBON DEPOSITION.

COMPONENT	FEED STREAM 3 MOLS	%	EXIT STREAM 8 MOLS	%
H2	15505.4	27.6	15505.4	25.5
CH4	3513.2	6.3	3513.2	5.8
CO	12318.5	21.9	12318.5	20.3
CO2	9197.6	16.4	9197.6	15.1
H2O	14906.7	26.5	19529.4	32.1
N2	200.5	0.4	200.5	0.3
C2H6	0.0	0.0	0.0	0.0
H2S	477.9	0.9	477.9	0.8
NH3	74.4	0.1	74.4	0.1
CUS	13.7	0.0	13.7	0.0
TOTAL	56208.0	100.0	60830.6	100.0
TEMPERATURE	1875.0 F		1507.8 F	
PRESSURE	500.0 PSIA		495.0 PSIA	

QUENCH WATER ADDED 4622.7 MOLS

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CASE AIR4.160
STEP 9
14 JAN 83 912114

AIR4.160 JPL4, 2-STAGE AIR COMPRESSOR, WITH AIR FEED @ 160.0% OF STOICHIOMETRIC

9 COOL RAW GAS TO 700 DEGREES FAHRENHEIT, MAKE STEAM.

THE DUTY OF THIS PROCFS COOLER IS: -450.66 MILLIONS BTU PER HOUR.

COMPONENT	STREAM NUMBER	%
H2	15505.4	25.5
CH4	3513.2	5.8
CU	12318.5	20.3
CO2	9197.6	15.1
H2O	19529.4	32.1
N2	200.5	0.3
C2H6	0.0	0.0
H2S	477.9	0.8
NH3	74.4	0.1
CUS	13.7	0.0
TOTAL	60830.6	100.0

PROCESS GAS TEMPERATURE, FEED= 1507.8 EXIT= 700.0 DEGREES FAHRENHEIT.
PROCESS GAS ENTHALPY, FEED= 770.06 EXIT= 311.40 MILLIONS BTU.

CASE AIR4.160
 STEP 10
 14 JAN 83 9:21:14

AIR4.160 JPL4, 2-STAGE AIR COMPRESSOR, WITH AIR FEED @ 160.08 OF STOICHIOMETRIC

10 H10118 COOL HAW GAS WARM CLEAN GAS #18.

THE DUTY OF THIS PROCESS COOLER IS: -31.77 MILLIONS BTU PER HOUR.

COMPONENT	STREAM NUMBER 10	MOLS	%
H2	15505.4	25.5	
CH4	3513.2	5.8	
CU	12318.5	20.3	
CU2	9197.6	15.1	
H2O	19529.4	32.1	
N2	200.5	0.3	
C2H6	0.0	0.0	
H2S	477.9	0.8	
NH3	74.4	0.1	
CUS	13.7	0.0	
TOTAL	60830.6	100.0	

PROCESS GAS TEMPERATURE, FEED= 700.0 EXIT= 640.0 DEGREES FARENHEIT.
 PROCESS GAS ENTHALPY, FEED= 311.40 EXIT= 279.63 MILLIONS BTU.

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AIR4.160 JPL4, 2-STAGE AIR COMPRESSOR, WITH AIR FEED @ 160.0% OF STOICHIOMETRIC

CASE AIR4.160
 STEP 11
 14 JAN 83 9:21:14

 11 R11117 COOL WARM GAS; WARM CLEAN GAS #17.

THE DUTY OF THIS PROCESS COOLER IS: -31.42 MILLIONS BTU PER HOUR.

COMPONENT	STREAM NUMBER 11	MOLS	%
H2	15505.4	25.5	
CH4	3513.2	5.8	
CO	12318.5	20.3	
CO2	9197.6	15.1	
H2O	19529.4	32.1	
N2	200.5	0.3	
C2H6	0.0	0.0	
H2S	477.9	0.8	
NH3	74.4	0.1	
CUS	13.7	0.0	
TOTAL	60830.6	100.0	

PROCESS GAS TEMPERATURE, FEED= 640.0 EXIT= 580.0 DEGREES FAHRENHEIT.
 PROCESS GAS ENTHALPY, FEED= 279.63 EXIT= 248.20 MILLIONS BTU.

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AIR4.160 JPL4, 2-STAGE AIR COMPRESSOR, WITH AIR FEED @ 160.0% OF STOICHIOMETRIC

CASE AIR4.160
STEP 12
14 JAN 83 9:21:14

12 H12T16 COOL HAW GAS WARM CLEAN GAS #16.

THE DUTY OF THIS PROCESS COOLER IS: -31.07 MILLIONS BTU PER HOUR.

COMPONENT STREAM NUMBER 12
 MOLS %

H2	15505.4	25.5
CH4	3513.2	5.8
CO	12318.5	20.3
CO2	9197.6	15.1
H2O	19529.4	32.1
N2	200.5	0.3
C2H6	0.0	0.0
H2S	477.9	0.8
NH3	74.4	0.1
CUS	13.7	0.0

TOTAL 60830.6 100.0

PROCESS GAS TEMPERATURE,
PROCESS GAS ENTHALPY,

FEED= 580.0
FEED= 246.20

EXIT= 520.0 DEGREES FAHRENHEIT.
EXIT= 217.14 MILLIONS BTU.

AIR4.160 JPL4. 2-STATE AIR COMPRESSOR, WITH AIR FEED @ 160.0% OF STOICHIOMETRIC
 CASE AIR4.160
 STEP 13
 14 JAN 83 9121114

 13 RIJTI'S COOL HAW GAS! WARM CLEAN GAS #15.

THE DUTY OF THIS PROCESS COOLER IS: -30.70 MILLIONS BTU PER HOUR.

COMPONENT	STREAM NUMBER 13	MOLS	%
H2	15505.4	25.5	
CH4	3513.2	5.8	
CO	12318.5	20.3	
CO2	9197.6	15.1	
H2O	19529.4	32.1	
N2	200.5	0.3	
COH6	0.0	0.0	
H2S	477.9	0.8	
NH3	74.4	0.1	
CUS	13.7	0.0	
TOTAL	60830.6	100.0	

PROCESS GAS TEMPERATURE, FEED= 520.1
 PROCESS GAS ENTHALPY, FEED= 217.14
 EXIT= 460.0 DEGREES FAHRENHEIT.
 EXIT= 186.44 MILLIONS BTU.

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CASE AIR4.160
STEP 14
14 JAN 83 9121114

AIR4.160 JPL4, 2-STAGE AIR COMPRESSOR, WITH AIR FEED @ 160.0% OF STOICHIOMETRIC

14 COOL MOIST RAW GAS; SCHUB DUST; REMOVE COND. AND NH3.

THIS STEP SUBTRACTS CONDENSATE FROM THE GAS AS DIRECTED BY USER.

COMPONENT	FEED STREAM 13 MOLS	EXIT STREAM 14 MOLS
H2	15505.4	25.8
CH4	3513.2	5.8
CU	12318.5	20.5
CU2	9197.6	15.3
H2O	19529.4	31.4
N2	200.5	0.3
C2H6	0.0	0.0
H2S	477.9	0.8
NH3	74.4	0.0
CUS	13.7	0.0
TOTAL	60830.6	60132.3

TEMPERATURE 460.0 F 358.3 F
PRESSURE 482.0 PSIA 477.0 PSIA

GAS DEW POINT 360.1 F 358.3 F
WATER SUBTRACTED 698.3 MOLS
HEAT EXCHANGE DUTIES:
GAS COOLING 51.161 MILLIONS BTU
WATER CONDENSING 10.916 MILLIONS BTU
TOTAL 62.077 MILLIONS BTU

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CASE AIR4.160
STEP 15
14 JAN 83 9:21:14

AIR4.160 JPL4; 2-STAGE AIR COMPRESSOR, WITH AIR FEED @ 160.0% OF STOICHIOMETRIC

15 RIGHT15 WARM SHIFT FEED GAS FROM RAW GAS #13.

THIS STEP RECUPERATES HEAT AGAINST STEP # 13
THE ENTHALPY TRANSFERRED IS: 30.70 MILLIONS BTU

COMPONENT	STREAM NUMBER 15	MOLS	%
H2	15505.4	25.8	
CH4	3513.2	5.8	
CO	12318.5	20.5	
CO2	9197.6	15.3	
H2O	18905.5	31.4	
N2	200.5	0.3	
C2H6	0.0	0.0	
H2S	477.9	0.8	
NH3	0.0	0.0	
CS2	13.7	0.0	
TOTAL	60132.3	100.0	

PROCESS GAS TEMPERATURE: FEED= 358.3 EXIT= 420.3 DEGREES FAHRENHEIT.
PROCESS GAS ENTHALPY: FEED= 133.66 EXIT= 164.36 MILLIONS BTU.

CASE AIR4.160
 STEP 16
 14 JAN 83 9:21:14

AIR4.160 JPL4, 2-STAGE AIR COMPRESSOR, WITH AIR FEED = 160.08 OF STOICHIOMETRIC

16 K12T16 WARM SHIFT FEED GAS FROM RAW GAS #12.

THIS STEP RECUPERATES HEAT AGAINST STEP # 12
 THE ENTHALPY TRANSFERRED IS: 31.07 MILLIONS BTU

COMPONENT	STREAM NUMBER 16	MOLS	%
H2	15505.4	25.8	
CH4	3513.2	5.8	
CU	12319.5	20.5	
CU2	9197.6	15.3	
H2O	18905.5	31.4	
N2	200.5	0.3	
CH6	0.0	0.0	
H2S	477.9	0.8	
NH3	0.0	0.0	
CUS	13.7	0.0	
TOTAL	60132.3	100.0	

PROCESS GAS TEMPERATURE, FEED= 420.3 EXIT= 482.3 DEGREES FAHRENHEIT.
 PROCESS GAS ENTHALPY, FEED= 164.36 EXIT= 195.43 MILLIONS BTU.

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AIR4.160 JPL4, 2-STAGE AIR COMPRESSOR, WITH AIR FLED @ 160.0% OF STOICHIOMETRIC
 CASE AIR4.160
 STEP 17
 14 JAN 83 9121114

17 HLT17 WARM SHIFT FEED GAS FROM RAW GAS #11.
 THIS STEP RECUPERATES HEAT AGAINST STEP # 11
 THE ENTHALPY TRANSFERRED IS: 31.42 MILLIONS BTU

COMPONENT	STREAM NUMBER 17	MOLS	%
H2	15505.4	25.8	
CH4	3513.2	5.8	
CO	12318.5	20.5	
CU2	9197.6	15.3	
H2O	18905.5	31.4	
N2	200.5	0.3	
C2H6	0.0	0.0	
H2S	477.9	0.8	
NH3	0.0	0.0	
CUS	13.7	0.0	
TOTAL	60132.3	100.0	

PROCESS GAS TEMPERATURE: FEED= 482.3 EXIT= 544.1 DEGREES FAHRENHEIT.
 PROCESS GAS ENTHALPY: FEED= 195.43 EXIT= 226.85 MILLIONS BTU.

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CASE AIR4.160
STEP 18
14 JAN 83 9121114

AIR4.160 JPL4, 2-STAGE AIR COMPRESSOR, WITH AIR FEED @ 160.0% OF STOICHIOMETRIC

10 RI0110 WARM SHIFT FEED GAS FROM RAW GAS #10.

THIS STEP RECUPERATES HEAT AGAINST STEP # 10
THE ENTHALPY TRANSFERRED IS: 31.77 MILLIONS BTU

COMPONENT	STREAM NUMBER 10	MOLS	%	PROCESS GAS TEMPERATURE,	FEED#	544.1	EXIT#	605.9 DEGREES FAHRENHEIT.
				PROCESS GAS ENTHALPY,	FEED#	226.85	EXIT#	250.62 MILLIONS BTU.
H2	15505.4	25.0						
CH4	3513.2	5.0						
CO	12318.5	20.5						
CO2	9197.6	15.3						
H2O	18905.5	31.4						
N2	200.5	0.3						
C2H6	0.0	0.0						
H2S	477.9	0.8						
NH3	0.0	0.0						
CUS	13.7	0.0						
TOTAL	60132.3	100.0						

AIR4.160 JPL4. 2-STAGE AIR COMPRESSOR, WITH AIR FEED @ 160.0% OF STOICHIOMETRIC

19 SHIFT CONVERSION, FIRST STAGE, HIGH TEMPERATURE & PRESS.

GAS COMPOSITION IS MODIFIED BY WATER-GAS SHIFT REACTION. HYDROCARBONS ARE NOT AFFECTED.

EQUILIBRIUM TEMPERATURE = 870.0 DEGREES F PRESSURE = 464.0 PSIA

COMPONENT	FEED GAS # LB MOLES	PRODUCT GAS MOLES	%	PRODUCT GAS STREAM # LB PSIA
H2	15505.4	22915.6	38.1	176.8
CH4	3513.2	3513.2	5.8	27.1
CO	12318.5	4908.2	8.2	37.9
CO2	9197.6	16607.8	27.6	128.2
H2O	18905.5	11495.3	19.1	88.7
N2	200.5	200.5	0.3	1.5
C2H6	0.0	0.0	0.0	0.0
H2S	477.9	477.9	0.8	3.7
NH3	0.0	0.0	0.0	0.0
CUS	13.7	13.7	0.0	0.1
TOTAL	60132.3	60132.3	100.0	464.0

THE HEAT OF REACTION IS APPROXIMATELY 124.6 MILLIONS BTU

GAS ENTERS REACTOR AT 605.9 AND EXITS AT 821.6 DEGREES F.

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CASE AIR4.160
STEP 20
14 JAN 83 9:21:14

AIR4.160 JPL4, 2-STAGE AIR COMPRESSOR, WITH AIR FEED @ 160.0% OF STOICHIOMETRIC

20 EXPAND SHIFT (1) EXIT GAS THROUGH POWER RECOVERY TURBINE.

THIS STEP DESCRIBES A PROCESS GAS TURBINE-EXPANDER

COMPONENT	STREAM NUMBER 20	MOLS	%
H2	22915.6	38.1	
CH4	3513.2	5.8	
CO	4908.2	8.2	
CO2	16607.8	27.6	
H2O	11495.3	19.1	
N2	200.5	0.3	
COH6	0.0	0.0	
H2S	477.9	0.8	
NH3	0.0	0.0	
CS	13.7	0.0	
TOTAL	60132.3	100.0	

THE PRESSURE LET-DOWN RATIO IS 0.422

SHAFT ENERGY DELIVERED, HORSEPOWER: 42599.
KILOWATTS: 31756.

ISENTROPIC GAS EXPANSION EFFICIENCY IS 92.0 % (SET BY USER)
MECHANICAL EFFICIENCY ASSUMED IS 98.0%

INLET TEMPERATURE IS: 821.6 DISCHARGE IS: 618.8 DEGREES F.

INLET PRESSURE IS: 464.0 DISCHARGE IS: 196.0 PSIA

INLET VOLUME IS: 28627.0 DISCHARGE IS: 57045.0 ACFM

AIR4.160
 CASE AIR4.160
 STEP 21
 14 JAN 83 9121114

AIR4.160 JPL4, 2-STAGE AIR COMPRESSOR, WITH AIR FEED # 160.0% OF STOICHIOMETRIC

 21 COOL INVER-STAGE SHIFT GAS IN WHBI RAISE STEAM.

THE DUTY OF THIS PROCESS COOLER IS: -10.06 MILLIONS BTU PER HOUR.

COMPONENT	STREAM NUMBER 21	MOLS	%
H2	22415.6	38.1	
CH4	3513.2	5.8	
CO	4908.2	8.2	
CO2	16607.8	27.6	
H2O	11495.3	19.1	
NC	200.5	0.3	
C2H6	0.0	0.0	
H2S	477.4	0.8	
NH3	0.0	0.0	
CUS	13.7	0.0	
TOTAL	60132.3	100.0	

PROCESS GAS TEMPERATURE, FEED= 618.8 EXIT= 600.0 DEGREES FAHRENHEIT.
 PROCESS GAS ENTHALPY, FEED= 272.64 EXIT= 262.58 MILLIONS BTU.

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CASE AIR4.160
 STEP 22
 14 JAN 83 9121114

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AIR4.160 JPL4. 2-STAGE AIR COMPRESSOR, WITH AIR FEED # 160.08 OF STOICHIOMETRIC

 22 SECOND STAGE SHIFT, MODERATE TEMP., LOWER PRES.

GAS COMPOSITION IS MODIFIED BY WATER-GAS SHIFT REACTION. HYDROCARBONS ARE NOT AFFECTED.

EQUILIBRIUM TEMPERATURE = 695.0 DEGREES F PRESSURE = 189.0 PSIA

COMPONENT	FEED GAS # 21 MOLS	PRODUCT GAS MOLS	PRODUCT GAS STREAM %	PSIA
H2	22915.6	24963.3	41.5	78.5
CH4	3513.2	3513.2	5.8	11.0
CO	4908.2	2860.6	4.8	9.0
CO2	16607.4	16555.5	31.0	58.6
H2O	11495.3	9447.7	15.7	29.7
N2	200.5	200.5	0.3	0.6
C2H6	0.0	0.0	0.0	0.0
H2S	477.9	477.9	0.8	1.5
NH3	0.0	0.0	0.0	0.0
CUS	13.7	13.7	0.0	0.0
TOTAL	60132.3	60132.3	100.0	189.0

THE HEAT OF REACTION IS APPROXIMATELY 34.5 MILLIONS BTU

GAS ENTERS REACTOR AT 600.0 AND EXITS AT 660.1 DEGREES F.

CASE AIR4.160
 STEP 23
 14 JAN 83 9:21:14

AIR4.160 JPL4, 2-STAGE AIR COMPRESSOR, WITH AIR FEED @ 160.08 OF STOICHIOMETRIC

23 H23T35 COOL SHIFTED GAS TO WARM SWEET FUEL #35.

THE DUTY OF THIS PROCESS COOLER IS: -27.13 MILLIONS BTU PER HOUR.

COMPONENT	STREAM NUMBER	MOLS	%
H2	24963.3	41.5	
CH4	3513.2	5.8	
CU	2860.6	4.8	
CU2	18655.5	31.0	
H2O	9447.7	15.7	
N2	200.5	0.3	
CO2	0.0	0.0	
H2S	477.9	0.8	
NH3	0.0	0.0	
CUS	13.7	0.0	
TOTAL	60132.3	100.0	

PROCESS GAS TEMPERATURE, FEED= 660.1 EXIT= 610.0 DEGREES FAHRENHEIT.
 PROCESS GAS ENTHALPY, FEED= 297.04 EXIT= 269.91 MILLIONS BTU.

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CASE: AIR4.160
STEP 24
14 JAN 83 9:21:14

AIR4.160 JPL4, 2-STAGE AIR COMPRESSOR, WITH AIR FEED @ 160.0% OF STOICHIOMETRIC

24 R24T34 COOL SHIFTED GAS TO WARM SWEET FUEL #34.
THE DUTY OF THIS PROCESS COOLER IS 48.08 MILLIONS BTU PER HOUR.

COMPONENT	STREAM NUMBER 24	MOLS	%
H2	24963.3	41.5	
CH4	3513.2	5.8	
CO	2860.6	4.8	
CO2	18655.5	31.0	
H2O	9447.7	15.7	
N2	200.5	0.3	
C2H6	0.0	0.0	
H2S	477.9	0.8	
NF3	0.0	0.0	
CUS	13.7	0.0	
TOTAL	60132.3	100.0	

PROCESS GAS TEMPERATURE, FEED= 610.0 EXIT= 520.0 DEGREES FAHRENHEIT.
PROCESS GAS ENTHALPY, FEED= 269.91 EXIT= 221.83 MILLIONS BTU.

AIR4.160 JPL4, 2-STAGE AIR COMPRESSOR, WITH AIR FEED @ 160.0% OF STOICHIOMETRIC
CASE AIR4.160
STEP 25
14 JAN 83 9121114

25 R25TJ3 COOL SHIFTED GAS TO WARM SWEET FUEL #33.
THE DUTY OF THIS PROCESS COOLER IS: -47.17 MILLIONS BTU PER HOUR.

COMPONENT	STREAM NUMBER	MOLS	%
H2	24963.3	41.5	
CH4	3513.2	5.8	
CO	2860.6	4.8	
CO2	18655.5	31.0	
H2O	9447.7	15.7	
N2	200.5	0.3	
C2H6	0.0	0.0	
H2S	477.9	0.8	
NH3	0.0	0.0	
CUS	13.7	0.0	
TOTAL	60132.3	100.0	

PROCESS GAS TEMPERATURE, FEED= 520.0 EXIT= 430.0 DEGREES FARENHEIT.
PROCESS GAS ENTHALPY, FEED= 221.83 EXIT= 174.66 MILLIONS BTU.

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CASE AIR4.160
 STEP 26
 14 JAN 83 9:21:14

AIR4.160 JPL4, 2-STAGE AIR COMPRESSOR, WITH AIR FEED @ 160.0% OF STOICHIOMETRIC

26 DESTROY CARBONYL SULFIDE BY HYDROLYSIS.
 THIS STEP HEATS CUS WITH WATER VAPOR TO PRODUCE H2S AND CARBON DIOXIDE.

COMPONENT	FEED STREAM 25 MOLS	%	EXIT STREAM 26 MOLS	%
H2	24953.3	41.5	24963.3	41.5
CH4	3513.2	5.8	3513.2	5.8
CU	2860.6	4.8	2860.6	4.8
CU2	18655.5	31.0	18668.8	31.0
H2O	9447.7	15.7	9434.4	15.7
NC	200.5	0.3	200.5	0.3
C2H6	0.0	0.0	0.0	0.0
H2S	477.9	0.8	491.2	0.8
NH3	0.0	0.0	0.0	0.0
CUS	13.7	0.0	0.5	0.0
TOTAL	60132.3	100.0	60132.3	100.0

EQUILIBRIUM TEMPERATURE = 430.0 DEGREES F PRESSURE = 179.0 PSIA

RESIDUAL CARBONYL SULFIDE IS 0.4572E+00 MOLS.
 MOL FRACTION IS 9.02 PPM ON A WATER-FREE BASIS.

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CASE AIR4.160
 STEP 27
 14 JAN 83 9121:14

AIR4.160 JPL4, 2-STAGE AIR COMPRESSOR, WITH AIR FEED @ 160.0% OF STOICHIOMETRIC

 27 R27132 COOL SHIFTED GAS TO WARM SWEET FUEL #32.

THE DUTY OF THIS PROCESS COOLER IS: -56.31 MILLIONS BTU PER HOUR.

COMPONENT	STREAM NUMBER 27	MOLS	%
H2	24963.3	41.5	
CH4	3513.2	5.8	
CO	2860.6	4.8	
CU2	18668.8	31.0	
H2O	9434.4	15.7	
N2	200.5	0.3	
C2H6	0.0	0.0	
H2S	491.2	0.8	
NH3	0.0	0.0	
CU5	0.5	0.0	
TOTAL	60132.3	100.0	

PROCESS GAS TEMPERATURE, FEED= 430.0 EXIT= 320.0 DEGREES FAHRENHEIT.
 PROCESS GAS ENTHALPY, FEED= 174.66 EXIT= 118.35 MILLIONS BTU.

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CASE AIR4.160
 STEP 28
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AIR4.160 JPL4, 2-STAGE AIR COMPRESSOR, WITH AIR FEED @ 160.0% OF STOICHIOMETRIC

28 COOL TO AMBIENT DRUP CONDENSATE DUMP HEAT TO AIR & CW.

THIS STEP COOLS THE GAS TO 100.0 F AND SUBTRACTS CONDENSATE.

COMPONENT	FEED STREAM 27 MOLS	%	EXIT STREAM 28 MOLS	%
H2	24963.3	41.5	24963.3	49.0
CH4	3513.2	5.8	3513.2	6.9
CO	2860.6	4.8	2860.6	5.6
CU2	18668.8	31.0	18668.8	36.6
H2O	9434.4	15.7	280.2	0.5
IN2	200.5	0.3	200.5	0.4
C2H6	0.0	0.0	0.0	0.0
H2S	491.2	0.8	491.2	1.0
NH3	0.0	0.0	0.0	0.0
CUS	0.5	0.0	0.5	0.0
TOTAL	60132.3	100.0	50978.1	100.0

TEMPERATURE 320.0 F 100.0 F
 PRESSURE 177.0 PSIA 173.0 PSIA

GAS DUMP POINT 244.8 F 9154.2 MOLS 100.0 F
 WATER SUBTRACTED
 HEAT EXCHANGE DUTIES:
 GAS COOLING 107.551 MILLIONS BTU
 WATER CONDENSING 171.093 MILLIONS BTU
 TOTAL 278.644 MILLIONS BTU

AIR4.160 JPL4 2-STAGE AIR COMPRESSOR, WITH AIR FEED @ 160.0% OF STOICHIOMETRIC

CASE AIR4.160
STEP 29
14 JAN 83 9121114

29 REMOVE SULFUR FROM RAW GAS BY SELEXOL SYSTEM.

THIS STEP REMOVES ALL H2S BUT ONLY A FRACTION OF THE CO2.
IT USES THE SELEXOL PROCESS.

COMPONENT	FEED GAS STREAM # 28		EXIT GAS STREAM # 29		EXIT GAS STREAM # 94	
	MOLS	% PSIA	MOLS	% PSIA	MOLS	% PSIA
H2	24953.3	49.0	24962.9	51.9	0.4	0.0
CH4	3513.2	6.9	3510.9	7.3	2.2	0.0
CO	2860.6	5.6	2860.4	5.9	0.2	0.0
CO2	18668.8	36.6	16574.9	34.5	2093.9	74.5
H2O	280.2	0.5	0.3	0.0	222.8	11.0
N2	200.5	0.4	200.5	0.4	0.0	1.2
C2H6	0.0	0.0	0.0	0.0	0.0	0.0
H2S	491.2	1.0	0.0	0.0	0.0	0.0
NH3	0.0	0.0	0.0	0.0	491.1	17.5
CUS	0.5	0.0	0.3	0.0	0.0	0.0
TOTAL	50978.1	100.0	48110.2	100.0	2810.9	100.0

LOW-PRESSURE STEAM USAGE IS 214. THOUSANDS POUNDS PER HOUR.

POWER CONSUMPTION IS 6958. KILOWATTS.

DRY ACID GAS SENT TO CLAUSS PLANT IS 19.00 % H2S.

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CASE AIR4.160
 STEP 30
 14 JAN 83 9121:14

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AIR4.160 JPL4, 2-STAGE AIR COMPRESSOR, WITH AIR FEED M 160.0% OF STOICHIOMETRIC

SPLIT RECYCLE GAS #30 FOR COAL FEED; GAS #31 GOES FORWARD.
 THIS STEP SPLITS STREAM 29 INTO SIDESTREAM 30 AND DOWNSTREAM, 31 SIDESTREAM 30 IS RECYCLED.

COMPONENT	FEED STREAM 29		EXIT STREAM 30		EXIT STREAM 31	
	MOLS	%	MOLS	%	MOLS	%
H2	24962.9	51.4	1872.2	51.9	23090.6	51.9
CH4	3510.9	7.3	263.3	7.3	3247.6	7.3
CO	2360.4	5.4	214.5	5.9	2645.9	5.9
CO2	10574.9	34.5	1243.1	34.5	15331.7	34.5
H2O	0.3	0.0	0.0	0.0	0.3	0.0
N2	200.5	0.4	15.0	0.4	185.5	0.4
C2H6	0.0	0.0	0.0	0.0	0.0	0.0
H2S	0.0	0.0	0.0	0.0	0.0	0.0
NH3	0.0	0.0	0.0	0.0	0.0	0.0
CO2	0.3	0.0	0.0	0.0	0.3	0.0
TOTAL	48110.2	100.0	3608.3	100.0	44502.0	100.0

THE TEMPERATURE OF ALL THREE GAS STREAMS IS 100.0 DEGREES FARENHEIT.
 THE PRESSURE OF THE TWO EXIT GAS STREAMS IS 168.0 PSIA.

AIR4.16U JPL4, 2-STAGE AIR COMPRESSOR, WITH AIR FEED # 160.0% OF STOICHIOMETRIC

CASE: AIR4.160
STEP 32
14 JAN 83 9:21:14

32 K27132 WARM SWEETENED GAS COOL SHIFT EXIT #27.

THIS STEP RECUPERATES HEAT AGAINST STEP # 27
THE ENTHALPY TRANSFERRED IS: 56.31 MILLIONS BTU

COMPONENT	STREAM NUMBER 32	MOLS	%
H2	23090.6	51.9	
CH4	3247.6	7.3	
CU	2665.9	5.9	
CU2	15331.7	34.5	
H2O	0.3	0.0	
N2	185.5	0.4	
C2H6	0.0	0.0	
H2S	0.0	0.0	
NH3	0.0	0.0	
CUS	0.3	0.0	
TOTAL	44502.0	100.0	

PROCESS GAS TEMPERATURE,	FEED=	100.0	EXIT=	258.4	DEGREES FARENHEIT.
PROCESS GAS ENTHALPY,	FEED=	7.92	EXIT=	64.22	MILLIONS BTU.

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CASE AIR4.160
 STEP 33
 14 JAN 83 9121:14

AIR4.160 JPL4. 2-STAGE AIR COMPRESSOR, WITH AIR FEED # 160.08 OF STOICHIOMETRIC

JJ H25133 WARM SWEETENED GAS COOL SHIFT EXIT #25.

THIS STEP RECOVERS HEAT AGAINST STEP # 25
 THE ENTHALPY TRANSFERRED IS 47.17 MILLIONS BTU

COMPONENT	STREAM NUMBER 33	SOLES	8
H2	23090.6	51.9	
CH4	3247.6	7.3	
CO	2645.9	5.9	
CO2	15331.7	34.5	
H2O	0.3	0.0	
N2	185.5	0.4	
C2H6	0.0	0.0	
H2S	0.0	0.0	
NH3	0.0	0.0	
CUS	0.3	0.0	
TOTAL	44502.0	100.0	

PROCESS GAS TEMPERATURE, FEED= 258.4 EXIT= 385.4 DEGREES FAHRENHEIT.
 PROCESS GAS ENTHALPY, FEED= 64.22 EXIT= 111.39 MILLIONS BTU.

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CASE: AIR4.160
 STFP 34
 14 JAN 83 9:21:14

AIR4.160 JPL4: 2-STAGE AIR COMPRESSOR, WITH AIR FEED @ 160.0% OF STOICHIOMETRIC

J4 H24T34 WARM SWEETENED GAS: COOL SHIFT EXIT #24.

THIS STEP RECOVERATES HEAT AGAINST STEP # 24
 THE ENTHALPY TRANSFERRED IS: 48.08 MILLIONS BTU

COMPONENT	STREAM NUMBER J4	PULS	8
H2	23090.6	51.9	
CH4	3247.6	7.3	
CO	2645.9	5.9	
CO2	15331.7	34.5	
H2O	0.3	0.0	
N2	185.5	0.4	
C2H6	0.0	0.0	
H2S	0.0	0.0	
NH3	0.0	0.0	
CUS	0.3	0.0	
TOTAL	44502.0	100.0	

PROCESS GAS TEMPERATURE: FEED= 385.4 EXIT= 510.9 DEGREES FAHRENHEIT.
 PRUCSS GAS ENTHALPY: FEED= 111.39 EXIT= 159.47 MILLIONS BTU.

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CAS# AIR4.160
 STEP 35
 14 JAN 83 9:21:14

AIR4.160 JPL4, 2-STAGE AIR COMPRESSOR, WITH AIR FEED @ 160.0% OF STOICHIOMETRIC

35 R23T35 WARM SWEETENED GAS COOL SHIFT EXIT #23.

THIS STEP RECUPERATES HEAT AGAINST STEP # 23
 THE ENTHALPY TRANSFERRED IS: 27.13 MILLIONS BTU

COMPONENT	STREAM NUMBER	%
H2	23090.6	51.9
CH4	3247.6	7.3
CO	2645.9	5.9
CO2	15331.7	34.5
H2O	0.3	0.0
N2	185.5	0.4
C2H6	0.0	0.0
H2S	0.0	0.0
NH3	0.0	0.0
CUS	0.3	0.0
TOTAL	44502.0	100.0

PROCESS GAS TEMPERATURE, FEED= 510.9 EXIT= 580.3 DEGREES FARENHEIT.
 PROCESS GAS ENTHALPY, FEED= 159.47 EXIT= 186.60 MILLIONS BTU.

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CASE: AIR4.160
 STEP: 36
 14 JAN 83 9:21:14

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AIR4.160 JPL4, 2-STAGE AIR COMPRESSOR, WITH AIR FEED @ 160.0% OF STOICHIOMETRIC

 36 RJ6F55 WARM SWEETENED GAS COOL FUEL CELL ANODE EXIT #55.

THIS STEP RECUPERATES HEAT AGAINST STEP # 55
 THE ENTHALPY TRANSFERRED IS: 68.53 MILLIONS BTU

COMPONENT	STREAM NUMBER 36	MOLS	%
H2	23090.6		51.9
CH4	3247.6		7.3
CO	2645.9		5.9
CO2	15331.7		34.5
H2O	0.3		0.0
N2	185.5		0.4
CCM6	0.0		0.0
HCS	0.0		0.0
NH3	0.0		0.0
CVS	0.3		0.0
TOTAL	44506.0		100.0

PROCESS GAS TEMPERATURE, FEED= 580.3 EXIT= 751.3 DEGREES FAHRENHEIT.
 PROCESS GAS ENTHALPY, FEED= 186.60 EXIT= 255.14 MILLIONS BTU.

CASE A1M4.160
 STEP 37
 14 JAN 83 9:21:14

A1M4.160 JPL4 2-STAGE AIR COMPRESSOR WITH AIR FEED @ 100.0% OF STOICHIOMETRIC

J7 REMOVE LAST TRACES OF SULFUR BY ZINC OXIDE METATHESIS.
 THIS STEP SIMULATES THE REMOVAL OF THE LAST TRACES OF SULFUR FROM THE GAS BY A BED OF ZINC OXIDE.

COMPONENT	FEED STREAM 36		EXIT STREAM 37	
	MOLS	%	MOLS	%
H2	23090.6	51.9	23090.6	51.9
CH4	3247.6	7.3	3247.6	7.3
CO	2645.9	5.9	2645.9	5.9
CO2	15331.7	34.5	15331.7	34.5
H2O	0.3	0.0	0.3	0.0
N2	185.5	0.4	185.5	0.4
C2H6	0.0	0.0	0.0	0.0
H2S	0.0	0.0	0.0	0.0
NH3	0.0	0.0	0.0	0.0
CUS	0.3	0.0	0.0	0.0
TOTAL	44502.0	100.0	44501.7	100.0

ZINC OXIDE CONSUMPTION WILL BE ABOUT 0.1245E+03 POUNDS PER HOUR AT 20% LOADING.

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AIR4.160 JPL4, 2-STAGE AIR COMPRESSOR, WITH AIR FEED @ 160.0% OF STOICHIOMETRIC
CASE AIR4.160
STEP 38
14 JAN 83 9:21:14

J8 TEST GAS FOR DANGER OF CARBON DEPOSITION.

THIS IS A TEST WHETHER THE GAS IS STABLE WITH RESPECT TO CARBON DEPOSITION.

GAS COMPOSITION FROM STREAM # J7
CARBON STABILITY REPORTED AS STREAM # 38

TEMPERATURE: 1200.0 DEGREES F
PRESSURE: 150.0 PSIA

	GAS STREAM MOLS	GAS STREAM MOL FRACTION	GRAPHITE LIMIT		AMORPHOUS LIMIT	
			MOL FRACTION	MOL FRACTION	MOL FRACTION	MOL FRACTION
H2	23090.6	0.5189	0.2263	0.2240		
CH4	3247.6	0.0730	0.1261	0.1331		
CO	2645.9	0.0595	0.0860	0.1005		
CO2	15331.7	0.3445	0.2468	0.2439		
H2O	0.3	0.0000	0.3148	0.2984		
N2	185.5	0.0042				
C2H6	0.0	0.0000				
H2S	0.0	0.0000				
NH3	0.0	0.0000				
CO5	0.0	0.0000				
TOTAL	44501.7	1.0000	1.0000	1.0000		

GAS HUMIDIFICATION REQUIREMENTS TO ACHIEVE STABILITY ARE CALCULATED BELOW

GAS IS UNSTABLE. 6537.76 MOLS STEAM NEEDED.
STEAM INJECTION CAN DILUTE CARBON CONTENT TO HALFWAY BETWEEN GRAPHITE AND AMORPHOUS CARBON LIMIT.

AT 1100.0 DEGREES F. AND 150.0 PSIA GAS NEEDS 7991.03 MOLS WATER. 0.1598 = AMORPHOUS CARBON LIMIT.
AT 1000.0 DEGREES F. AND 150.0 PSIA GAS NEEDS 8014.76 MOLS WATER. 0.1644 = AMORPHOUS CARBON LIMIT.
AT 900.0 DEGREES F. AND 150.0 PSIA GAS NEEDS 6759.28 MOLS WATER. 0.1767 = AMORPHOUS CARBON LIMIT.

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CASE AIR4.160
 STEP 42
 14 JAN 83 9:21:14

AIR4.160 JPL4, 2-STAGE AIR COMPRESSOR, WITH AIR FEED @ 160.0% OF STOICHIOMETRIC

 42 ADD STEAM VIA TURBINE; PREVENT CARBON DROP IN FUEL CELL.

THIS IS A POWER RECOVERY TURBINE WHICH EXPANDS 8015. MOLS OF HIGH PRESSURE SUPERHEATED STEAM INTO PROCESS STREAM NUMBER 42

COMPONENT	FEED STREAM 37 MOLS	%	EXIT STREAM 42 MOLS	%
H2	23090.6	51.9	23090.6	44.0
CH4	3247.6	7.3	3247.6	6.2
CU	2645.9	5.9	2645.9	5.0
CO2	15331.7	34.5	15331.7	29.2
H2O	0.3	0.0	8015.1	15.3
N2	185.5	0.4	185.5	0.4
C2H6	0.0	0.0	0.0	0.0
H2S	0.0	0.0	0.0	0.0
NH3	0.0	0.0	0.0	0.0
CUS	0.0	0.0	0.0	0.0
TOTAL	44501.7	100.0	52516.4	100.0

THE STEAM PRESSURE LET-DOWN RATIO IS 0.102
 FROM 1500.0 PSIA TO 153.0 PSIA

INLET STEAM TEMPERATURE AT THROTTLE IS: 1000.0 DEGREES F.
 DRY FEED GAS TEMPERATURE IS: 751.3 DEGREES F.
 MOIST EXIT GAS TEMPERATURE IS: 703.1 DEGREES F.

ISENTROPIC EXPANSION EFFICIENCY IS 92.0 %.
 MECHANICAL EFFICIENCY ASSUMED IS 98.0 %.

SHAFT ENERGY DELIVERED, HORSEPOWER: 14310.
 KILOWATTS: 10668.

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AIR4.160 JPL4, 2-STAGE AIR COMPRESSOR, WITH AIR FEED @ 160.0% OF STOICHIOMETRIC

CASE AIR4.160
STEP 43
14 JAN 83 9:21:14

43 FEED AIR AT 160.0% OF STOICHIOMETRIC. (INCLUDING SCALEX).

THIS INTRODUCES A SECOND UR SURSEQUENT FEED GAS AS STREAM NUMBER 43

COMPONENT	STREAM NUMBER 43	MOLS	%
CU2	0.0	0.0	
H2O	3755.8	2.5	
N2	117312.1	77.1	
O2	30989.3	20.4	
TOTAL	152057.2	100.0	

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CASE AIR4.160
 STEP 44
 14 JAN 83 9:21:14

AIR4.160 JPL4, 2-STAGE AIR COMPRESSOR, WITH AIR FEED @ 160.0% OF STOICHIOMETRIC

44 COMPRESS COMBUSTION AIR TO FUEL CELL PRESSURE.

THIS STEP DESCRIBES A PROCESS GAS COMPRESSOR OF 2 STAGES.

COMPONENT	STREAM NUMBER	MOLS	%
CO2	0.0	0.0	0.0
H2O	3755.8	2.5	
N2	117312.1	77.1	
O2	30989.3	20.4	
TOTAL	152057.2	100.0	

THE COMPRESSION RATIO (PER STAGE) IS 3.60

ENERGY REQUIREMENTS
 PER STAGE TOTAL
 HORSEPOWER 116715. 233431.
 KILOWATTS 87008. 174015.

ISENTROPIC GAS COMPRESSION EFFICIENCY IS 88.0 %. (BY DEFAULT)
 MECHANICAL EFFICIENCY ASSUMED IS 98.0%

AT EACH STAGE SUCTION TEMPERATURE IS: 90.0 DISCHARGE IS: 360.7 DEGREES F.

INTERSTAGE COOLING DUTY IS: 296.94 MILLIONS BTU

PRESSURE (PSIA): SUCTION DISCHARGE VOLUME (ACFM): SUCTION DISCHARGE

STAGE # 1	14.3	51.5	1007706.2	417301.4
STAGE # 2	45.5	164.0	316664.3	131133.9

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AIR4.160 JPL4, 2-STAGE AIR COMPRESSOR, WITH AIR FEED @ 160.0% OF STOICHIOMETRIC
 CASE AIR4.160
 STEP 45
 14 JAN 83 9:21:14

 45 R45F65 - WARM COMBUSTION AIR FROM STACK GAS #65.

THIS STEP RECUPERATES HEAT AGAINST STEP # 65
 THE ENTHALPY TRANSFERRED IS: 156.12 MILLIONS BTU

COMPONENT	STREAM NUMBER 45	MOLS	%
CU2		0.0	0.0
H2O		3755.8	2.5
N2		117312.1	77.1
O2		30984.3	20.4
TOTAL		152057.2	100.0

PROCESS GAS TEMPERATURE, FEED= 360.7 EXIT= 504.4 DEGREES FAHRENHEIT.
 PROCESS GAS ENTHALPY, FEED= 304.74 EXIT= 462.86 MILLIONS BTU.

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CASE AIR4.160
STEP 46
14 JAN 83 9121114

AIR4.160 JPL4, 2-STAGE AIR COMPRESSOR, WITH AIR FEED @ 160.08 OF STOICHIOMETRIC

46 H46F62 - WARM COMBUSTION AIR FROM CATHODE RECYCLE #62.
THE DUTY OF THIS PROCESS HEATER IS: 219.06 MILLIONS BTU PER HOUR.

COMPONENT	STREAM NUMBER 46	MOLS	%
CO2	0.0	0.0	0.0
H2O	3755.8	2.5	
NC	117312.1	77.1	
O2	30989.3	20.4	
TOTAL	152057.2	100.0	

PROCESS GAS TEMPERATURE, FEED= 504.4 EXIT= 700.0 DEGREES FAHRENHEIT.
PROCESS GAS ENTHALPY, FEED= 462.86 EXIT= 681.91 MILLIONS BTU.

AIR4.160 JPL4, 2-STAGE AIR COMPRESSOR, WITH AIR FEED @ 160.0% OF STOICHIOMETRIC
CASE AIR4.160
STEP 47
14 JAN 83 9:21:14

47 R47f61 - WARM COMBUSTION AIR FROM CATHODE RECYCLE #61.
THE DUTY OF THIS PROCESS HEATER IS: 240.09 MILLIONS BTU PER HOUR.

COMPONENT	STREAM NUMBER 47	MOLS	%
CU2		0.0	0.0
H2O		3755.8	2.5
N2		117312.1	77.1
O2		30989.3	20.4
TOTAL		152057.2	100.0

PROCESS GAS TEMPERATURE, FEED= 700.0 EXIT= 910.0 DEGREES FAHRENHEIT.
PROCESS GAS ENTHALPY, FEED= 681.91 EXIT= 922.01 MILLIONS BTU.

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CASE AIR4.160
 STEP 48
 14 JAN 63 9:21:14

AIR4.160 JPL4+ 2-STAGE AIR COMPRESSOR, WITH AIR FEED W 160.0% OF STOICHIOMETRIC

 48 MERGE FRESH AIR #47 WITH RECYCLED CATHODE EXIT GAS #63.

THIS STEP ADUS STREAM # 47 & 63 TO PRODUCE STREAM # 48

COMPONENT	FEED STREAM 47 MOLS	%	FEED STREAM 63 MOLS	%	EXIT STREAM 48 MOLS	%
CO2	0.0	0.0	13915.9	11.1	13915.9	5.0
H2O	3755.8	2.5	27114.7	21.6	30870.5	11.1
N2	117312.1	77.1	77034.9	61.3	194347.0	70.0
O2	30989.3	20.4	7622.2	6.1	38611.5	13.9
TOTAL	152057.2	100.0	125687.6	100.0	277744.8	100.0

TEMPERATURE 910.0 728.1 823.9
 PRESSURE 158.0 158.0 156.0

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CASE: AIR4.160
 STEP 49
 14 JAN 83 9121:14

AIR4.160 JPL4, 2-STAGE AIR COMPRESSOR, WITH AIR FEED @ 160.0% OF STOICHIOMETRIC

49 BURN ANODE EXHAUST GAS #57 WITH DILUTED AIR #48.

THIS IS A CATALYTIC BURNER FOR FUEL GAS # 57, WITH AIR/OXYGEN SUPPLY # 48. THE FLUE GAS IS # 49.

STREAM NUMBER 57

COMPONENT	MOLS	%
H2	3326.2	3.6
CH4	12.4	0.0
CO	2433.1	2.6
CO2	51697.6	56.3
H2O	34249.9	37.3
N2	185.5	0.2
C2H6	0.0	0.0
H2S	0.0	0.0
NH3	0.0	0.0
CUS	0.0	0.0
TOTAL	91904.7	100.0

TEMPERATURE 883.5 DEGREES F
 PRESSURE 156.0 PSIA

COMPONENT	FEED STREAM 48 MOLS	%	EXIT STREAM 49 MOLS	%
CO2	13915.9	5.0	68059.0	18.6
H2O	30870.5	11.1	68471.4	18.7
N2	194347.0	70.0	194532.5	53.0
O2	38611.5	13.9	35706.9	9.7
TOTAL	277744.8	100.0	366769.9	100.0

TEMPERATURE 823.9 F
 PRESSURE 156.0 PSIA

1046.4 F
 152.0 PSIA

THE MIXED FEED TEMPERATURE IS ONLY 842.3 F.
 THIS MAY NOT BE HOT ENOUGH TO LIGHT OFF THE CATALYST.

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CASE AIR4.160
 STEP 50
 14 JAN 83 9:21:14

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AIR4.160 JPL4, 2-STAGE AIR COMPRESSOR, WITH AIR FEED @ 160.0% OF STOICHIOMETRIC

50 JOIN THE TWO FUEL CELL FEEDS, ANODE #42, CATHODE #49.

THIS STEP ADDS STREAM # 42 & 49 TO PRODUCE STREAM # 50

COMPONENT	FEED STREAM 42 MOLS	FEED STREAM 42 %	FEED STREAM 49 MOLS	FEED STREAM 49 %	EXIT STREAM 50 MOLS	EXIT STREAM 50 %
H2	23090.6	44.0	0.0	0.0	23090.6	44.0
CH4	3247.6	6.2	0.0	0.0	3247.6	6.2
CO	2645.9	5.0	0.0	0.0	2645.9	5.0
CO2	15331.7	29.2	0.0	0.0	15331.7	29.2
H2O	8015.1	15.3	0.0	0.0	8015.1	15.3
N2	185.5	0.4	0.0	0.0	185.5	0.4
C2H6	0.0	0.0	0.0	0.0	0.0	0.0
H2S	0.0	0.0	0.0	0.0	0.0	0.0
NH3	0.0	0.0	0.0	0.0	0.0	0.0
COS	0.0	0.0	0.0	0.0	0.0	0.0
TOTAL	52516.4	100.0	0.0	0.0	52516.4	100.0

COMPONENT	FEED STREAM 42 MOLS	FEED STREAM 42 %	FEED STREAM 49 MOLS	FEED STREAM 49 %	EXIT STREAM 50 MOLS	EXIT STREAM 50 %
CO2	0.0	0.0	68059.0	18.6	68059.0	18.6
H2O	0.0	0.0	68471.4	18.7	68471.4	18.7
N2	0.0	0.0	194532.5	53.0	194532.5	53.0
O2	0.0	0.0	35706.9	9.7	35706.9	9.7
TOTAL	0.0	0.0	366769.9	100.0	366769.9	100.0

TEMPERATURE 703.1 1001.5
 PRESSURE 1500.0 152.0

CASE AIR4.160
 STEP 51
 14 JAN 83 9:21:14

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AIR4.160 JPL4, 2-STAGE AIR COMPRESSOR, WITH AIR FELD @ 160.0% OF STOICHIOMETRIC

 51 THE FUEL CELL, MAXIMUM CONVERSION, REFORM WITHIN ANODE.

ANODE
 IN : PRESSURE= 1500.0 PSIA TEMPERATURE= 703.1 DEG. F
 CATHODE
 IN : PRESSURE= 152.0 PSIA TEMPERATURE= 1046.4 DEG. F
 ANODE
 OUT : PRESSURE= 150.0 PSIA TEMPERATURE= 1259.6 DEG. F
 CATHODE
 OUT : PRESSURE= 150.0 PSIA TEMPERATURE= 1259.6 DEG. F

ANODE

COMPONENT	FEED STREAM S0 MOLS	FEED STREAM S0 %	EXIT STREAM S1 MOLS	EXIT STREAM S1 %
H2	23090.6	44.0	3326.2	3.6
CH4	3247.6	6.2	12.4	0.0
CO	2645.9	5.0	2433.1	2.6
CO2	15331.7	29.2	51697.6	56.3
H2O	8015.1	15.3	34249.9	37.3
N2	185.5	0.4	185.5	0.2
C2H6	0.0	0.0	0.0	0.0
H2S	0.0	0.0	0.0	0.0
NH3	0.0	0.0	0.0	0.0
CUS	0.0	0.0	0.0	0.0
TOTAL	52516.4	100.0	91904.7	100.0

CATHODE

COMPONENT	FEED STREAM S0 MOLS	FEED STREAM S0 %	EXIT STREAM S1 MOLS	EXIT STREAM S1 %
CO2	68059.0	18.6	35141.1	11.1
H2O	68471.4	18.7	68471.4	21.6
N2	194532.5	53.0	194532.5	61.3
U2	35706.9	9.7	19248.0	6.1
TOTAL	366769.9	100.0	317393.0	100.0

51 THE FUEL CELL, MAXIMUM CONVERSION, REFURM WITHIN ANODE.
 AVERAGE QUANTITIES (LOG MEAN):

ANODE		CATHODE	
MOL FRACTION	PARTIAL PRESSURE (ATM)	MOL FRACTION	PARTIAL PRESSURE (ATM)
H2	0.1616	CO2	0.1449
CH4	0.0101	H2O	0.2009
CO	0.0372	N2	0.5707
CO2	0.4125	O2	0.0776
H2O	0.2465		
N2	0.0027		

CHARACTERISTIC CELL TEMPERATURE= 922.0 DEG. K 1253.0 DEG. F (INPUT)
 EXIT GAS TEMPERATURE OF CELL= 955.2 DEG. K 1259.6 DEG. F (CALCULATED FOR ADIABATIC CELL)
 DELTA G=-47.065 KCAL/GMOL -84718. BTU/LB-MOL 1.021 VOLTS
 NERNST OPEN CIRCUIT POTENTIAL 0.958 VOLTS
 CURRENT DENSITY 150.0 MILLIAMPS/CM2 139.4 AMPS/FT2
 FUEL CONVERSION 85.0 % (ALL FUEL COMPONENTS)
 OXYGEN CONVERSION 46.1 %
 CO2 CONVERSION 48.4 %
 GROSS POWER OUTPUT= 6.646E+05 KW 664.63 MEGA-W
 HEAT EQUIVALENT TO NET POWER OUTPUT 0.2268E+10 BTU/HR.
 TOTAL HEAT RELEASE, INCLUDING ELECTROCHEMICAL AND GAS PHASE REACTIONS 0.3263E+10 BTU/HR. (ISOTHERMAL)
 FRACTION OF POTENTIAL ENTHALPY RELEASE EXTRACTED AS POWER: 70.0 %
 CELL AREA = 0.5338E+10 CM2
 TOTAL CURRENT = 0.8005E+09 AMPS

VOLTAGE SUMMARY (VOLTS)

DELIVERED VOLTAGE	0.8303
CATHODE GAS ACTIVITY	-0.0109
CATHODE POLARIZATION	0.0312
OHMIC RESISTANCE	0.0407
ANODE POLARIZATION	0.0559
ANODE GAS ACTIVITY	0.0739
TOTAL THERMODYNAMIC EQUILIBRIUM CELL	1.0212

CASE: AIR4.160
 STEP: 52
 14 JAN 83 9:21:14

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AIR4.160 JPL4. 2-STAGE AIR COMPRESSOR, WITH AIR FEED @ 160.0% OF STOICHIOMETRIC

SEPARATE ANODE EXIT GAS, #52, FROM CATHODE EXIT GAS, #53.

THIS STEP SEPARATES HYBRID STREAM # 51 INTO FUEL GAS COMPONENTS, STREAM # 52
 AND OXIDIZING GAS COMPONENTS, STREAM # 53.

COMPONENT	FEED GAS # 51 MOLS	PRODUCT GAS STREAM # 52 MOLS	PRODUCT GAS STREAM # 52 PSIA
H2	3326.2	3326.2	3.6
CH4	12.4	12.4	0.0
CO	2433.1	2433.1	2.6
CU2	51697.6	51697.6	56.3
H2O	34249.9	34249.9	37.3
N2	185.5	185.5	0.2
C2F6	0.0	0.0	0.0
H2S	0.0	0.0	0.0
NH3	0.0	0.0	0.0
COS	0.0	0.0	0.0
TOTAL	91904.7	91904.7	100.0

COMPONENT	FEED GAS # 51 MOLS	PRODUCT GAS STREAM # 53 MOLS	PRODUCT GAS STREAM # 53 PSIA
CU2	35141.1	35141.1	11.1
H2O	68471.4	68471.4	21.6
N2	194532.5	194532.5	61.3
O2	19248.0	19248.0	6.1
TOTAL	317393.0	317393.0	100.0

TEMPERATURE = 1259.6 DEGREES F.
 PRESSURE = 145.0 PSIA

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CASE AIR#160
STEP 54
14 JAN 83 9:21:14

AIR#160 JPL4, 2-STAGE AIR COMPRESSOR, WITH AIR FEED @ 160.0% OF STOICHIOMETRIC

54 COOL ANNU EXIT GAS IN #HB - MAISE STEAM.

THE DUTY OF THIS PROCESS COOLER IS: -338.74 MILLIONS BTU PER HOUR.

COMPONENT	STREAM NUMBER 54	MOLS	%
H2	3326.2	3.6	
CH4	12.4	0.0	
CU	2433.1	2.6	
CU2	51697.6	56.3	
H2O	34249.9	37.3	
N2	185.5	0.2	
C2H6	0.0	0.0	
H2S	0.0	0.0	
NH3	0.0	0.0	
CUS	0.0	0.0	
TOTAL	91904.7	100.0	

PROCESS GAS TEMPERATURE, FEED= 1259.6 EXIT= 925.0 DEGREES FAHRENHEIT.
PROCESS GAS ENTHALPY, FEED= 1095.98 EXIT= 757.24 MILLIONS BTU.

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CASE AIR4.160
STEP 55
14. JAN 83 9121114

AIR4.160 JPL4. 2-STAGE AIR COMPRESSOR, WITH AIR FEED @ 160.0% OF STOICHIOMETRIC

55 H36F55 COOL ANGE EXIT GAS TO WARM ANGE FEED GAS #36.

THE DUTY OF THIS PROCESS COOLER IS: -68.53 MILLIONS BTU PER HOUR.

COMPONENT	STREAM NUMBER 55	MOLS	%
H2	3326.2	3.6	
GM4	12.4	0.0	
CU	2433.1	2.6	
CO2	51697.6	56.3	
H2O	34249.9	37.3	
NC	185.5	0.2	
CH4	0.0	0.0	
H2S	0.0	0.0	
NP3	0.0	0.0	
CUS	0.0	0.0	
TOTAL	91904.7	100.0	

PROCESS GAS TEMPERATURE: FEED= 925.0 EXIT= 855.0 DEGREES FARENHEIT.
PROCESS GAS ENTHALPY: FEED= 757.24 EXIT= 688.70 MILLIONS BTU.

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CASE AIR4.160
STEP 57
14 JAN 83 9121114

AIR4.160 JPL4, 2-STAGE AIR COMPRESSOR, WITH AIR FEED @ 160.0% OF STOICHIOMETRIC *****

57 ANODE EXIT GAS ROASTER FAN, TO FEED CATALYTIC BURNER.

THIS STEP DESCRIBES A PROCESS GAS COMPRESSOR

COMPONENT	STREAM NUMBER 57	MULS	%
H2	3326.2	3.6	
CH4	12.4	0.0	
CU	2433.1	2.6	
CU2	51697.6	56.3	
H2O	34249.9	37.3	
N2	185.5	0.2	
C2H6	0.0	0.0	
H2S	0.0	0.0	
NH3	0.0	0.0	
CUS	0.0	0.0	
TOTAL	91904.7	100.0	

THE COMPRESSION RATIO (PER STAGE) IS 1.11

ENERGY REQUIREMENTS
PER STAGE TOTAL

HORSEPOWER
KILOWATTS

11167.
8325.

ISENTROPIC GAS COMPRESSION EFFICIENCY IS 88.0 % (BY DEFAULT)
MECHANICAL EFFICIENCY ASSUMED IS 98.0%

AT EACH STAGE SUCTION TEMPERATURE IS: 855.0 DISCHARGE IS: 883.5 DEGREES F.

PRESSURE (PSIA): SUCTION DISCHARGE VOLUME (ACFM): SUCTION DISCHARGE

STAGE # 1 141.0 156.0 147734.7 136428.8

CASE: AIR4.160
 STEP 58
 14 JAN 83 9:21:14

AIR4.160 JPL4, 2-STAGE AIR COMPRESSOR, WITH AIR FEED @ 160.0% OF STOICHIOMETRIC

SPLIT CATHODE EXIT. COOLED RECYCLE (#58), STACK EXIT (#59).
 THIS STEP SPLITS STREAM 53 INTO SIDESTREAM 58 AND DOWNSTREAM 59. SIDESTREAM 58 IS RECYCLED.

COMPONENT	FEED STREAM 53 MOLS	FEED STREAM 53 %	EXIT STREAM 58 MOLS	EXIT STREAM 58 %	EXIT STREAM 59 MOLS	EXIT STREAM 59 %
CO2	35141.1	11.1	13915.9	11.1	21225.2	11.1
H2O	68471.4	21.6	27114.7	21.6	41356.7	21.6
N2	194532.5	61.3	77034.9	61.3	117497.6	61.3
O2	19248.0	6.1	7622.2	6.1	11625.8	6.1
TOTAL	317393.0	100.0	125687.6	100.0	191705.4	100.0

THE TEMPERATURE OF ALL THREE GAS STREAMS IS 1259.6 DEGREES FAHRENHEIT.
 THE PRESSURE OF THE TWO EXIT GAS STREAMS IS 142.0 PSIA.

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AIR4.160 JPL4, 2-STAGE AIR COMPRESSOR, WITH AIR FEED @ 160.0% OF STOICHIOMETRIC

CASE AIR4.160
 STEP 50
 14 JAN 83 9121114

60 COOL CATHODE RECYCLE GAS IN WHB - RAISE STEAM.

THE DUTY OF THIS PROCESS COOLER IS: -157.28 MILLIONS BTU PER HOUR.

COMPONENT	STREAM NUMBER	MOLS	%
CU2	13915.9	11.1	
H2O	27114.7	21.6	
NC	77034.9	61.3	
OC	7622.2	6.1	
TOTAL	125687.6	100.0	

PROCESS GAS TEMPERATURE, FEED= 1259.6 EXIT= 1115.0 DEGREES FAHRENHEIT.
 PROCESS GAS ENTHALPY, FEED= 1201.63 EXIT= 1044.35 MILLIONS BTU.

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AIR4.160 JPL4, 2-STAGE AIR COMPRESSOR, WITH AIR FEED @ 150.0% OF STOICHIOMETRIC
 CASE AIR4.160
 STEP 61
 14 JAN 83 9:21:14

 61 N4761 COOL CATHODE RECYCLE TO WARM COMBUSTION AIR #47.

THIS STEP RECUPERATES HEAT AGAINST STEP # 47
 THE ENTHALPY TRANSFERRED IS: -240.09 MILLIONS BTU

COMPONENT	STREAM NUMBER 61	MOLS	%
CU2	13915.9	11.1	
H2O	27114.7	21.6	
N2	77034.9	61.3	
O2	7622.2	6.1	
TOTAL	125687.6	100.0	

PROCESS GAS TEMPERATURE, FEED= 1115.0 EXIT= 889.3 DEGREES FAHRENHEIT.
 PROCESS GAS ENTHALPY, FEED= 1044.35 EXIT= 804.25 MILLIONS BTU.

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CASE AIR4.160
 STEP 62
 14 JAN 83 9:21:14

AIR4.160 JPL4. 2-STAGE AIR COMPRESSOR, WITH AIR FEED # 160.0% OF STADIOMERIC

02 H46F62 COOL CATHODE RECYCLE TO WARM COMBUSTION AIR #46.

THIS STEP RECOVERATES HEAT AGAINST STEP # 46
 THE ENTHALPY TRANSFERRED IS: -219.86 MILLIONS BTU

COMPONENT	STREAM NUMBER 62	MOLS	%
CO2	13915.9	11.1	
H2O	27114.7	21.6	
NE	77034.9	61.3	
O2	7622.2	6.1	
TOTAL	125687.6	100.0	

PROCESS GAS TEMPERATURE, FEED# 889.3 EXIT# 677.6 DEGREES FAHRENHEIT.
 PROCESS GAS ENTHALPY, FEED# 804.25 EXIT# 585.20 MILLIONS BTU.

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CASE AIR4.160
 STEP 63
 14 JAN 83 9:21:14

AIR4.160 JPL4, 2-STAGE AIR COMPRESSOR, WITH AIR FEED @ 160.0% OF STOICHIOMETRIC

6J CATHODE RECYCLE GAS CIRCULATION FAN.

THIS STEP DESCRIBES A PROCESS GAS COMPRESSOR

COMPONENT	STREAM NUMBER 63	MOLS	%
CO2	13915.9	11.1	
H2O	27114.7	21.6	
N2	77034.9	61.3	
O2	7622.2	6.1	
TOTAL	125667.6	100.0	

THE COMPRESSION RATIO (PER STAGE) IS 1.17

ENERGY REQUIREMENTS
 PER STAGE TOTAL
 HORSEPOWER 20749.
 KILOWATTS 15468.

ISENTROPIC GAS COMPRESSION EFFICIENCY IS 88.0 %. (BY DEFAULT)
 MECHANICAL EFFICIENCY ASSUMED IS 98.0%

AT EACH STAGE SUCTION TEMPERATURE IS: 677.6 DISCHARGE IS: 728.1 DEGREES F.

PRESSURE (PSIA): SUCTION DISCHARGE VOLUME (ACFM): SUCTION DISCHARGE

STAGE # 1 135.0 158.0 182548.4 162904.9

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64 EXPAND STACK GAS THROUGH TURBINE TO RECOVER POWER.

THIS STEP DESCRIBES A PROCESS GAS TURBINE-EXPANDER

COMPONENT	STREAM NUMBER	64
	MOLS	%
CO2	21225.2	11.1
H2O	41356.7	21.6
N2	117497.6	61.3
O2	11625.8	6.1
TOTAL	191705.4	100.0

THE PRESSURE LET-DOWN RATIO IS 0.128

SHAFT ENERGY DELIVERED, HORSEPOWER: 382962.
KILOWATTS: 285486.

ISENTROPIC GAS EXPANSION EFFICIENCY IS 92.0 %. (SET BY USER)
MECHANICAL EFFICIENCY ASSUMED IS 98.0%

INLET TEMPERATURE IS: 1259.6 DISCHARGE IS: 642.9 DEGREEIS F.
INLET PRESSURE IS: 142.0 DISCHARGE IS: 17.0 PSIA
INLET VOLUME IS: 400170.4 DISCHARGE IS: 2143694.8 ACFM

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CASE AIR6.160
 STEP 65
 14 JAN 83 9:21:14

AIR4.160 JPL4, 2-STAGE AIR COMPRESSOR, WITH AIR FEED @ 160.0% OF STOICHIOMETRIC

 65 H45F65 COOL STACK GAS TO WARM COMBUSTION AIR #45.

THE DUTY OF THIS PROCESS COOLER IS: -158.12 MILLIONS BTU PER HOUR.

COMPONENT	STREAM NUMBER 65	MOLS	%
CO2	21225.2		11.1
H2O	41356.7		21.6
N2	117497.6		61.3
O2	11625.8		6.1
TOTAL	191705.4		100.0

PROCESS GAS TEMPERATURE, FEED= 642.9 EXIT= 540.0 DEGREES FARENHEIT.
 PROCESS GAS ENTHALPY, FEED= 838.61 EXIT= 680.49 MILLIONS BTU.

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CASE AIR4.160
STEP 66
14 JAN 83 9:21:14

AIR4.160 JPL4, 2-STAGE AIR COMPRESSOR, WITH AIR FEED @ 160.0% OF STOICHIOMETRIC

66 READ COMBUSTION FUEL VALUE OF RAW GAS FROM GASIFIER.
THIS ROUTINE GIVES COMBUSTION HEAT VALUES.

STEP IDENTIFICATION # 66 SOURCE STREAM # 2

COMPONENT	TOTAL FUEL GAS MOLS	%	UNRIED FUEL GAS MOLS	%	SWEETENED GAS MOLS	%
H2	13633.2	25.9	13633.2	36.2	13633.2	46.7
CH4	3249.8	6.2	3249.8	8.6	3249.8	11.1
CO	12103.9	23.0	12103.9	32.2	12103.9	41.5
CO2	7954.5	15.1	7954.5	21.1		
H2O	14906.7	28.3				
N2	185.5	0.4	185.5	0.5	185.5	0.6
CSH6	0.0	0.0	0.0	0.0	0.0	0.0
H2S	477.9	0.9	477.9	1.3		
NH3	74.4	0.1				
CUS	13.7	0.0	13.7	0.0		
TOTAL	52599.7	100.0	37618.6	100.0	29172.5	100.0

BTU/MOL, NET 78601. 109633.
GROSS XXXXXX 120101.

BTU/CU FT, NET 207. 289.
GROSS XXX 316.

TOTAL STREAM COMBUSTION HEAT RELEASE, MILLIONS BTU:
NET 4134.4 4014.4
GROSS XXXXXX 4399.1 4399.1

CASE AIR4.160
 STFM 67
 14 JAN 83 9:21:14

AIR4.160 JPL4. 2-STAGE AIR COMPRESSOR, WITH AIR FEED # 160.0% OF STOICHIOMETRIC

67 READ COMBUSTION FUEL VALUE OF SHIFTED SOUR GAS.
 THIS ROUTINE GIVES COMBUSTION HEAT VALUES.

STEP IDENTIFICATION # 67 SOURCE STREAM # 31

COMPONENT	TOTAL FUEL GAS MOLS	%	DRIED FUEL GAS MOLS	%	SWEETENED GAS MOLS	%
H2	23090.6	51.9	23090.6	51.9	23090.6	79.2
CH4	3247.6	7.3	3247.6	7.3	3247.6	11.1
CO	2645.9	5.9	2645.9	5.9	2645.9	9.1
CO2	15331.7	34.5	15331.7	34.5		
H2O	0.3	0.0				
N2	185.5	0.4	185.5	0.4	185.5	0.6
C2H6	0.0	0.0	0.0	0.0	0.0	0.0
H2S	0.0	0.0	0.0	0.0		
NH3	0.0	0.0				
CUS	0.3	0.0	0.3	0.0		
TOTAL	44502.0	100.0	44501.7	100.0	29169.6	100.0

BTU/MOL, NET 86431.
 GROSS XXXXX

BTU/CU FT, NET 227.
 GROSS XXX

TOTAL STREAM COMBUSTION HEAT RELEASE, MILLIONS BTU:
 NET 3846.4
 GROSS XXXXX

131659.
 151238.

347.
 398.

3846.3
 4411.5

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CASE AIR4.160
 STEP 68
 14 JAN 83 9121114

AIR4.160 JPL4, 2-STAGE AIR COMPRESSOR, WITH AIR FEED @ 160.0% OF STOICHIOMETRIC

68 READ COMBUSTION FUEL VALUE OF SWEET ANODE FEED GAS.
 THIS ROUTINE GIVES COMBUSTION HEAT VALUES.

STEP IDENTIFICATION # 68 SOURCE STREAM # 37

COMPONENT	TOTAL FUEL GAS MOLS	%	DRYED FUEL GAS MOLS	%	SWEETENED GAS MOLS	%
H2	23090.6	51.9	23090.6	51.9	23090.6	79.2
CH4	3247.6	7.3	3247.6	7.3	3247.6	11.1
CO	2645.9	5.9	2645.9	5.9	2645.9	9.1
CO2	15331.7	34.5	15331.7	34.5		
H2O	0.3	0.0				
N2	185.5	0.4	185.5	0.4	185.5	0.6
C2H6	0.0	0.0	0.0	0.0	0.0	0.0
H2S	0.0	0.0	0.0	0.0		
NH3	0.0	0.0				
CUS	0.0	0.0	0.0	0.0		
TOTAL	44501.7	100.0	44501.4	100.0	29169.6	100.0

BTU/MOL, NET GROSS 86430. 86431.
 XXXXXX 99133. 131859.
 151238.

BTU/CU FT, NET GROSS 227. 227.
 XXX 261. 347.
 398.

TOTAL STREAM COMBUSTION HEAT RELEASE, MILLIONS BTU:
 NET 3846.3 3846.3
 GROSS XXXXX 4411.5 4411.5

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69 HEAD COMBUSTION FUEL VALUE OF ANODE EXIT GAS.
THIS ROUTINE GIVES COMBUSTION HEAT VALUES.

STEP IDENTIFICATION # 69 SOURCE STREAM # 54

COMPONENT	TOTAL FUEL GAS MOLS	%	URIED FUEL GAS MOLS	%	SWEETENED GAS MOLS	%
H2	3326.2	3.6	3326.2	5.8	3326.2	55.8
CH4	12.4	0.0	12.4	0.0	12.4	0.2
CO	2433.1	2.6	2433.1	4.2	2433.1	40.8
CO2	51697.6	56.3	51697.6	89.7		
H2O	34249.9	37.3				
N2	185.5	0.2	185.5	0.3	185.5	3.1
C2H6	0.0	0.0	0.0	0.0	0.0	0.0
H2S	0.0	0.0	0.0	0.0		
NH3	0.0	0.0				
CO5	0.0	0.0	0.0	0.0		
TOTAL	94904.7	100.0	57654.8	100.0	5957.2	100.0

BTU/MOL, NET 7037.
GROSS XXXXXX

BTU/CO FT, NET 19.
GROSS XXX

TOTAL STREAM COMBUSTION HEAT RELEASE, MILLIONS BTU
NET 646.7
GROSS XXXXXX

108558.
119306.

286.
314.

646.7
710.7

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CASE: AIR4.160
 STEP: 70
 14 JAN 83 9:21:14

AIR4.160 JPL4: 2-STAGE AIR COMPRESSOR, WITH AIR FEED @ 160.0% OF STOICHIOMETRIC

70 COMPRESS RECYCLE GAS TO INJECT FEED CUAL TO GASIFIER.

THIS STEP DESCRIBES A PROCESS GAS COMPRESSOR OF 2 STAGES.

COMPONENT	STREAM NUMBER 70	MOLS	%
H2	1872.2	51.9	
CH4	263.3	7.3	
CO	214.5	5.9	
CO2	1243.1	34.5	
H2O	0.0	0.0	
N2	15.0	0.4	
C2H6	0.0	0.0	
H2S	0.0	0.0	
NH3	0.0	0.0	
CUS	0.0	0.0	
TOTAL	3608.3	100.0	

THE COMPRESSION RATIO (PER STAGE) IS 1.84

ENERGY REQUIREMENTS
 PER STAGE TOTAL
 HORSEPOWER 1206.
 KILOWATTS 899.

IS(ENTROPIC GAS COMPRESSION EFFICIENCY IS 88.0 %. (BY DEFAULT)
 MECHANICAL EFFICIENCY ASSUMED IS 98.0%

AT EACH STAGE SUCTION TEMPERATURE IS: 100.0 DISCHARGE IS: 205.2 DEGREES F.

INTERSTAGE COOLING DUTY IS: 3.07 MILLIONS BTU

PRESSURE (PSIA): SUCTION DISCHARGE VOLUME (ACFM): SUCTION DISCHARGE

STAGE #	SUCTION	DISCHARGE	VOLUME (ACFM)	SUCTION	DISCHARGE
1	168.0	309.6	2072.4	1335.8	
2	298.4	550.0	1166.7	752.0	

AIR4.160 JPL4, 2-STAGE AIR COMPRESSOR, WITH AIR FEED @ 160.0% OF STOICHIOMETRIC

CASE AIR4.160
STEP 72
14 JAN 83 9:21:14

72 READ COMBUSTION FUEL VALUE OF RECYCLE COAL INJECTION GAS.
THIS ROUTINE GIVES COMBUSTION HEAT VALUES.

STEP IDENTIFICATION # 72 SOURCE STREAM # 71

COMPONENT	TOTAL FUEL GAS MOLS	%	DRIED FUEL GAS MOLS	%	SWEETENED GAS MOLS	%
H2	1872.2	51.9	1872.2	51.9	1872.2	79.2
CH4	263.3	7.3	263.3	7.3	263.3	11.1
CO	214.5	5.9	214.5	5.9	214.5	9.1
CO2	1243.1	34.5	1243.1	34.5		
H2O	0.0	0.0				
N2	15.0	0.4	15.0	0.4	15.0	0.6
C2H6	0.0	0.0	0.0	0.0	0.0	0.0
H2S	0.0	0.0	0.0	0.0		
NH3	0.0	0.0				
CUS	0.0	0.0	0.0	0.0		
TOTAL	3608.3	100.0	3608.2	100.0	2365.1	100.0

BTU/MOL, NET	86431.		86432.		131859.	
GROSS	XXXXXX		99134.		151238.	
BTU/CU FT, NET	227.		227.		347.	
GROSS	XXX		261.		398.	
TOTAL STREAM COMBUSTION HEAT RELEASE, MILLIONS BTU:						
NET	311.9		311.9		311.9	
GROSS	XXXXXX		357.7		357.7	

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CASE AIR4.160
STEP 73
14 JAN 83 9:21:14

AIR4.160 JPL4, 2-STAGE AIR COMPRESSOR, WITH AIR FEED @ 160.0% OF STOICHIOMETRIC

73 HEAD SENSIBLE HEATS OF REACTOR EXIT RAW GAS (INCL. #71)
THE DUTY OF THIS PROCESS COOLER IS: -208.24 MILLIONS BTU PER HOUR.

COMPONENT	STREA	MULS	NUMBER	%
			73	
H2	15505.4		27.6	
CH4	3513.2		6.3	
CO	12318.5		21.9	
CU2	9197.6		16.4	
H2O	14906.7		26.5	
N2	200.5		0.4	
C2H6	0.0		0.0	
H2S	477.9		0.9	
NH3	74.4		0.1	
GUS	13.7		0.0	
TOTAL	56208.0		100.0	

PROCESS GAS TEMPERATURE, FEED= 1875.0 EXIT= 1507.8 DEGREES FAHRENHEIT.
PROCESS GAS ENTHALPY, FEED= 918.83 EXIT= 710.59 MILLIONS BTU.

END OF THIS CASE.

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*****CONNECT TIME= 0MINS 36SECS *****

APPENDIX Ce. Printouts of Gasifier Balance for
Non-Confidential Fluidized-Bed Cases

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SOLIDS FLOW, MOLES	FEED	FINES	AGGLOMERATES	GAS FLOW, MOLES	FEED1	FEED2	RECYCLE GAS	PRODUCT GAS
C	57.8803	1.0386	0.7662	CO	0.0000	0.0000	2.5778	35.1678
H(2)	26.4385	0.0372	0.0389	CO(2)	0.0000	0.0000	1.4737	20.1049
O	6.2629	0.0000	0.0000	H(2)	0.0000	0.0000	3.0311	41.3523
N(2)	0.4462	0.0022	0.0035	H(2)O	0.0000	57.8803	0.0000	36.2855
S	1.2038	0.0094	0.0128	CH(4)	0.0000	0.0000	0.3814	5.2028
ASH*	100.2000	12.0875	88.1125	O(2)	20.3021	0.0000	0.0000	0.0000
				N(2)	0.1020	0.0000	0.0360	0.4893
MOIST.	1.4233	0.0000	0.0000	H(2)S	0.0000	0.0000	0.0000	1.1488
				CO ₂	0.0000	0.0000	0.0000	0.0328
				NH(3)	0.0000	0.0000	0.0000	0.1785
				HCN	0.0000	0.0000	0.0000	0.0000
TOTAL*	1025.6410	25.0000	97.9028	TOTAL	20.4041	57.8803	7.5000	139.9627
TEMPERATURE, DEGREE F	80.	1875.	1875.	TEMPERATURE, DEGREE F	250.	570.	600.	1875.

* FLOW RATE IS IN LBS

U-GAS GASIFIER MATERIAL AND ENERGY BALANCE

SOLIDS FLOW, LBS	FEED	FINES	AGGLOMERATES	GAS FLOW, LBS	FEED1	FEED2	RECYCLE GAS	PRODUCT GAS
C	695.2000	12.4750	9.2029	CO	0.0000	0.0000	72.2049	985.0507
H	53.3000	0.0750	0.0783	CO(2)	0.0000	0.0000	64.8550	884.7952
O	100.2000	0.0090	0.0000	H(2)	0.0000	0.0000	6.1107	83.3662
N	12.5000	0.0625	0.0979	H(2)O	0.0000	1042.7136	0.0000	653.6830
S	38.6000	0.3000	0.4112	CH(4)	0.0000	0.0000	6.1184	83.4686
ASH	100.2000	12.0875	88.1125	O(2)	649.6264	0.0000	0.0000	0.0000
				N(2)	2.8580	0.0000	1.0085	13.7061
MOIST.	25.6410	0.0000	0.0000	H(2)S	0.0000	0.0000	0.0000	39.1524
				CO ₂	0.0000	0.0000	0.0000	1.9719
				NH(3)	0.0000	0.0000	0.0000	3.0397
				HCN	0.0000	0.0000	0.0000	0.0000
TOTAL	1025.6410	25.0000	97.9028	TOTAL	652.4844	1042.7136	150.2975	2748.2337
TEMPERATURE, DEGREE F	80.	1875.	1875.	TEMPERATURE, DEGREE F	250.	570.	600.	1875.

GASIFIER PRESSURE = 500.00 PSIA, 34.023 ATM, 34.472 BAR

GASIFIER HEAT LOSS = 0. BTU
 COAL FEED (MOISTURE-FREE) = 1000.0 LBS
 FEEDSTOCK HEATING VALUE = 12771.0 BTU/LB (MF)
 FEEDSTOCK MOISTURE CONTENT = 2.500 %
 N₂ CONTENT IN OXIDANT FEED = 0.500 %
 PRODUCT GAS SHIFT TEMP. = 1725.0 DEGREE F

ATOMIC WEIGHTS USED: C = 12.0110 H = 1.0080 O = 15.9990 N = 14.0070 S = 32.0640
 GASEOUS SPECIES : CO CO₂ H₂ CH₄ O₂ H₂S
 MOLECULAR WEIGHT : 28.0100 44.0090 2.0160 18.0150 16.0430 31.9980 28.0140 34.0800 60.0740
 HHV, BTU/SCF : 321.30 0.00 324.90 0.00 1012.00 0.00 0.00
 LHV, BTU/SCF : 321.30 0.00 274.50 0.00 911.20 0.00 0.00

COAL FEED, LBS = 1000.0000 (MF) 1025.6410 (AS FED)
 STEAM IN GAS FEED2 = 57.8803 MOLES (1042.714 LBS)
 RECYCLE GAS FEED = 7.5000 MOLES (150.297 LBS)
 O(2) + N(2) IN GAS FEED1 = 20.4041 MOLES (652.484 LBS)
 O(2) IN GAS FEED1 = 20.3021 MOLES (649.626 LBS)
 N(2) IN GAS FEED1 = 0.1020 MOLES (2.858 LBS)
 TOTAL GAS FEED1 + FEED2 = 78.2844 MOLES (O2 = 25.93 %, STEAM = 73.94 %, REMAINING = 0.13 %)

PRODUCT GAS COMPOSITION, MOLE PERCENT:
 CO 25.127 H2O 29.545 H2 3.717 O2 0.000 H2S 0.821 HCN 0.000
 H2O AND NH3 FREE: 33.979 19.425 39.954 5.027 0.000 0.473 1.110 0.000
 MOLECULAR WEIGHT = 19.6355 (GAS VOLUME = 7013.403 CU FT : AT REACTOR TEMP AND PRESSURE)
 RECYCLE GAS COMPOSITION, MOLE PERCENT:
 CO 34.371 H2O 19.649 40.415 0.000 5.085 0.000 0.480 0.000 0.000
 MOLECULAR WEIGHT = 20.0397

TOTAL H2O,H2S,COS,NH3,HCN FREE PRODUCT GAS = 102.3170 MOLES (2050.387 LBS)
 COMPOSITION, MOLE PERCENT : 34.371 19.650 40.416 5.085 0.478 (ORDER: CO CO2 H2 CH4 N2)
 MOLECULAR WEIGHT = 20.0395
 HIGHER HEATING VALUE = 293.206 BTU/SCF (LHV = 267.711 BTU/SCF)
 NET GAS MAKE = 94.8171 MOLES
 NET GAS PRODUCTION = 0.1052548E+08 BTU
 NET PRODUCT GAS HHV/COAL MHV = 0.8242

CARBON GASIFICATION EFFICIENCY = 98.49 %
 CARBON UTILIZATION EFFICIENCY = 96.88 %
 HAF COAL GASIFICATION EFFICIENCY = 98.76 %
 HAF COAL UTILIZATION EFFICIENCY = 97.48 %
 CONVERSIONS IN FINES : C = 85.12 % N = 95.86 % S = 93.56 %
 CONVERSIONS IN AGGLOMERATES : C = 98.49 % N = 99.11 % S = 98.79 %
 OVERALL CONVERSIONS : C = 96.88 % N = 98.72 % S = 98.16 %
 FINES LOSS (MF) = 2.50 % OF FEED COAL (MF)
 FINES LOSS (MAF) = 1.44 % OF FEED COAL (MAF)
 CARBON IN FINES LOSS = 1.79 % OF FEED COAL CARBON

EXTERNAL STEAM/CARBON = 1.0000 MOL/MOL 1.4999 LB/LB
 EXTERNAL STEAM/COAL (AS FED) = 1.0166 LB/LB
 EXTERNAL STEAM/COAL (MF) = 1.0427 LB/LB
 EXTERNAL STEAM/COAL (MAF) = 1.1588 LB/LB
 TOTAL MOISTURE/CARBON = 1.0246 MOL/MOL 1.5368 LB/LB
 TOTAL MOISTURE/COAL (AS FED) = 1.0416 LB/LB
 TOTAL MOISTURE/COAL (MF) = 1.0684 LB/LB
 TOTAL MOISTURE/COAL (MAF) = 1.1873 LB/LB
 FEED OXYGEN (OR AIR)/CARBON = 0.3525 MOL/MOL 0.9386 LB/LB
 FEED OXYGEN (OR AIR)/COAL (AS FED) = 0.6362 LB/LB
 FEED OXYGEN (OR AIR)/COAL (MF) = 0.6525 LB/LB
 FEED OXYGEN (OR AIR)/COAL (MAF) = 0.7251 LB/LB
 RECYCLE GAS/COAL (AS FED) = 0.1465 LB/LB 0.0823 CU FT/LB (AT 515.0 PSIA AND 80. F)
 RECYCLE GAS/COAL (MF) = 0.1503 LB/LB 0.0844 CU FT/LB (AT 515.0 PSIA AND 80. F)
 GOAL FEED (AS FED)/RECYCLE GAS = 6.8241 LB/LB 12.1548 LB/CU FT (AT 515.0 PSIA AND 80. F)
 COAL FEED (MF)/RECYCLE GAS = 6.6535 LB/LB 11.8509 LB/CU FT (AT 515.0 PSIA AND 80. F)

EXTERNAL STEAM CONVERSION = 37.31 %
 TOTAL FEED MOISTURE CONVERSION = 38.81 %
 NET METHANE PRODUCED = 8.33 % OF FEED CARBON
 KINETIC RATE CONSTANT = 0.1697315E-01

IF PRODUCT GAS IS SHIFTED AT THE GASIFIER TEMPERATURE THEN SPECIES CO CO2 H2 H2O : 36.8480 18.4247 39.6721 37.9656

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SOLIDS FLOW, MOLES	FEED	FINES	AGGLOMERATES	GAS FLOW, MOLES	FEED1	FEED2	RECYCLE GAS	PRODUCT GAS
C	57.8803	1.0386	0.7662	CO	0.0000	0.0000	2.4457	31.5488
H(2)	26.4385	0.0372	0.0389	CO(2)	0.0000	0.0000	1.6060	20.7317
O	6.2629	0.0000	0.0000	H(2)	0.0000	0.0000	2.7559	35.5357
N(2)	0.4462	0.0022	0.0035	H(2)O	0.0000	57.8803	0.0000	35.8421
S	1.2038	0.0094	0.0128	CH(4)	0.0000	0.0000	0.6548	8.4687
ASH*	100.2000	12.0875	88.1125	D(2)	18.8314	0.0000	0.0000	0.0000
				N(2)	0.0746	0.0000	0.0376	0.4834
MOIST.	1.4233	0.0000	0.0000	H(2)S	0.0000	0.0000	0.0000	1.1488
				CO ₂	0.0000	0.0000	0.0000	0.0328
				NH(3)	0.0000	0.0000	0.0000	0.1785
				H ₂ N	0.0000	0.0000	0.0000	0.0000
TOTAL*	1025.6410	25.0000	97.9026	TOTAL	18.9261	57.8803	7.5000	133.9705
TEMPERATURE, DEGREE F	80.	1875.	1875.	TEMPERATURE, DEGREE F	250.	570.	600.	1875.

* FLOW RATE IS IN LBS

U-GAS GASIFIER MATERIAL AND ENERGY BALANCE

SOLIDS FLOW, LBS	FEED	FINES	AGGLOMERATES	GAS FLOW, LBS	FEED1	FEED2	RECYCLE GAS	PRODUCT GAS
C	695.2000	12.4750	9.2029	CO	0.0000	0.0000	68.5034	883.6811
H	53.3000	0.0750	0.0783	CO(2)	0.0000	0.0000	70.6774	912.3796
O	100.2000	0.0000	0.0000	H(2)	0.0000	0.0000	5.5560	71.6399
N	12.5000	0.0625	0.0979	H(2)O	0.0000	1012.7136	0.0000	645.6957
S	38.6000	0.3000	0.4112	CH(4)	0.0000	0.0000	10.5054	135.8627
ASH	100.2000	12.0875	88.1125	D(2)	602.5687	0.0000	0.0000	0.0000
				N(2)	2.6510	0.0000	1.0526	13.5432
MOIST.	25.6410	0.0000	0.0000	H(2)S	0.0000	0.0000	0.0000	39.1524
				CO ₂	0.0000	0.0000	0.0000	1.9719
				NH(3)	0.0000	0.0000	0.0000	3.0397
				H ₂ N	0.0000	0.0000	0.0000	0.0000
TOTAL	1025.6410	25.0000	97.9028	TOTAL	605.2197	1042.7136	156.2947	2706.9662
TEMPERATURE, DEGREE F	80.	1875.	1875.	TEMPERATURE, DEGREE F	250.	570.	600.	1875.

GASIFIER PRESSURE = 500.00 PSIA, 34.023 ATM, 34.472 BAR
 GASIFIER HEAT LOSS = 0. BTU
 COAL FEED (MOISTURE-FREE) = 1000.0 LBS
 FEEDSTOCK HEATING VALUE = 12771.0 BTU/LB (MF)
 FEEDSTOCK MOISTURE CONTENT = 2.500 %
 N₂ CONTENT IN OXIDANT FEED = 0.500 %
 PRODUCT GAS SHIFT TEMP. = 1725.0 DEGREE F

ATOMIC WEIGHTS USED: C = 12.0110 H = 1.0080 O = 15.9990 N = 14.0070 S = 32.0640
 GASEOUS SPECIES : CO CO₂ H₂ H₂O CH₄ O₂ N₂ H₂S COS NH₃ HCN
 MOLECULAR WEIGHT : 28.0100 44.0090 2.0160 18.0150 16.0430 31.9980 28.0140 34.0800 60.0740 17.0310 27.0260
 HHV, BTU/SCF : 321.30 0.00 324.90 0.00 1012.00 0.00 0.00
 LHV, BTU/SCF : 321.30 0.00 274.50 0.00 911.20 0.00 0.00

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COAL FEED, LBS = 1000.0000 (MF) 1025.6410 (AS FED)
 STEAM IN GAS FEED2 = 57.8803 MOLES (1042.714 LBS)
 RECYCLE GAS FEED = 7.5000 MOLES (156.295 LBS)
 O(2) + N(2) IN GAS FEED1 = 18.9261 MOLES (605.220 LBS)
 O(2) IN GAS FEED1 = 18.8314 MOLES (602.569 LBS)
 N(2) IN GAS FEED1 = 0.0946 MOLES (2.651 LBS)
 TOTAL GAS FEED1 + FEED2 = 76.8064 MOLES (O2 = 24.52 %, STEAM = 75.36 %, REMAINING = 0.12 %)

PRODUCT GAS COMPOSITION, MOLE PERCENT:
 CO 23.549 15.475 26.525 26.754 6.321 0.000 0.361 0.858 0.000
 H2O AND NH3 FREE: 32.209 21.166 36.279 8.646 0.000 0.494 1.173 0.034 0.000
 MOLECULAR WEIGHT = 20.2057 (GAS VOLUME = 6713.139 CU FT : AT REACTOR TEMP AND PRESSURE)
 RECYCLE GAS COMPOSITION, MOLE PERCENT: CO CO2 H2 CH4 O2 N2 H2S COS NH3 HCN
 32.609 21.913 36.746 0.000 8.731 0.000 0.501 0.000 0.000 0.000 0.000

MOLECULAR WEIGHT = 20.8393

TOTAL H2O,H2S,COS,NH3,HCN FREE PRODUCT GAS = 96.7682 MOLES (2017.107 LBS)
 COMPOSITION, MOLE PERCENT : 32.602 21.424 36.722 8.751 0.500 (ORDER: CO CO2 H2 CH4 N2)
 MOLECULAR WEIGHT = 20.8447
 HIGHER HEATING VALUE = 312.628 BTU/SCF (LHV = 285.298 BTU/SCF)
 NET GAS MAKE = 89.2708 MOLES
 NET GAS PRODUCTION = 0.1056621E+08 BTU
 NET PRODUCT GAS HHV/COAL HHV = 0.8274

CARBON GASIFICATION EFFICIENCY = 98.49 %
 CARBON UTILIZATION EFFICIENCY = 96.89 %
 MAF COAL GASIFICATION EFFICIENCY = 98.73 %
 MAF COAL UTILIZATION EFFICIENCY = 97.48 %
 CONVERSIONS IN FINES : C = 85.12 % N = 95.86 % S = 93.56 %
 CONVERSIONS IN AGGLOMERATES : C = 98.49 % N = 99.11 % S = 98.79 %
 OVERALL CONVERSIONS : C = 96.88 % N = 98.72 % S = 98.16 %
 FINES LOSS (MF) = 2.50 % OF FEED COAL (MF)
 FINES LOSS (MAF) = 1.44 % OF FEED COAL (MAF)
 CARBON IN FINES LOSS = 1.79 % OF FEED COAL CARBON

EXTERNAL STEAM/CARBON = 1.0000 MOL/MOL 1.4999 LB/LB
 EXTERNAL STEAM/COAL (AS FED) = 1.0166 LB/LB
 EXTERNAL STEAM/COAL (MF) = 1.0427 LB/LB
 EXTERNAL STEAM/COAL (MAF) = 1.1588 LB/LB
 TOTAL MOISTURE/CARBON = 1.0246 MOL/MOL 1.5368 LB/LB
 TOTAL MOISTURE/COAL (AS FED) = 1.0416 LB/LB
 TOTAL MOISTURE/COAL (MF) = 1.0684 LB/LB
 TOTAL MOISTURE/COAL (MAF) = 1.1873 LB/LB
 FEED OXYGEN (OR AIR)/CARBON = 0.3270 MOL/MOL 0.8706 LB/LB
 FEED OXYGEN (OR AIR)/COAL(AS FED) = 0.5901 LB/LB
 FEED OXYGEN (OR AIR)/COAL (MF) = 0.6052 LB/LB
 FEED OXYGEN (OR AIR)/COAL (MAF) = 0.6726 LB/LB
 RECYCLE GAS/COAL (AS FED) = 0.1524 LB/LB
 RECYCLE GAS/COAL (MF) = 0.1563 LB/LB
 COAL FEED (AS FED)/RECYCLE GAS = 6.5622 LB/LB 12.1548 LB/CU FT (AT 515.0 PSIA AND 80. F)
 COAL FEED (MF)/RECYCLE GAS = 6.3982 LB/LB 11.8509 LB/CU FT (AT 515.0 PSIA AND 80. F)

EXTERNAL STEAM CONVERSION = 38.08 %
 TOTAL FEED MOISTURE CONVERSION = 39.56 %
 NET METHANE PRODUCED = 13.50 % OF FEED CARBON
 KINETIC RATE CONSTANT = 0.1911822E-01

IF PRODUCT GAS IS SHIFTED AT THE GASIFIER TEMPERATURE THEN SPECIES CO CO2 H2 H2O : 33.1571 19.1234 33.9274 37.9504

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APPENDIX Da. Analysis of Variations in Basic Assumptions

ANALYSIS OF VARIATIONS IN BASIC ASSUMPTIONS

General

During the course of the work effort, it became apparent that significant variations in the system efficiency could be achieved with relatively minor adjustments in the operating parameters. The systems analyst must use extreme care to be certain that two systems to be compared are identical in all respects except the variable being altered, to avoid clouding the results by the effects of unwanted changes. In fact, many of the cases presented in Appendix Ca were reevaluated several times during the course of the program as the analyst became more familiar with the overall system and sought to ensure strictly comparable results for analysis. For that matter, the results presented in the body of the text, using essentially the same systems as employed in Appendix Ca, show slightly different efficiencies because one last iteration was used to ensure identical assumptions.

The above effect occurs because of the relative complexity of the system and the number of elements employed. At each step of the process, many hundreds of millions of Btu's per hour are being transferred or converted. The final efficiency is determined by summing the power productions and consumptions in the overall system — the sums and differences of very large numbers. However, a difference of only 25,000 to 50,000 Btu/hr, in recovered power, might be considered a significant change in efficiency. Only a small variation in certain of the operating parameters is sufficient to generate such an effect. This change might be as simple as the change in the initial assumptions on one of the subelements of the process.

The above effect raises concern about the comparability of results prepared by different organizations, or even different analysts, unless the bases for these results are clearly defined. For example, this analysis showed a much lower efficiency for the primary base case (using the entrained-bed gasifier) than had been developed by earlier investigators. Similar discrepancies with the literature were also noted for other cases.

Consequently, a subprogram was undertaken to illustrate the apparent variation in overall system performance, employing essentially the same system, depending upon the assumptions selected by the analyst for various elements in the subsystem.

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This appendix is a discussion of the results of that subinvestigation. Table Da-1 presents the result of 21 variations on the same system, yielding overall system efficiencies between 45.4% and 56.5%, depending upon the initial assumptions employed. The data of Table Da-1 were generated in a fashion consistent with the original efficiency analysis of Appendix Ca — a constant quantity of coal was fed to the system and variation in power output were analyzed. Summary printouts for the primary cases of this analysis are presented in Appendix Db.

The first analysis presented uses the same degree of conservatism that was employed throughout the majority of the program; somewhat less restrictive assumptions were then employed for each of the 20 other variations upon the system. These assumptions, however, would not result in an inoperable system — a quench is used on the raw gas in each case to avoid carbon deposition and trains of crossflow heat exchangers are used for heat recovery, rather than the countercurrent heat exchange reported in many other designs. Fuel gas conversions was held constant at 85% in the fuel cell in each case. Also, the current density and, after initial evaluation, the quantity of air supplied to the cathode were held constant. Thus, the modifications employed are primarily variations in the assumptions of the operating conditions of the various elements of the system.

Case A — The Base Case

The basic system for this analysis was the combination of a methane-producing, fluidized-bed gasifier integrated with an internally-reforming molten carbonate fuel cell as illustrated in Figure Da-1. This is the case labeled IV in Appendix Ca. This base case system had indicated significant promise in the earlier evaluations and appears to be a near-term process efficiency improvement that is currently undergoing evaluation in several fuel cell laboratories. Because all of the systems employed in this analysis are variations in the assumptions underlying this basic case, Figure Da-1 is applicable for all of the cases reported in this appendix.

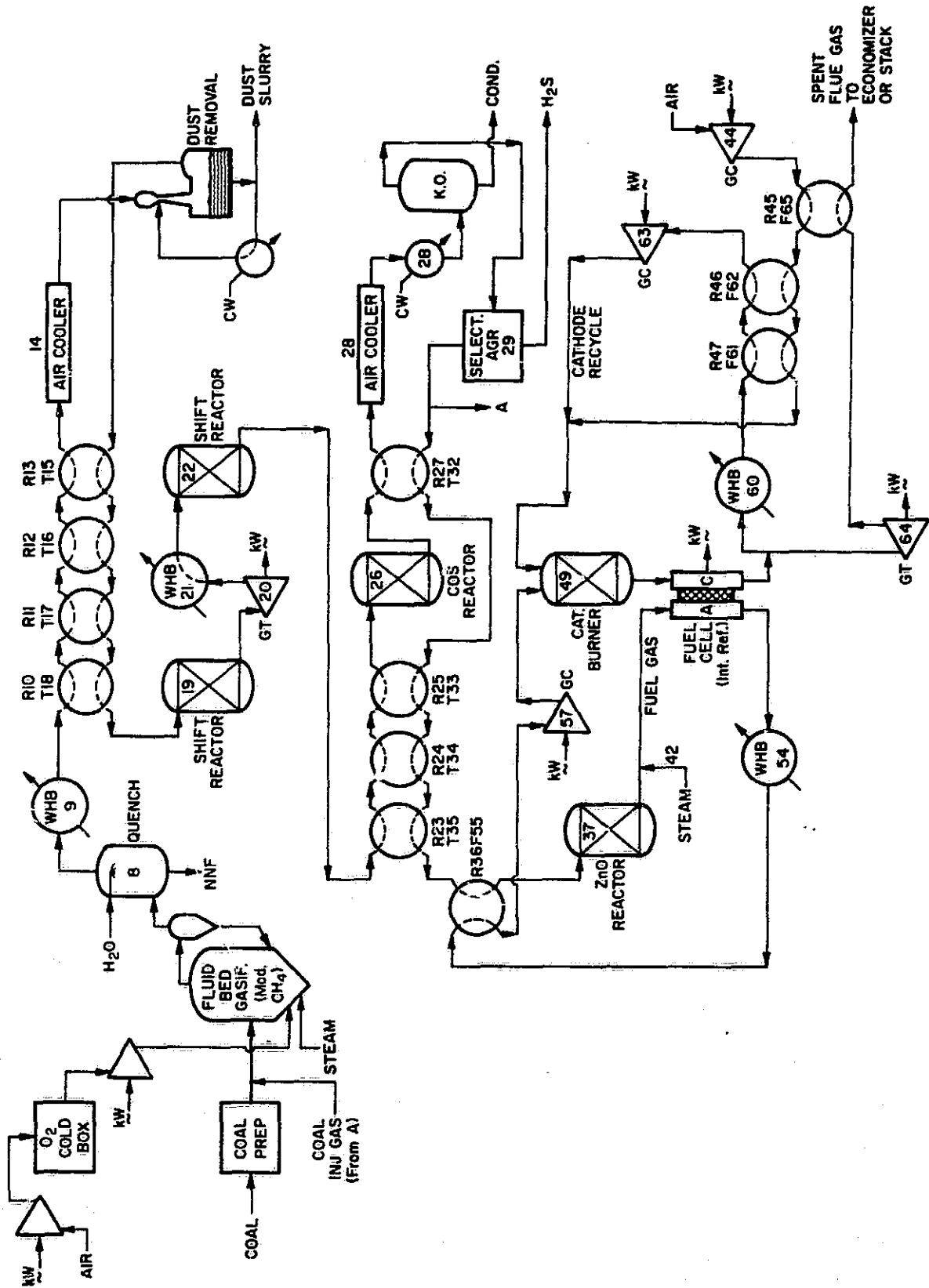
The major assumptions underlying this case have been outlined in the general discussion of this document. Because many of these assumptions are critical to understanding the overall impact of this evaluation of initial assumptions, they will be reviewed here.

Table Da-1. ANALYSIS OF VARIATIONS IN INITIAL ASSUMPTIONS
(Normal Feed - 4991 Tons/Day)

Case	System	Power Output			Eff. %	Comments
		Net Mech	Fuel Cell	Steam		
		MW				
		Net	Net	Net		
A	Base (No. IV of Table Ca-1)	51	651	4	706	45.4
B	Add 2-Stage Compression of Air	54	651	5	710	45.6
C	Add Optimized Air Rate (Case IVe)	60	651	1	(694)-712	45.8 (5 Cases)
D	Add High-Temperature Heat Recovery	60	653	15	728	46.8
E	Add Expanders for low-Pressure Steam Use	96	653	15	764	49.1
F	Add Tail Gas Economizer to 250°F	93	653	52	798	51.3
G	Add Minimum Quenching to Amorphous Carbon	90	656	59	805	51.7
H	Change Gasifier to Optimistic Design	86	665	52	803	51.6
I	Change Quench From Water to Steam	99	665	58	822	52.8
J	Remove More CO ₂ With H ₂ S				806-808	(2 Cases)
K	As I, Burn Part of Anode Exhaust to Cathode Exhaust				816-817	(2 Cases)
L	As I, Low AP in System	128	665	48	841	54.0
M	Add Increase Fuel Cell Exhaust to 1300°F	137	667	44	848	54.5
N	Add Increase Steam System Efficiency to 40%	137	667	50	854	54.9 Estimated
O	Add Correct Cell Voltage Based on Experimental	137	710	32	879	56.5 Estimated

Other Potential Changes; Not Made:
 -- Shift Reaction Temperature-of-Approach Minimized
 -- Heat Exchanger Temperature-of-Approach Minimized
 -- As J but Recompress Purified CO₂ to Cathode
 -- As O but Minimum Cell Thickness, Other Future Improvements.

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Figure Da-1. PROCESS FLOW DIAGRAM FOR THE GENERAL SYSTEM EMPLOYED IN ANALYSIS
OF INITIAL ASSUMPTIONS

The gasifier operates in a fluidized-bed mode at approximately 500 psi using a steam/carbon molar ratio of 1.0. Dust losses, carbon reporting to the ash, and the need for fuel injection gas are considered in the design. The gasifier is in thermal and mass balance and the off-gas is assumed to contain 13.5% of the feed carbon in the form of methane (6.1% methane in the raw, wet gas), based upon extrapolation of current pilot plant operating data.

After initial dust recovery to regasification, the raw gas is quenched with sufficient liquid water to render the gas thermodynamically stable for carbon deposition, using a factor of safety in the water addition such that the resulting gas composition is midway between that required for stability when considering both amorphous and graphitic carbon. In other words, the minimum amount of water required for this system is that calculated for the deposition of carbon in the amorphous form, which has a higher thermodynamic activity than carbon in a graphitic form. This minimum water is required because carbon is anticipated to condense in the amorphous form. However, the factor of safety is employed by calculating the amount of water required for deposition of graphitic carbon and selecting the average quantity of water.

The quench gas is cooled in a recuperative heat exchange train so that it may be washed for final dust removal and then it is reheated in the same train. Much of the water content is converted to hydrogen by reaction over a water-gas shift catalyst with a temperature-of-approach to thermodynamic equilibrium of approximately 50°F.

The heat of the exothermic shift reaction elevates the temperature of the gas. Power is then recovered by expansion of the gas to a pressure level consistent with the 150 psi fuel cell operating pressure (and intermediate pressure drops within the system). After heat recovery, the gas is passed through another shift reactor to take advantage of the improved shift equilibrium at lower temperatures.

The gas is then cooled through a second recuperative heat exchange train. Midway within the train is a reactor to drive the carbonyl sulfide hydrolysis reaction toward completion at a preferred operating condition, thus minimizing the amount of COS that must be removed from the system. After final air- and water-cooling, the gas is scrubbed for selective sulfur removal in a Selexol system. The acid-gas removed from the system is assumed to contain 19% H₂S; COS is distributed between the acid gas and the treated gas according to data from several designs.

The sweetened gas is reheated in the heat exchange train, finally passing over zinc oxide to remove the last traces of sulfur. Steam, after energy recovery in an expansion turbine, is added to the system to eliminate the potential for carbon deposition in the downstream fuel cell. Again, the quantity of steam added is taken as midway between that required for amorphous carbon deposition and graphitic carbon deposition.

The fuel cell system of Figure Da-1 incorporates significant heat exchange to alleviate operating conditions in the recycle blowers and to ensure operability of the catalytic combustor in the temperature range of commercially available catalysts. In most systems, the fuel cell off-gas temperature is 1260°F. The air supply to the fuel cell employs a conservative compressor design of three stages. The final exhaust from the system is about 500°F to minimize plume formation from the very wet off-gas discharged from the atmosphere. Pressure drops throughout the system were identified in the body of this document, and the efficiency of recovery of waste heat to steam and power is assumed to be 35.2%, as discussed elsewhere in this document. The overall power produced from this system, using the base amount of coal, was 706 MW for a system efficiency of 45.4%.

Case B — Modified Air Compressor

The first modification of the base case was to change the air compressor from three stages to two stages. For this modification, it is assumed that compressors are available with outboard bearings and seals that can withstand the higher off-gas temperature encountered. This approach minimizes the heat lost to interstage cooling and results in more heat present in the compressed cathode air. Note that not all of this gain was readily recoverable — the balancing of the heat exchange system required that the system off-gas temperature be increased to approximately 540°F. Nevertheless, the addition of this modification increased the power output of the system to 710 MW and the efficiency to 45.6%.

Case C — Optimized Cathode Air Supply

Five cases were run at varying quantities of air flow to the fuel cell, to determine if significant improvement can be achieved. The initially assumed air flow was near optimum; modification raised the system output to 712 MW and a system efficiency of 45.8%. This case became Case IVe of the

initial analysis and Case 5 in the body of the report. The complete system computer printout of this case is included in Appendix Cd.

Case D — Add High-Temperature Heat Recovery

For this modification, high-temperature heat recovery was added. Heat is available both before and after the quench that can be recovered to more useful application than the initial assumption of converting it to steam or steam superheat. Heat exchangers were added on the raw gas stream, both before and after the initial quench, to drive a greater quantity of heat into the shifted off-gas, prior to the expansion turbine. Caution was exercised to ensure that the raw gas was not cooled into the carbon-depositing regime before quench water was added. For this modification, the availability of high-temperature heat recuperation equipment was assumed. Incoloy 800 or Inconel 600 would probably be the preferred materials of construction. This modification increases the overall power output of the system to 728 MW and the system efficiency increased to 46.8%.

Case E — Add Steam Expanders

This system modification included steam expanders for all the low-pressure steam used throughout the system. Prior examples had assumed that the steam employed in the gasifier and the acid-gas removal system was raised at the conditions necessary. For this case, it is assumed that all steam is raised at 1500 psi and superheated to 1000°F. As is discussed elsewhere, the accurate steam balance cannot be projected in this effort because the total system of steam make and consumption is not defined (sulfur recovery, tail gas treatment, power water stripping, ammonia recovery, etc.). Therefore, the steam system had been assumed to correspond to that required and the quantity of steam raised is assumed to correspond with the Btu's available.

Again, for this case the steam quantity corresponds to Btu's available; however, it is assumed that low-temperature heat is used for BFW heating, moderate-temperature heat is used for vaporization, and high-temperature heat is available for superheat. Thus all the steam is produced at high pressure and superheat. For this option, the quantity of steam make decreases, but the quality is higher. Delivered shaft horsepower increased significantly as expanders were included between the steam header and the lower pressure steam consumers. The total system power increased to 764 MW and the efficiency climbed 49.1%.

Case F — Add Flue-Gas Economizer

Case F includes an economizer on the cathode exhaust tail gas. In prior examples, the tail gas was exhausted to the atmosphere at temperatures of about 500°F to avoid the political problems of plumes from the wet exhaust. Common power plant practice is to discharge the stack gas at about 320° to 350°F to avoid a stack plume (except with cold ambient air.) With the very wet exhaust of this system, a higher stack-gas temperature of 500° had been selected for the same reason.

For this option, it was assumed that such political objections would not be present and that the exhaust could be cooled to 250°F. The heat is recovered into the boiler feedwater preheat and, eventually, into the high-pressure steam system. The improved heat recovery results in more power recovered from the steam system, increasing the total system power to 798 MW and the system efficiency to 51.3%.

Case G — Minimum Quench

For this option, the quantity of water used to quench the gas is reduced to the thermodynamic minimum. Thus, both the raw gas and the sweetened gas, which are unstable for carbon deposition in certain temperature ranges, are humidified only to the amorphous carbon limit. This effect improves the heat recovery from the system and, also, slightly improves the fuel cell power because the Nernst concentration polarization at the anode is reduced as the steam fed to the anode is minimized. The total system power increased to 805 MW and the efficiency increased to 51.7%.

Case H — Optimistic Gasifier Design

The design of the fluidized-bed gasifier was optimized to minimize the factor of safety employed. The steam/carbon ratio was reduced to 0.75 and the operating pressure was increased to 600 psi, the limit of currently available lockhopper valves. Interestingly, the total system performance does not change significantly, on an overall basis. The altered gas composition required more quench water, thus negating the effect of the reduced gasifier steam consumption. The gas quality was somewhat improved, resulting in higher fuel cell output, but the total system power remained the essentially unchanged at 803 MW with a system efficiency of 51.6%.

Case I — Quench With Steam

The form of water used in the quench system was changed from liquid water (Case J) to steam for this modification. Liquid water had been used in the earlier analyses to minimize the size of BFW treatment system. Although the total moles of water (or water vapor) added are the same, the effect improves system efficiency because high-temperature heat is not used to provide the latent heat of water vaporization. Thus, the overall system improved to 822 MW and the system efficiency improved to 52.8%.

Case J — Remove More Carbon Dioxide

For this option, the acid gas was removed from the sour gas at lower H₂S concentration. This approach removes more CO₂ with the H₂S and permits reduced steam requirement for stabilization of the anode fuel. Operationally, it would be assumed that a Stretford process or the Ralph N. Parsons "Selectox" system would be used for sulfur recovery. Two cases were analyzed in this approach. The overall power from the facility declined because of greater cathode concentration polarization. Therefore, this option was not considered further. It should be noted, however, that system performance may well increase if the purified CO₂, exiting from the Stretford process, were to be recompressed into the cathode feed.

Case K — Burn a Portion of Anode Exhaust Into Cathode Exhaust

For this modification, the system of Case I was employed, but a portion of the anode exhaust was combusted directly into the cathode exhaust, rather than all of the anode exhaust being combusted into the cathode feed. This modification was made because it had been noted that the power recovered from the cathode exhaust turbine was very sensitive to the operating temperature; slightly increasing the temperature by combustion was attempted to improve the overall system power. However, the increased Nernst concentration voltage loss at the cathode, due to bypassing part of the carbon dioxide, caused a slight reduction in the system performance. Therefore, this system was not considered further.

Case L — Reduced Pressure Drops

For this modification, the basic system was the same as Case I but it was assumed that the pressure drop across each element of the system could be decreased by a factor of two. As was discussed in the body of this report,

the pressure drops assumed for the various elements in the system are relatively low, compared with normal chemical process industry practice. However, several designs appearing in the literature have assumed constructional characteristics that would further decrease the pressure drops significantly. Therefore, this case was developed to indicate the effect of lower pressure drop. The reduced pressure losses in the system permitted much greater shaft horsepower recovery, providing 841 MW from the base coal fed to the system and a system efficiency of 54.0%.

Case M -- Higher-Temperature Cathode Exhaust

As was discussed under Case K, higher temperatures in the cathode exhaust system are desirable. For this modification, the flow of the cathode recycle cooling gas around the fuel cell was controlled to result in a fuel cell off-gas temperature of 1300°F -- a value assumed to be operable without excessive electrolyte loss. The increase in the fuel cell exhaust gas temperature of only 40°F improved the total system performance to 848 MW, with a total system efficiency of 54.5%.

Case N -- Higher Steam System Efficiency

For this option, the efficiency of recovery of heat from waste steam into power is assumed to increase to 40%. All earlier analyses had used an efficiency of 35.2%, a value that includes penalties for boiler feedwater treatment and the cooling water systems. However, modern power plants operate near this efficiency and include penalties for SO₂ scrubbing (not required here). Thus, an improvement of efficiency to 40% would be considered reasonable. For this presentation, the modification was estimated, rather than calculated on the system simulator. Such simulations had been made earlier in the program and the impact is the same as is estimated here. The total system power increased to 854 MW and the efficiency climbed to 54.9%.

Case O -- Correct Fuel Cell Voltage

As is discussed elsewhere in this document, the modified fuel cell simulator subpacket reported a fuel cell voltage that was somewhat lower than the current experimental performance. Consequently, the cell potential was increased by 55 mV with a corresponding reduction in the waste heat from the system. This change resulted in a total system performance improvement to 879 MW with a system efficiency of 56.5%. Again, the process simulator was not

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used for this comparison because of the time required for simulator modification; rather, hand calculation was employed and the results are reported as estimated.

Summary

The analysis above indicates that a wide variation in the reported efficiency of the gasifier/fuel cell system can be obtained, depending upon the assumptions employed by the system analyst in setting the design conditions for each element within the system. It is therefore suggested that comparisons of systems derived by different analysts be evaluated carefully, because even minor variations in basic assumptions may cause an apparent system efficiency change of over 1 percentage point. Any such comparative analysis must be made with care to be certain that differences in apparent system performance are real.

APPENDIX Db. Summary Printouts for Cases
in Appendix Da

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CASE A

The Base Case

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TANULATION UP SHAFI ENERGY AVAILABLE OR NEEDED.

97	-39844.7	KILOWATTS	AIR COMPRESSOR FOR SUPPLY TO CRYOGENIC OXYGEN PLANT
99	-16426.3	KILOWATTS	CRYOGENIC OXYGEN COMPRESSOR FOR SUPPLY TO COAL GASIFIER
2	-5240.3	KILOWATTS	2 U-GAS GASIFIER NORMAL OPERATION. STREAM #2 = NET MAKE.
20	31756.2	KILOWATTS	20 EXHAUST SHIF (1) EXIT GAS THROUGH POWER RECOVERY TURBINE.
24	-8457.4	KILOWATTS	24 REMOVE SULFUR FROM RAW GAS BY SCLEROL SYSTEM.
42	10647.5	KILOWATTS	42 ADD STEAM VIA TURBINE TO PREVENT CARBON DROIP IN FUEL CELL.
44	-170942.0	KILOWATTS	44 COMPRESS COMBUSTION AIR TO FUEL CELL PRESSURE.
51	663492.0	KILOWATTS	51 FINE FUEL CELL, MAXIMUM CONVERSION, REFORM WITHIN ANODE.
57	-8324.8	KILOWATTS	57 ANODE EXIT GAS BOOSTER FAN, TO FEED CATALYTIC BURNER.
63	-14446.7	KILOWATTS	63 CATHODE RECYCLE GAS CIRCULATION FAN.
64	274032.7	KILOWATTS	64 EXPAND STACK GAS THROUGH TURBINE TO RECOVER POWER.
70	-1747.3	KILOWATTS	70 COMPRESS RECYCLE GAS TO INJECT FEED COAL TO GASIFIER.

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JPL 06.06
CINSE
STREAM LIST
4 JAN 83 14135:19

JPL-0604 FUEL CELL REFORMER, 454 TONS, 4000 PSI, 13.5% CH4 RAW

INDICATION OF STEAM NEEDED IN PROCESS.

2	500.00 PSIA	24072.0	MOLS STEAM	2.0	U-GAS GASIFIER NORMAL OPERATION. STREAM #2 & #1 MAKE.
29	45.00 PSIA	17819.0	MOLS STEAM	7.0	REMOVE SULFUR FROM RAW GAS BY SFLXOL SYSTEM.
42	15.00 PSIA	8015.0	MOLS STEAM	4.2	AND STEAM VIA TURBINE TO PREVENT CARBON DROP IN FUEL CELL.

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CASE JPLUG-4
WASTE HEAT OUT
6 JAN 83 14135119

JPLUG-4 FLOID BED HEATERS FOR CONVE QUENCH, 1st-SHIFT, COSA, 13.5% CHG RAW

TABULATION OF WASTE HEAT AVAILABLE FROM PROCESS HEATERS.

	5	700.0 DEGREES F.	428.66 MILLIONS BTU	9 COOL RAW GAS TO 700 DEGREES FARENHEIT, MAKE STEAM.
21	600.0 DEGREES F.	10.00 MILLIONS BTU	21 COOL INTER-STAGE SHIFT GAS IN WHBI RAISE STEAM.	
54	925.0 DEGREES F.	319.00 MILLIONS BTU	56 COOL ANODE EXHAUST GAS IN MMH - RAISE STEAM.	
60	1000.0 DEGREES F.	199.00 MILLIONS BTU	60 COOL CATHODE RECYCLE GAS IN MMH - RAISE STEAM.	

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CASE B

Utilize Two-Stage Air Compression

JPL.25.4 FLUID BED, REFORM, 85% CONV, QUENCH, PRE-SHIFT, COSX, 13.5% CH4 RAW

TABULATION OF SHAFT ENERGY AVAILABLE OR NEEDED.

97	-39844.7	KILOWATTS	AIR COMPRESSOR FOR SUPPLY TO CRYOGENIC OXYGEN PLANT
99	-16426.3	KILOWATTS	CRYOGENIC OXYGEN COMPRESSOR FOR SUPPLY TO COAL GASIFIER
2	-5240.3	KILOWATTS	U-GAS GASIFIER NORMAL OPERATION, STREAM #2 = NET MAKE.
20	31756.2	KILOWATTS	EXPAND SHIFT (1) EXIT GAS THROUGH POWER RECOVERY TURBINE.
29	-6958.0	KILOWATTS	REMOVE SULFUR FROM RAW GAS BY SELEXOL SYSTEM.
42	10667.7	KILOWATTS	ADD STEAM VIA TURBINE PREVENT CARBON DROP IN FUEL CELL.
44	-171083.5	KILOWATTS	COMPRESS COMBUSTION AIR TO FUEL CELL PRESSURE.
51	663891.7	KILOWATTS	THE FUEL CELL, MAXIMUM CONVERSION, REFORM WITHIN ANODE.
57	-8924.9	KILOWATTS	ANODE EXIT GAS BOOSTER FAN, TO FEED CATALYTIC BURNER.
63	-20371.1	KILOWATTS	CATHODE RECYCLE GAS CIRCULATION FAN.
64	281757.5	KILOWATTS	EXPAND STACK GAS THROUGH TURBINE TO RECOVER POWER.
70	-1797.3	KILOWATTS	COMPRESS RECYCLE GAS TO INJECT FEED COAL TO GASIFIER.

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JPL-2S.4
CASE JPL-2S.4
STEAM LIST
13 JAN 83 10114126

JPL-2S.4 FLUID BED, REFORM, 85% CONV, QUENCH, PRE-SHIFT, COSA, 13.5% CH4 RAW

TABULATION OF STEAM NEEDED IN PROCESS.

2	500.0 PSIA	24072. MOLS STEAM	2 U-GAS GASIFIER NORMAL OPERATION. STREAM #2 = NET MAKE.
29	65.0 PSIA	11889. MOLS STEAM	11889. MOLS STEAM 29 REMOVE SULFUR FROM RAW GAS BY SELEXOL SYSTEM.
42	1500.0 PSIA	8015. MOLS STEAM	8015. MOLS STEAM 42 ADD STEAM VIA TURBINE; PREVENT CARBON DROP IN FUEL CELL.

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JPL-2S.4
WASTE HEAT OUT
13 JAN 83 13114126

JPL-2S.4 FLUID BEU, REFORM, 85% CUNV, QUENCH, PRE-SHIFT, COSX, 13.5% CH4 RAW

TABULATION OF WASTE HEAT AVAILABLE FROM PROCESS BOILERS.

Y	700.0 DEGREES F.	458.66 MILLIONS BTU	9 COOL RAW GAS TO 700 DEGREES FAHRENHEIT, MAKE STEAM.
21	600.0 DEGREES F.	10.06 MILLIONS BTU	21 COOL INTER-STAGE SHIFT GAS IN WHB; RAISE STEAM.
54	925.0 DEGREES F.	338.82 MILLIONS BTU	54 COOL ANODE EXIT GAS IN WHR - RAISE STEAM.
60	1115.0 DEGREES F.	192.44 MILLIONS BTU	60 COOL CATHODE RECYCLE GAS IN WHB - RAISE STEAM.

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CASE C

Optimize Air Flow Rate

TABULATION OF SHAFT ENERGY AVAILABLE OR NEEDED.

97	-39844.7	KILOWATTS	AIR COMPRESSOR FOR SUPPLY TO CRYOGENIC OXYGEN PLANT
99	-16426.3	KILOWATTS	CRYOGENIC OXYGEN COMPRESSOR FOR SUPPLY TO COAL GASIFIER
2	-5240.3	KILOWATTS	2 U-GAS GASIFIER NORMAL OPERATION, STREAM #2 = NET MAKE.
20	31756.2	KILOWATTS	20 EXPAND SHIFT (I) EXIT GAS THROUGH POWER RECOVERY TURBINE.
29	-6958.0	KILOWATTS	29 REMOVE SULFUR FROM RAW GAS BY SELEXOL SYSTEM.
42	10667.7	KILOWATTS	42 ADD STEAM VIA TURBINE1 PREVENT CARBON DROP IN FUEL CELL.
44	-174015.4	KILOWATTS	44 COMPRESS COMBUSTION AIR TO FUEL CELL PRESSURE.
51	664628.2	KILOWATTS	51 THE FUEL CELL, MAXIMUM CONVERSION, REFORM WITHIN ANODE.
57	-8324.9	KILOWATTS	57 ANODE EXIT GAS BOOSTER FAN, TO FEED CATALYTIC BURNER.
63	-15467.8	KILOWATTS	63 CATHODE RECYCLE GAS CIRCULATION FAN.
64	285486.4	KILOWATTS	64 EXPAND STACK GAS THROUGH TURBINE TO RECOVER POWER.
70	-1797.3	KILOWATTS	70 COMPRESS RECYCLE GAS TO INJECT FEED GOAL TO GASIFIER.

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AIR4.160 JPL4, 2-STAGE AIR COMPRESSOR, WITH AIR FEED @ 160.0% OF STOICHIOMETRIC

CASE AIR4-160
STEAM LIST
14 JAN 83 9:21:14

TABULATION OF STEAM NEEDED IN PROCESS.

2	500.0 PSIA	24072. MOLS STEAM	2 U-GAS GASIFIER NORMAL OPERATION. STREAM #2 = NET MAKE.
29	65.0 PSIA	11889. MOLS STEAM	29 REMOVE SULFUR FROM RAW GAS BY SELEXOL SYSTEM.
42	1500.0 PSIA	8015. MOLS STEAM	42 ADD STEAM VIA TURBINE; PREVENT CARBON DROP IN FUEL CELL.

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AIR4-160 JPL4, 2-STAGE AIR COMPRESSOR, WITH AIR FEED @ 160.0% OF STOICHIOMETRIC

CASE AIR4-160
WASTE HEAT OUT
14 JAN 83 9:21:14

TABULATION OF WASTE HEAT AVAILABLE FROM PROCESS BOILERS.

9	700.0 DEGREES F.	458.66 MILLIONS BTU	9 COOL RAW GAS TO 700 DEGREES FAHRENHEIT. MAKE STEAM.
21	600.0 DEGREES F.	10.06 MILLIONS BTU	21 COOL INTER-STAGE SHIFT GAS IN WHBI RAISE STEAM.
54	425.0 DEGREES F.	338.74 MILLIONS BTU	54 COOL ANODE EXIT GAS IN WHB - RAISE STEAM.
60	1115.0 DEGREES F.	157.28 MILLIONS BTU	60 COOL CATHODE RECYCLE GAS IN WHB - RAISE STEAM.

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CASE D

Employ High-Temperature Heat Recovery

TABULATION OF SHAFT ENERGY AVAILABLE OR NEEDED.

97	-39844.7	KILOWATTS	AIR COMPRESSOR FOR SUPPLY TO CRYOGENIC OXYGEN PLANT
99	-16426.3	KILOWATTS	CRYOGENIC OXYGEN COMPRESSOR FOR SUPPLY TO COAL GASIFIER
2	-5240.3	KILOWATTS	U-GAS GASIFIER NORMAL OPERATION, STEAM #2, 7 NET MAKE.
29	43214.6	KILOWATTS	29 EXPAND SHIFT (1) EXIT GAS THROUGH POWER RECOVERY TURBINE.
38	-6638.7	KILOWATTS	38 REMOVE SULFUR FROM RAW GAS BY SELEXOL SYSTEM.
52	10627.7	KILOWATTS	52 ADD STEAM VIA TURBINE; PREVENT CARBON DROP IN FUEL CELL.
56	-179574.5	KILOWATTS	56 COMPRESS COMBUSTION AIR TO FUEL CELL PRESSURE.
64	666577.7	KILOWATTS	NORMAL FUEL CELL, 1200 F INPUT SETS VOLTAGE, 1260 F EXIT.
71	-22185.6	KILOWATTS	71 CATHODE RECYCLE GAS CIRCULATION FAN.
73	286161.3	KILOWATTS	73 EXPAND CATHODE EXIT STACK GAS TO RECOVER POWER.
82	-8482.7	KILOWATTS	82 ANODE EXIT GAS BOOSTER FAN, TO FEED CATALYTIC BURNER.
83	-1795.5	KILOWATTS	83 COMPRESS COAL FEED INJECTION GAS, A SMALL RECYCLE.

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CASE RUN# 01
STEAM LIST
4 PPH #3 14152102

HUN4.1 JPL.UG.4 BASE CASE, FC REFURM, PRE-SHIFT, 13.5% CARB EX COAL TO CH4.

TABULATION OF STEAM NEEDED IN PROCESS.

2	550.0 PSIA	24072. MOLS STEAM	2 U-GAS GASIFIER NORMAL OPERATION. STREAM #2 = NET MAKE.
38	65.0 PSIA	11823. MOLS STEAM	38 REMOVE SULFUR FROM RAW GAS BY SFLEXOL SYSTEM.
52	1500.0 PSIA	8002. MOLS STEAM	52 ADD STEAM VIA TURBINE1 PREVENT CARBON DROP IN FUEL CELL.

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RUN4.1 JPL-UG.4 BASE CASE, FC REFORM, PRE-SHIFT, 13.5% CARB EX COAL TO CH4.
CASE: HUN4.1
WASTE HEAT OUT
4 FEB 83 14:52:02

TABULATION OF WASTE HEAT AVAILABLE FROM PROCESS BOILERS.

	690.0 DEGREES F.	208.77 MILLIONS BTU	15 COOL RAW GAS TO 690 DEGREES FAHRENHEIT; MAKE STEAM.
15	360.0 DEGREES F.	42.76 MILLIONS BTU	20 COOL RAW GAS; PRE-HEAT BOILER FEED WATER.
20	600.0 DEGREES F.	248.26 MILLIONS BTU	30 COOL INTER-STAGE SHIFT GAS IN WHB; RAISE STEAM.
30	880.0 DEGREES F.	434.77 MILLIONS BTU	70 COOL CATHODE RECYCLE GAS IN WHB - RAISE STEAM.
70	880.0 DEGREES F.	164.56 MILLIONS BTU	81 COOL ANODE EXIT GAS; RAISE STEAM.
81			

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CASE E

Utilize Expansion Turbines on
Low-Pressure Steam Used

TABULATION OF SHAF1 ENERGY AVAILABLE OR NEEDED.

97	-39844.7	KILOWATTS	AIR COMPRESSOR FOR SUPPLY TO CRYOGENIC OXYGEN PLANT
99	-16426.3	KILOWATTS	CRYOGENIC OXYGEN COMPRESSOR FOR SUPPLY TO COAL GASIFIER
2	-5240.3	KILOWATTS	2 U-GAS GASIFIER NORMAL OPERATION. STREAM #2 = NET MAKE.
2	16031.3	KILOWATTS	3 GET POWER BY EXPANDING GASIFIER STEAM NEEDS FROM 1500 PSI.
29	43214.6	KILOWATTS	29 EXPAND SHIFT (1) EXIT GAS THROUGH POWER RECOVERY TURBINE.
38	-6638.7	KILOWATTS	38 REMOVE SULFUR FROM RAW GAS BY SELEXOL SYSTEM.
38	19909.1	KILOWATTS	39 GET LP STEAM FOR SELEXOL REBOILER BY EXPANDING HP STEAM.
52	10627.7	KILOWATTS	52 ADD STEAM VIA TURBINE; PREVENT CARBON DROP IN FUEL CELL.
56	-179574.5	KILOWATTS	56 COMPRESS COMBUSTION AIR TO FUEL CELL PRESSURE.
64	666579.4	KILOWATTS	NORMAL FUEL CELL, 1200 F INPUT SETS VOLTAGE, 1260 F EXIT.
71	-22176.6	KILOWATTS	71 CATHODE RECYCLE GAS CIRCULATION FAN.
73	286170.9	KILOWATTS	73 EXPAND CATHODE EXIT STACK GAS TO RECOVER POWER.
82	-8482.7	KILOWATTS	82 ANODE EXIT GAS BOOSTER FAN, TO FEED CATALYTIC BURNER.
83	-1795.5	KILOWATTS	83 COMPRESS COAL FEED INJECTION GAS, A SMALL RECYCLE.

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RUN#2 JPL-UG-4 PLUS BACK-PRESSURE STEAM TURBINES TO GET LP STEAM.

CASE RUN#2
STEAM LIST
5 FEB 83 10:34:00

TABLATION OF STEAM NEEDED IN PROCESS.

2	550.0 PSIA	24072.	MULS STEAM	2	U-GAS GASIFIER NORMAL OPERATION.	STREAM #2 = NFI MAKE.
2	1500.0 PSIA	0.	MULS STEAM	3	GET POWER BY EXPANDING GASIFIER	STEAM NEEDS FROM 1500 PSI.
38	65.0 PSIA	11823.	MULS STEAM	38	REMOVE SULFUR FROM RAW GAS BY SFLEXUL SYSTEM.	
38	1500.0 PSIA	0.	MULS STEAM	39	GET LP STEAM FOR SELEXOL REBOILER	BY EXPANDING HP STEAM.
52	1500.0 PSIA	8002.	MULS STEAM	52	ADD STEAM VIA TURBINE	PREVENT CARBON DROP IN FUEL CELL.

RUN4.2
CAST WASTE HEAT OUT
5 FEB 83 10134100

JPL-UG.4 PLUS BACK-PRESSURE STEAM TURBINES TO GET LP STEAM.

TABLATION OF WASTE HEAT AVAILABLE FROM PROCESS BOILERS.

	690.0 DEGREES F.	204.77 MILLIONS BTU	15 COOL RAW GAS TO 490 DEGRES FARENHEIT; MAKE STEAM.
15	360.0 DEGREES F.	42.76 MILLIONS BTU	20 COOL RAW GAS; PHE-HEAT BOILER FEED WATER.
20	600.0 DEGREES F.	248.26 MILLIONS BTU	30 COOL INTER-STAGE SHIFT GAS IN WHB; RAISE STEAM.
30	880.0 DEGREES F.	434.58 MILLIONS BTU	70 COOL CATHODE RECYCLE GAS IN WHB - RAISE STEAM.
70	880.0 DEGREES F.	164.62 MILLIONS BTU	81 COOL ANODE EXIT GAS; RAISE STEAM.
81			

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CASE F

Employ Economizer on Final
Tail Gas to 250°F

TABULATION OF SHAFT ENERGY AVAILABLE OR NEEDED.

97	-39844.7	KILOWATTS	AIR COMPRESSOR FOR SUPPLY TO CRYOGENIC OXYGEN PLANT
99	-16426.3	KILOWATTS	CRYOGENIC OXYGEN COMPRESSOR FOR SUPPLY TO COAL GASIFIER
2	-5240.3	KILOWATTS	2 U-GAS GASIFIER NORMAL OPERATION. STEAM #2 = NET MAKE.
2	16031.3	KILOWATTS	3 GET POWER BY EXPANDING GASIFIER STEAM NETS FROM 1500 PSI.
29	43214.6	KILOWATTS	29 EXPAND SHIF (1) EXIT GAS THROUGH POWER RECOVERY TURBINE.
38	-6638.7	KILOWATTS	38 REMOVE SULFUR FROM RAW GAS BY SLEXOL SYSTEM.
38	19909.1	KILOWATTS	39 GET LP STEAM FOR SELEXOL REHEATER BY EXPANDING HP STEAM.
52	10627.7	KILOWATTS	52 ADD STEAM VIA TURBINE PREVENT CARBON DROP IN FUEL CELL.
56	-179574.5	KILOWATTS	56 COMPRESS COMBUSTION AIR TO FUEL CELL PRESSURE.
64	666579.4	KILOWATTS	NORMAL FUEL CELL. 1200 F INPUT SETS VOLTAGE, 1260 F EXIT.
71	-22176.6	KILOWATTS	71 CATHODE RECYCLE GAS CIRCULATION FAN.
73	283199.7	KILOWATTS	73 EXPAND CATHODE EXIT STACK GAS TO RECOVER POWER.
82	-8482.7	KILOWATTS	82 ANODE EXIT GAS BOOSTER FAN. TO FEED CATALYTIC BURNER.
83	-1795.5	KILOWATTS	83 COMPRESS COAL FEED INJECTION GAS. A SMALL RECYCLE.

ORIGINAL PAGE IS
OF POOR QUALITY

RUN#3 JPL.UG.4 + LP ST-TURBIS. STACK GAS ECONOMIZER TO 250 DEGREES FAHR.

CASE HUN4.3
STEAM TEST
5 FEB 83 10:40:36

TABULATION OF STEAM NEEDED IN PROCESS.

2	550.0 PSIA	24072.	MULS STEAM	2	U-GAS GASIFIER NORMAL OPERATION. STREAM #2 = NET MAKE.
2	1500.0 PSIA	0.	MULS STEAM	3	GET POWER BY EXPANDING GASIFIER STEAM NEEDS FROM 1500 PSI.
38	65.0 PSIA	11823.	MULS STEAM	38	REMOVE SULFUR FROM RAW GAS BY SELEXOL SYSTEM.
38	1500.0 PSIA	0.	MULS STEAM	39	GET LP STEAM FOR SELEXOL REBOILER BY EXPANDING HP STEAM.
52	1500.0 PSIA	8002.	MULS STEAM	52	ADD STEAM VIA TURBINE! PREVENT CARBON DROP IN FUEL CELL.

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OF POOR QUALITY

RUN4.3 JPL.UG.4 * LP ST-TURBIS, STACK GAS ECONOMIZER TO 250 DEGREES FAKH.
CASE: MUN4.3
WASTE HEAT OUT
5 PER HJ 10140136

TABULATION OF WASTE HEAT AVAILABLE FROM PROCESS BOILERS.

15	690.0	DEGREES F.	208.77	MILLIONS BTU	15	COOL RAW GAS TO 690 DEGREES FAHRENHEIT MAKE STEAM.
20	360.0	DEGREES F.	42.76	MILLIONS BTU	20	COOL RAW GAS PRE-HEAT BOILER FEED WATER.
30	600.0	DEGREES F.	248.26	MILLIONS BTU	30	COOL INTER-STAGE SHIFT GAS IN WHB1 RAISE STEAM.
70	880.0	DEGREES F.	434.58	MILLIONS BTU	70	COOL CATHODE RECYCLE GAS IN WHB - RAISE STEAM.
77	250.0	DEGREES F.	356.77	MILLIONS BTU	77	FINAL ECONOMIZER ON STACK GAS TO 250. DEG. F.
81	880.0	DEGREES F.	164.62	MILLIONS BTU	81	COOL ANODE EXIT GAS RAISE STEAM.

ORIGINAL PAGE IS
OF POOR QUALITY

CASE G

Employ Minimum Quenching
to Amorphous Carbon

TABULATION OF SHAFT ENERGY AVAILABLE OR NEEDED.

97	-39844.7 KILOWATTS	AIR COMPRESSOR FOR SUPPLY TO CRYOGENIC OXYGEN PLANT
99	-16426.3 KILOWATTS	CRYOGENIC OXYGEN COMPRESSOR FOR SUPPLY TO COAL GASIFIER
2	-5240.3 KILOWATTS	2 U-GAS GASIFIER NORMAL OPERATION, STREAM #2 = NET MAKE.
2	16031.3 KILOWATTS	3 GET POWER BY EXPANDING GASIFIER STEAM NEEDS FROM 1500 PSI.
29	44416.1 KILOWATTS	29 EXPAND SHIFT (I) EXIT GAS THROUGH POWER RECOVERY TURBINE.
38	-6499.5 KILOWATTS	38 REMOVE SULFUR FROM RAW GAS BY SELEXOL SYSTEM.
38	19772.7 KILOWATTS	39 GET LP STEAM FOR SELEXOL REBOILER BY EXPANDING HP STEAM.
52	8854.6 KILOWATTS	52 ADD STEAM VIA TURBINE; PREVENT CARBON DROP IN FUEL CELL.
56	-179574.5 KILOWATTS	56 COMPRESS COMBUSTION AIR TO FUEL CELL PRESSURE.
64	669379.3 KILOWATTS	NORMAL FUEL CELL, 1200 F INPUT SETS VOLTAGE, 1260 F EXIT.
71	-22330.4 KILOWATTS	71 CATHODE RECYCLE GAS CIRCULATION FAN.
73	280787.6 KILOWATTS	73 EXPAND CATHODE EXIT STACK GAS TO RECOVER POWER.
82	-8329.5 KILOWATTS	82 ANODE EXIT GAS BOOSTER FAN, TO FEED CATALYTIC BURNER.
83	-1782.6 KILOWATTS	83 COMPRESS COAL FEED INJECTION GAS, A SMALL RECYCLE.

ORIGINAL PAGE IS
OF POOR QUALITY

CASE RUN4.4
 STEAM LIST
 2 FEB 93 8:05:21

RUN4.4 JPL.UG.4 + LP ST-TURR'S, ECON 250, MINIMUM HUMIDITY TO AMORPHOUS LIMIT

TABULATION OF STEAM NEEDED IN PROCESS

2	550.0 PSIA	24072.	MOLS STEAM	2	U-GAS GASIFIER NORMAL OPERATION. STREAM #2 = NET MAKE.
2	1500.0 PSIA	0.	MOLS STEAM	3	GET POWER BY EXPANDING GASIFIER STEAM NEEDS FROM 1500 PSI.
38	65.0 PSIA	11742.	MOLS STEAM	38	REMOVE SULFUR FROM RAW GAS BY SELEXOL SYSTEM.
38	1500.0 PSIA	0.	MOLS STEAM	39	GET LP STEAM FOR SELEXOL REBOILER BY EXPANDING HP STEAM.
52	1500.0 PSIA	6667.	MOLS STEAM	52	ADD STEAM VIA TURBINE; PREVENT CARBON DROP IN FUEL CELL.

ORIGINAL PAGE IS
 OF POOR QUALITY

CASE RUN4.4
WASTE HEAT OUT
2 FEB 83 8:05:21

JPL.UG.4 + LP ST-TURB'S, ECON 25D, MINIMUM HUMIDITY TO AMORPHOUS LIMIT

TABULATION OF WASTE HEAT AVAILABLE FROM PROCESS BOILERS.

15	690.0 DEGREES F.	203.00 MILLIONS BTU	15 COOL RAW GAS TO 590 DEGREES FAHRENHEIT; MAKE STEAM.
20	360.0 DEGREES F.	41.57 MILLIONS BTU	20 COOL RAW GAS; PRE-HEAT BOILER FEED WATER.
30	600.0 DEGREES F.	291.28 MILLIONS BTU	30 COOL INTER-STAGE SHIFT GAS IN WHR; RAISE STEAM.
70	890.0 DEGREES F.	439.62 MILLIONS BTU	70 COOL CATHODE RECYCLE GAS IN WHB - RAISE STEAM.
77	250.0 DEGREES F.	350.50 MILLIONS BTU	77 FINAL ECONOMIZER ON STACK GAS TO 250. DEG. F.
81	890.0 DEGREES F.	165.37 MILLIONS BTU	81 COOL ANODE EXIT GAS; RAISE STEAM.

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OF POOR QUALITY

CASE H

Change Gasifier to
More Optimistic Design Point

TABULATION OF SHAFT ENERGY AVAILABLE OR NEEDED.

97	-38419.6 KILOWATTS	AIP COMPRESSOR FOR SUPPLY TO CRYOGENIC OXYGEN PLANT
99	-16762.0 KILOWATTS	CRYOGENIC OXYGEN COMPRESSOR FOR SUPPLY TO COAL GASIFIER
2	-5240.3 KILOWATTS	2 SUPER U-GAS. 600#, LO-STM, LO-OK. STREAM #2 = NET MAKE.
2	10197.4 KILOWATTS	3 GET POWER BY EXPANDING GASIFIER STEAM NEEDS FROM 1500 PSI.
29	46653.4 KILOWATTS	29 EXPAND SHIFT (1) EXIT GAS THROUGH POWER RECOVERY TURBINE.
38	-6438.0 KILOWATTS	38 REMOVE SULFUR FROM RAW GAS BY SELEXOL SYSTEM.
38	19872.1 KILOWATTS	39 GET LP STEAM FOR SELEXOL REBOILER BY EXPANDI/G HP STEAM.
52	9077.1 KILOWATTS	52 ADD STEAM VIA TURBINE; PREVENT CARBON DROP IN FUEL CELL.
56	-179574.5 KILOWATTS	56 COMPRESS COMBUSTION AIR TO FUEL CELL PRESSURE.
64	678920.7 KILOWATTS	NORMAL FUEL CELL, 1200 F INPUT SETS VOLTAGE, 17/60 F EXIT.
71	-23499.9 KILOWATTS	71 CATHODE RECYCLE GAS CIRCULATION FAN.
73	280989.2 KILOWATTS	73 EXPAND CATHODE EXIT STACK GAS TO RECOVER POWER.
82	-8410.2 KILOWATTS	82 ANODE EXIT GAS BOOSTER FAN, TO FEED CATALYTIC BURNER.
83	-2059.1 KILOWATTS	83 COMPRESS COAL FEED INJECTION GAS, A SMALL RECYCLE.

ORIGINAL PAGE IS
OF POOR QUALITY

CASE RUNS
STEAM LIST
2 FEB 83 7:55:43

RUNS 600PSI U-GAS, LP ST-TURB'S, ECON 250, MIN. H2O TO AMORF LIM., FC 1260

TABULATION OF STEAM NEEDED IN PROCESS.

2	650.0 PSIA	18054.	MOLS STEAM	2	SUPER U-GAS.	600#.	LO-STM.	LO-OK.	STREAM #2 = NET MAKE.
2	1500.0 PSIA	0.	MOLS STEAM	3	GET POWER BY EXPANDING GASIFIER	STEAM NEEDS FROM 1500 PSI.			
38	65.0 PSIA	11801.	MOLS STEAM	38	REMOVE SULFUR FROM RAW GAS BY SELEXOL SYSTEM.				
38	1500.0 PSIA	0.	MOLS STEAM	39	GET LP STEAM FOR SELEXOL REBOILER BY EXPANDING HP STEAM.				
52	1500.0 PSIA	6835.	MOLS STEAM	52	ADD STEAM VIA TURBINE; PREVENT CARBON DROP IN FUEL CELL.				

ORIGINAL PAGE IS
OF POOR QUALITY

CASE RUNS
WASTE HEAT OUT
2 FEB 83 7:55:43

RUNS 600PSI U-GAS, LP ST-TURB'S, ECON 250, MIN. H2O TO AMORF LIM., FC 1260

TABULATION OF WASTE HEAT AVAILABLE FROM PROCESS BOILERS.

15	690.0 DEGREES F.	114.65 MILLIONS BTU	15 COOL RAW GAS TO 690 DEGREES FAHRENHEIT; MAKE STEAM.
20	370.0 DEGREES F.	43.67 MILLIONS BTU	20 COOL RAW GAS; PRE-HEAT BOILER FEED WATER.
30	600.0 DEGREES F.	143.09 MILLIONS BTU	30 COOL INTER-STAGE SHIFT GAS IN WHB; RAISE STEAM.
70	880.0 DEGREES F.	475.22 MILLIONS BTU	70 COOL CATHODE RECYCLE GAS IN WHB - RAISE STEAM.
77	250.0 DEGREES F.	357.06 MILLIONS BTU	77 FINAL ECONOMIZER ON STACK GAS TO 250. DEG. F.
81	880.0 DEGREES F.	160.42 MILLIONS BTU	81 COOL ANODE EXIT GAS; RAISE STEAM.

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OF POOR QUALITY

CASE I

As Case G, but
Change Quench System From
Liquid Water to Steam

TABULATION OF SHAFT ENERGY AVAILABLE OR NEEDED.

97	-38418.6 KILOWATTS	AIR COMPRESSOR FOR SUPPLY TO CRYOGENIC OXYGEN PLANT
99	-16762.0 KILOWATTS	CRYOGENIC OXYGEN COMPRESSOR FOR SUPPLY TO COAL GASIFIER
2	-5240.3 KILOWATTS	2 SUPER U-GAS. 600#, LO-STM, LO-CX. STREAM #2 = NET MAKE.
2	10197.4 KILOWATTS	3 GET POWER BY EXPANDING GASIFIER STEAM NEEDS FROM 1500 PSI.
13	5274.5 KILOWATTS	13 STEAM INJECTION TO PREVENT CARBON DEPOSITION. (VIA TURB)
29	5399.3 KILOWATTS	29 EXPAND SHIFT (I) EXIT GAS THROUGH POWER RECOVERY TURBINE.
38	-6438.0 KILOWATTS	38 REMOVE SULFUR FROM RAW GAS BY SELEXOL SYSTEM.
38	19872.1 KILOWATTS	39 GET LP STEAM FOR SELEXOL REBOILER BY EXPANDING HP STEAM.
52	9077.1 KILOWATTS	52 ADD STEAM VIA TURBINE; PREVENT CARBON DROP IN FUEL CELL.
56	-179574.5 KILOWATTS	56 COMPRESS COMBUSTION AIR TO FUEL CELL PRESSURE.
64	678792.7 KILOWATTS	NORMAL FUEL CELL, 1200 F INPUT SETS VOLTAGE, 1260 F EXIT.
71	-23652.6 KILOWATTS	71 CATHODE RECYCLE GAS CIRCULATION FAN.
73	280991.1 KILOWATTS	73 EXPAND CATHODE EXIT STACK GAS TO RECOVER POWER.
82	-8410.2 KILOWATTS	82 ANODE EXIT GAS BOOSTER FAN, TO FEED CATALYTIC BURNER.
83	-2059.1 KILOWATTS	83 COMPRESS COAL FEED INJECTION GAS, A SMALL RECYCLE.

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OF POOR QUALITY

CASE RUNSA
STEAM LIST
2 FEB 93 11:09:43

RUNSA STEAM INJECTION IN PLACE OF RAW GAS QUENCH. OTHERWISE SAME AS CASE 5

TABULATION OF STEAM NEEDED IN PROCESS.

2	650.0 PSIA	18054.	MOLS STEAM	2	SUPER U-GAS.	600#,	LO-STM,	LO-OK.	STREAM #2 = NET MAKE.
2	1500.0 PSIA	0.	MOLS STEAM	3	GET POWER BY EXPANDING GASIFIER	STEAM NEEDS FROM 1500 PSI.			
13	1500.0 PSIA	8960.	MOLS STEAM	13	STEAM INJECTION TO PREVENT CARBON DEPOSITION. (VIA TURB)				
38	65.0 PSIA	11301.	MOLS STEAM	38	REMOVE SULFUR FROM RAW GAS BY SELEXOL SYSTEM.				
38	1500.0 PSIA	0.	MOLS STEAM	39	GET LP STEAM FOR SELEXOL REBOILER BY EXPANDING HP STEAM.				
52	1500.0 PSIA	6835.	MOLS STEAM	52	ADD STEAM VIA TURBINE; PREVENT CARBON DROP IN FUEL CELL.				

ORIGINAL PAGE IS
OF POOR QUALITY

CASE RUNSA
WASTE HEAT OUT
2 FEB 83 11:08:43

RUNSA STEAM INJECTION IN PLACE OF RAW GAS QUENCH. OTHERWISE SAME AS CASE 5

TABULATION OF WASTE HEAT AVAILABLE FROM PROCESS BOILERS.

15	690.0 DEGREES F.	244.40 MILLIONS BTU	15 COOL RAW GAS TO 690 DEGREES FAHRENHEIT; MAKE STEAM.
20	370.0 DEGREES F.	43.67 MILLIONS BTU	20 COOL RAW GAS; PRE-HEAT BOILER FEED WATER.
30	600.0 DEGREES F.	264.41 MILLIONS BTU	30 COOL INTER-STAGE SHIFT GAS IN WH3; RAISE STEAM.
70	880.0 DEGREES F.	480.66 MILLIONS BTU	70 COOL CATHODE RECYCLE GAS IN WHB - RAISE STEAM.
77	250.0 DEGREES F.	397.07 MILLIONS BTU	77 FINAL ECONOMIZER ON STACK GAS TO 250. DEG. F.
81	880.0 DEGREES F.	155.57 MILLIONS BTU	81 COOL ANODE EXIT GAS; RAISE STEAM.

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OF POOR QUALITY

CASE J

Remove More CO₂ With H₂S

RUNG LIKE RUN5A BUT REMOVE MORE CO2 @ #38 & RETIRE TURBINE #52.

TABULATION OF SHAFT ENERGY AVAILABLE OR NEEDED.

97	-38418.6	KILOWATTS	AIR COMPRESSOR FOR SUPPLY TO CRYOGENIC OXYGEN PLANT
99	-16762.0	KILOWATTS	CRYOGENIC OXYGEN COMPRESSOR FOR SUPPLY TO COAL GASIFIER
2	-5249.3	KILOWATTS	2 SUPER U-GAS. 600°, LO-STM, LO-OK. STREAM #2 = NET MAKE.
2	10197.4	KILOWATTS	3 GET POWER BY EXPANDING GASIFIER STEAM NEEDS FROM 1500 PSI.
13	5134.7	KILOWATTS	13 STEAM INJECTION TO PREVENT CARBON DEPOSITION. (VIA TURB)
29	53358.5	KILOWATTS	29 EXPAND SHIFT (1) EXIT GAS THROUGH POWER RECOVERY TURBINE.
38	-1508.0	KILOWATTS	38 REMOVE DILUTE SULFUR FROM RAW GAS BY SELEXOL SYSTEM.
38	19627.9	KILOWATTS	39 GET LP STEAM FOR SELEXOL REBOILER BY EXPANDING HP STEAM.
52	4070.8	KILOWATTS	52 ADD STEAM VIA TURBINE; PREVENT CARBON DROP IN FUEL CELL.
56	-179574.5	KILOWATTS	56 COMPRESS COMBUSTION AIR TO FUEL CELL PRESSURE.
64	675332.7	KILOWATTS	NORMAL FUEL CELL, 1200 F INPUT SETS VOLTAGE, 1260 F EXIT.
71	-24471.9	KILOWATTS	71 CATHODE RECYCLE GAS CIRCULATION FAN.
73	263416.9	KILOWATTS	73 EXPAND CATHODE EXIT STACK GAS TO RECOVER POWER.
82	-7256.6	KILOWATTS	82 ANODE EXIT GAS BOOSTER FAN, TO FEED CATALYTIC BURNER.
83	-1705.0	KILOWATTS	83 COMPRESS COAL FEED INJECTION GAS, A SMALL RECYCLE.

ORIGINAL PAGE IS
OF POOR QUALITY

CASE RUN6
STEAM LIST
2 FEB 83 15:58:41

RUN6 LIKE RUN5A BUT REMOVE MORE CO2 @ #39 & RETIRE TURBINE #52.

TABULATION OF STEAM NEEDED IN PROCESS.

2	650.0 PSIA	18954.	MOLS STEAM	2 SUPER U-GAS. 600W, LO-STM, LO-OX. STREAM #2 = NET MAKE.
2	1500.0 PSIA	0.	MOLS STEAM	3 GET POWER BY EXPANDING GASIFIER STEAM NEEDS FROM 1500 PSI.
13	1500.0 PSIA	8703.	MOLS STEAM	13 STEAM INJECTION TO PREVENT CARBON DEPOSITION. (VIA TURB)
38	65.0 PSIA	11656.	MOLS STEAM	38 REMOVE DILUTE SULFUR FROM RAW GAS BY SELEXOL SYSTEM.
38	1500.0 PSIA	0.	MOLS STEAM	39 GET LP STEAM FOR SELEXOL REBOILER BY EXPANDING HP STEAM.
52	1500.0 PSIA	3065.	MOLS STEAM	52 ADD STEAM VIA TURBINE; PREVENT CARBON DROP IN FUEL CELL.

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OF POOR QUALITY

CASE RUN6
WASTE HEAT OUT
2 FEB 83 15:58:41

RUN6 LIKE RUN5A BUT REMOVE MORE CO2 2 #18 & RETIRE TURBINE #52.

TABULATION OF WASTE HEAT AVAILABLE FROM PROCESS BOILERS.

15	690.C DEGREES F.	239.96 MILLIONS BTU	15 COOL RAW GAS TO 690 DEGREES FARENHEIT; MAKE STEAM.
20	370.0 DEGREES F.	42.90 MILLIONS BTU	20 COOL RAW GAS; PRE-HEAT BOILER FEED WATER.
30	600.0 DEGREES F.	262.27 MILLIONS BTU	30 COOL INTER-STAGE SHIFT GAS IN WMB; RAISE STEAM.
70	880.0 DEGREES F.	490.66 MILLIONS BTU	70 COOL CATHODE RECYCLE GAS IN MMB - RAISE STEAM.
77	250.C DEGREES F.	295.45 MILLIONS BTU	77 FINAL ECONOMIZER ON STACK GAS TO 250. DEG. F.
81	880.0 DEGREES F.	193.30 MILLIONS BTU	91 COOL ANODE EXIT GAS; RAISE STEAM.

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OF POOR QUALITY

CASE K

As Case I, but Burn Part of
Anode Exhaust into Cathode Exhaust

RUN7 90X TWO BURNS, 600# U-GAS, LP ST-TURB'S, ECON 250, MIN. H2O TO AMORF LIM.

TABULATION OF SHAFT ENERGY AVAILABLE OR NEEDED.

97	-38418.6 KILOWATTS	AIR COMPRESSOR FOR SUPPLY TO CRYOGENIC OXYGEN PLANT
99	-16762.0 KILOWATTS	CRYOGENIC OXYGEN COMPRESSOR FOR SUPPLY TO COAL GASIFIER
2	-5240.3 KILOWATTS	2 SUPER U-GAS, 600#, LO-STM, LO-OX, STREAM #2 = NET MAKE.
2	10197.4 KILOWATTS	3 GET POWER BY EXPANDING GASIFIED STEAM NEEDS FROM 1500 PSI.
13	5514.3 KILOWATTS	13 INJECT STEAM VIA POWER RECOVERY TURBINE TO STABILIZE GAS.
29	53957.2 KILOWATTS	29 EXPAND SHIT (I) EXIT GAS THROUGH POWER RECOVERY TURBINE.
38	-6439.0 KILOWATTS	38 REMOVE SULFUR FROM RAW GAS BY SELEXOL SYSTEM.
38	19572.1 KILOWATTS	39 GET LP STEAM FOR SELEXOL REBOILER BY EXPANDING HP STEAM.
52	9077.1 KILOWATTS	52 ADD STEAM VIA TURBINE; PREVENT CARBON DROP IN FUEL CELL.
56	-179574.5 KILOWATTS	56 COMPRESS COMBUSTION AIR TO FUEL CELL PRESSURE.
64	673685.6 KILOWATTS	NORMAL FUEL CELL, 1200 F INPUT SETS VOLTAGE, 1200 F EXIT.
71	-23577.7 KILOWATTS	71 CATHODE RECYCLE GAS CIRCULATION FAN.
73	276521.7 KILOWATTS	73 EXPAND CATHODE EXIT STACK GAS TO RECOVER POWER.
83	-8107.4 KILOWATTS	83 ANODE EXIT GAS BOOSTER FAN, TO CATH. FEED CAT. BURNER.
84	-2059.1 KILOWATTS	84 COMPRESS COAL FEED INJECTION GAS, A SMALL RECYCLE.

ORIGINAL PAGE IS
OF POOR QUALITY

CASE RUN7 90X
 STEAM LIST
 3 FEB 93 11:20:21

RUN7 90X TWO BURNS, 600# U-GAS, LP ST-TURB'S, ECON 250, MIN. H2O TO AMORF LIM.

TABULATION OF STEAM NEEDED IN PROCESS.

2	650.0 PSIA	18054. MOLS STEAM	2 SUPER U-GAS. 600#, LO-STM, LO-OX. STREAM #2 = NET MAKE.
2	1500.0 PSIA	0. MOLS STEAM	3 GET POWER BY EXPANDING GASIFIER STEAM NEEDS FROM 1500 PSI.
13	1500.0 PSIA	2940. MOLS STEAM	13 INJECT STEAM VIA POWER RECOVERY TURBINE TO STABILIZE GAS.
38	65.0 PSIA	11801. MOLS STEAM	38 REMOVE SULFUR FROM RAW GAS BY SELEXOL SYSTEM.
38	1500.0 PSIA	0. MOLS STEAM	39 GET LP STEAM FOR SELEXOL REBOILER BY EXPANDING HP STEAM.
52	1500.0 PSIA	6935. MOLS STEAM	52 ADD STEAM VIA TURBINE; PREVENT CARBON DROP IN FUEL CELL.

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CASE RUN7 90%
WASTE HEAT OUT
3 FEB 83 11:20:21

RUN7 90% TWO BURNS, 600# U-GAS, LP ST-TURR'S, ECON 250, MIN. H2O TO AMORF LIM.

TABULATION OF WASTE HEAT AVAILABLE FROM PROCESS BOILERS.

15	690.C DEGREES F.	244.40 MILLIONS BTU	15 COOL RAW GAS TO 690 DEGREES FAHRENHEIT; MAKE STEAM.
20	370.0 DEGREES F.	43.67 MILLIONS BTU	20 COOL RAW GAS; PRE-HEAT BOILER FEED WATER.
30	600.C DEGREES F.	263.72 MILLIONS BTU	30 COOL INTER-STAGE SHIFT GAS IN WHB; RAISE STEAM.
70	830.0 DEGREES F.	464.29 MILLIONS BTU	70 COOL CATHODE RECYCLE GAS IN WHB - RAISE STEAM.
77	250.0 DEGREES F.	490.59 MILLIONS BTU	77 FINAL ECONOMIZER ON STACK GAS TO 250. DEG. F.
30	230.0 DEGREES F.	160.15 MILLIONS BTU	30 COOL ANODE EXIT GAS; RAISE STEAM.

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OF POOR QUALITY

CASE L

As Case I, Employ Low-Pressure
Drops Throughout the System

TABULATION OF SHAFT ENERGY AVAILABLE OR NEEDED.

97	-38418.6 KILOWATTS	AIR COMPRESSOR FOR SUPPLY TO CRYOGENIC OXYGEN PLANT
99	-16762.0 KILOWATTS	CRYOGENIC OXYGEN COMPRESSOR FOR SUPPLY TO COAL GASIFIER
2	-5240.3 KILOWATTS	2 SUPER U-GAS, 600#, LO-STM, LO-OX, STREAM #2 = NET MAKE.
2	10197.4 KILOWATTS	3 GET POWER BY EXPANDING GASIFIER STEAM NEEDS FROM 1500 PSI.
13	5258.2 KILOWATTS	13 STEAM INJECTION TO PREVENT CARBON DEPOSITION. (VIA TURB)
29	60403.0 KILOWATTS	29 EXPAND SHIFT (1) EXIT GAS THROUGH POWER RECOVERY TURBINE.
38	-6440.2 KILOWATTS	38 REMOVE SULFUR FROM RAW GAS BY SELEXOL SYSTEM.
38	20448.0 KILOWATTS	38 GET LP STEAM FOR SELEXOL REBOILER BY EXPANDING HP STEAM.
52	9100.8 KILOWATTS	52 ADD STEAM VIA TURBINE; PREVENT CARBON DROP IN FUEL CELL.
56	-175065.9 KILOWATTS	56 COMPRESS COMBUSTION AIR TO FUEL CELL PRESSURE.
64	679073.2 KILOWATTS	NORVAL FUEL CELL, 1200 F INPUT SETS VOLTAGE, 1260 F EXIT.
71	-11912.0 KILOWATTS	71 CATHODE RECYCLE GAS CIRCULATION FAN.
73	282566.2 KILOWATTS	73 EXPAND CATHODE EXIT STACK GAS TO RECOVER POWER.
82	-4077.0 KILOWATTS	82 ANODE EXIT GAS BOOSTER FAN, TO FEED CATALYTIC BURNER.
84	-2115.7 KILOWATTS	84 COMPRESS COAL FEED INJECTION GAS, A SMALL RECYCLE.

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RUNB LO-OP, 600# U-GAS, LP ST-TURB'S, STM-STELE W. MIN. 420, FC 1250

CASE RUNR
STEAM LIST
3 FEB 83 14:46:29

TABULATION OF STEAM NEEDED IN PROCESS.

2	650.0 PSIA	19054.	MOLS STEAM	2	SUPER U-GAS.	600#	LO-STM,	LO-OK.	STREAM #2	=	NET MAKE.
2	1500.0 PSIA	0.	MOLS STEAM	3	GET POWER BY EXPANDING GASIFIER STEAM NEEDS FROM 1500 PSI.						
13	1500.0 PSIA	8942.	MOLS STEAM	13	STEAM INJECTION TO PREVENT CARBON DEPOSITION. (VIA TURB)						
38	65.0 PSIA	12143.	MOLS STEAM	38	REMOVE SULFUR FROM RAW GAS BY SELEXOL SYSTEM.						
38	1500.0 PSIA	0.	MOLS STEAM	39	GET LP STEAM FOR SELEXOL REBOILER BY EXPANDING HP STEAM.						
52	1500.0 PSIA	6234.	MOLS STEAM	52	ADD STEAM VIA TURBINE; PREVENT CARBON DROP IN FUEL CELL.						

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CASE RUN8
 WASTE HEAT OUT
 3 FEB 83 14:46:29

LO-DP, 600# U-GAS, LP ST-TURB'S, STM-STBLE W. MIN. H2O, FC 1260

TABULATION OF WASTE HEAT AVAILABLE FROM PROCESS BOILERS.

15	690.0 DEGREES F.	244.41 MILLIONS BTU	15 COOL RAW GAS TO 690 DEGREES FAHRENHEIT; MAKE STEAM.
20	370.0 DEGREES F.	43.68 MILLIONS BTU	20 COOL RAW GAS; PRE-HEAT BOILER FEED WATER.
30	600.0 DEGREES F.	242.95 MILLIONS BTU	30 COOL INTER-STAGE SHIFT GAS IN WH9; RAISE STEAM.
70	880.0 DEGREES F.	426.55 MILLIONS BTU	70 COOL CATHODE RECYCLE GAS IN WH8 - RAISE STEAM.
77	250.0 DEGREES F.	343.64 MILLIONS BTU	77 FINAL ECONOMIZER ON STACK GAS TO 250. DEG. F.
81	890.0 DEGREES F.	155.25 MILLIONS BTU	81 COOL ANODE EXIT GAS; RAISE STEAM.

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Case M

Increase Fuel Cell Exhaust
to 1300°F

TABULATION OF SHAFT ENERGY AVAILABLE OR NEEDED.

97	-38419.6	KILOWATTS	AIR COMPRESSOR FOR SUPPLY TO CRYOGENIC OXYGEN PLANT
99	-16762.0	KILOWATTS	CRYOGENIC OXYGEN COMPRESSOR FOR SUPPLY TO COAL GASIFIER
2	-5260.3	KILOWATTS	2 SUPER U-GAS. 600#, LO-STM, LO-OK. STREAM #2 = NET MAKE.
3	10197.4	KILOWATTS	3 GET POWER BY EXPANDING GASIFIER STEAM NEEDS FROM 1500 PSI.
13	5253.2	KILOWATTS	13 STEAM INJECTION TO PREVENT CARBON DEPOSITION. (VIA TURB)
29	60403.0	KILOWATTS	29 EXPAND SHIST (I) EXIT GAS THROUGH POWER RECOVERY TURBINE.
38	-6440.2	KILOWATTS	38 REMOVE SULFUR FROM RAW GAS BY SELEXOL SYSTEM.
39	20429.0	KILOWATTS	39 GET LP STEAM FOR SELEXOL REBOILER BY EXPANDING HP STEAM.
52	9100.8	KILOWATTS	52 ADD STEAM VIA TURBINE; PREVENT CARBON DROP IN FUEL CELL.
56	-175065.9	KILOWATTS	56 COMPRESS COMBUSTION AIR TO FUEL CELL PRESSURE.
64	680214.5	KILOWATTS	HI-T FUEL CELL, 1200 F INPUT SETS VOLTAGE, 1300 F EXIT.
71	-9057.1	KILOWATTS	71 CATHODE RECYCLE GAS CIRCULATION FAN.
73	289139.3	KILOWATTS	73 EXPAND CATHODE EXIT STACK GAS TO RECOVER POWER.
82	-4077.0	KILOWATTS	82 ANODE EXIT GAS BOOSTER FAN, TO FEED CATALYTIC BURNER.
84	-2115.7	KILOWATTS	84 COMPRESS COAL FEED INJECTION GAS, A SMALL RECYCLE.

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CASE RUN9
STEAM LIST
3 FEB 83 15:26:42

LO-DP, 600# U-GAS, LP ST-TURB'S, STM-ST9LE W. MIN. H2O, FC 1300 EXIT

TABULATION OF STEAM NEEDED IN PROCESS.

2	650.0 PSIA	19054.	MOLS STEAM	2 SUPER U-GAS. 600#, LO-STM, LC-OK. STREAM #2 = NET MAKE.	
2	1500.0 PSIA	0.	MOLS STEAM	3 GET POWER BY EXPANDING GASIFIER STEAM NEEDS FROM 1500 PSI.	
13	1500.0 PSIA	8942.	MOLS STEAM	13 STEAM INJECTION TO PREVENT CARBON DEPOSITION. (VIA TURB)	
38	65.0 PSIA	12143.	MOLS STEAM	38 REMOVE SULFUR FROM RAW GAS BY SELEXOL SYSTEM.	
38	1500.0 PSIA	J.	MOLS STEAM	39 GET LP STEAM FOR SELEXOL REPOILER BY EXPANDING HP STEAM.	
52	1500.0 PSIA	6834.	MOLS STEAM	52 ADD STEAM VIA TURBINE; PREVENT CARBON DROP IN FUEL CELL.	

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TABULATION OF WASTE HEAT AVAILABLE FROM PROCESS BOILERS.

15	690.0 DEGREES F.	244.41 MILLIONS BTU	15 COOL RAW GAS TO 690 DEGREES FAHRENHEIT; MAKE STEAM.
20	370.0 DEGREES F.	43.63 MILLIONS BTU	20 COOL RAW GAS; PRE-HEAT BOILER FEED WATER.
30	600.0 DEGREES F.	242.95 MILLIONS BTU	30 COOL INTER-STAGE SHIFT GAS IN WHB; RAISE STEAM.
70	980.0 DEGREES F.	303.22 MILLIONS BTU	70 COOL CATHODE RECYCLE GAS IN WHB - RAISE STEAM.
77	250.0 DEGREES F.	393.72 MILLIONS BTU	77 FINAL ECONOMIZER ON STACK GAS TO 250. DEG. F.
81	830.0 DEGREES F.	196.61 MILLIONS BTU	81 COOL ANODE EXIT GAS; RAISE STEAM.

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