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IGT Project No. 65060

SYSTEMS ANALYSIS OF ELECTRICITY PRODUCTION FROM COAL USING FUEL CELLS

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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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Final Report

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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

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Table ES-1. SUMMARY OF SYSTEMS ANALYZED

System No.	1	2	3	4	5
Gasifier Type Methane Make	Entrained Nil	Entrained Nil	Fluidized Bed Minimal	Fluidized Bed Minimal	Fluidized Bed 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	0.72	0.76	0.75	0.87	0.88
Total	8.5-	8.7+	8.7+	8.3-	8.2-

System No.	6	7	8	9
Gasifier Type Methane Make	Fluidized Bed Moderate	Fluidized Bed Moderate	Fluidized Bed Moderate	Fluidized Bed 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	0.86	0.81	0.86	0.81
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 definative 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 characterize 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 H_2S 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 ($H_2 + CO + CH_4$) 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

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* 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.

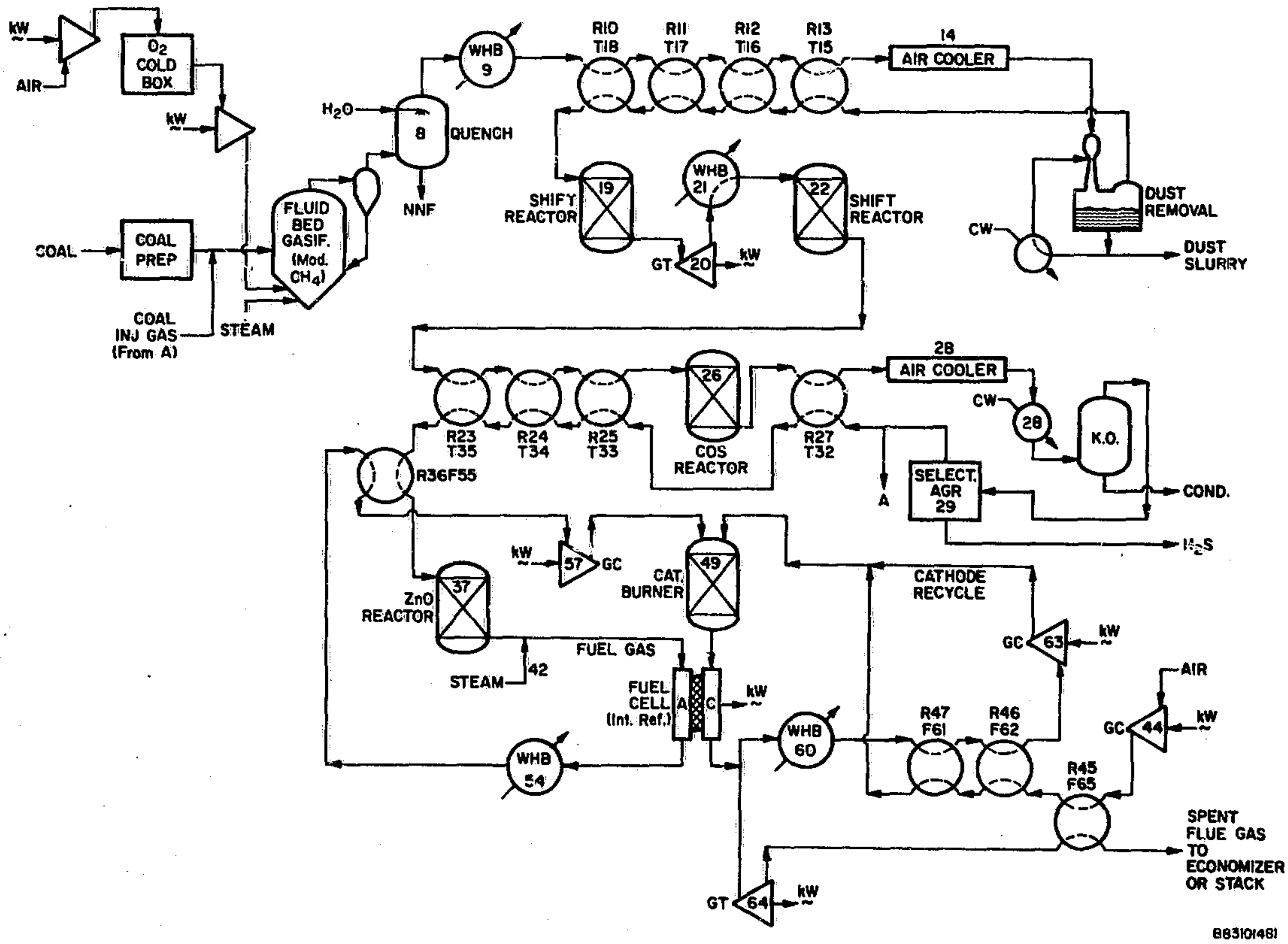

Figure 1. PROCESS FLOW DIAGRAM OF TYPICAL COAL GASIFICATION/MCFC POWER GENERATION SYSTEM

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 inputed), 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 preceeding 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

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[*] 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 data[3] (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) H_2S from the process gas, 2) CO_2 in accordance with an assumed H_2S 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 H_2S 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 H_2S-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 data[4] 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

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* 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 countercurrent 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 (H_2 + CO + 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

Fuel Processing Unit	Cost Base	Scaling Capacity Factor	Scaling Exponent
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 that 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 CO_2 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[7], 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.

Table 2. EFFICIENCY COMPARISONS OF ALTERNATIVE SYSTEMS

Case	Reference (Appendix Ca)	Features	Input 10^9 Btu/h	Power Output Net Mech	 Fuel Cell	 Steam Cycle	 Net	Eff
				MW				
1	I	Base Case, Entrained Gasifier Unstable Raw Gas - No Shift, Countercurrent 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	89	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

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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, CO_2 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
					$\$10^6$				
Component									
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	12.1	12.6	13.4	12.2	12.0	11.8	--	11.6	1.0
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	25.6	25.6	25.6	25.6	25.6	25.6	25.6	25.6	25.6
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
Allowance for Funds During Construction (AFDC)									
AFDC = TPC (0.0373)	35.1	36.0	35.0	33.5	33.1	32.7	29.3	33.6	28.8
Prepaid Royalties									
= 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
Preproduction or Start-Up Costs	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									
Inventory Capital									
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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Table 5. ANNUAL OPERATING COSTS FOR PRODUCING 675 MW POWER FROM COMBINED COAL GASIFICATION/FUEL CELL OPERATION

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Case No.*	1	2	3	4	5	6	7	8	9
	$\$10^6$								
Fixed Operating Costs									
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	3.19	3.33	3.21	3.06	3.02	3.00	2.67	3.09	2.69
Subtotal	24.14	25.44	24.37	22.99	22.63	22.41	19.31	23.19	19.52
Annual Variable Operating Costs									
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)	0.33	0.38	0.35	0.33	0.33	0.33	--	0.43	0.33
Subtotal	50.50	51.19	55.38	47.40	46.45	45.01	42.31	44.97	42.60
Levelized Revenue Required for First 10 Years									
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
Variable Operating Costs									
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	27.6	29.1	29.0	33.6	34.0	33.0	31.2	32.9	31.2
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	3.8435								
Cost of Electricity, \$/kWh	0.085	0.087	0.087	0.083	0.082	0.080	0.072	0.082	0.072

* 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 employes 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 CO_2 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 CO_2 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 CO_2 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 t 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 CO_2 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.

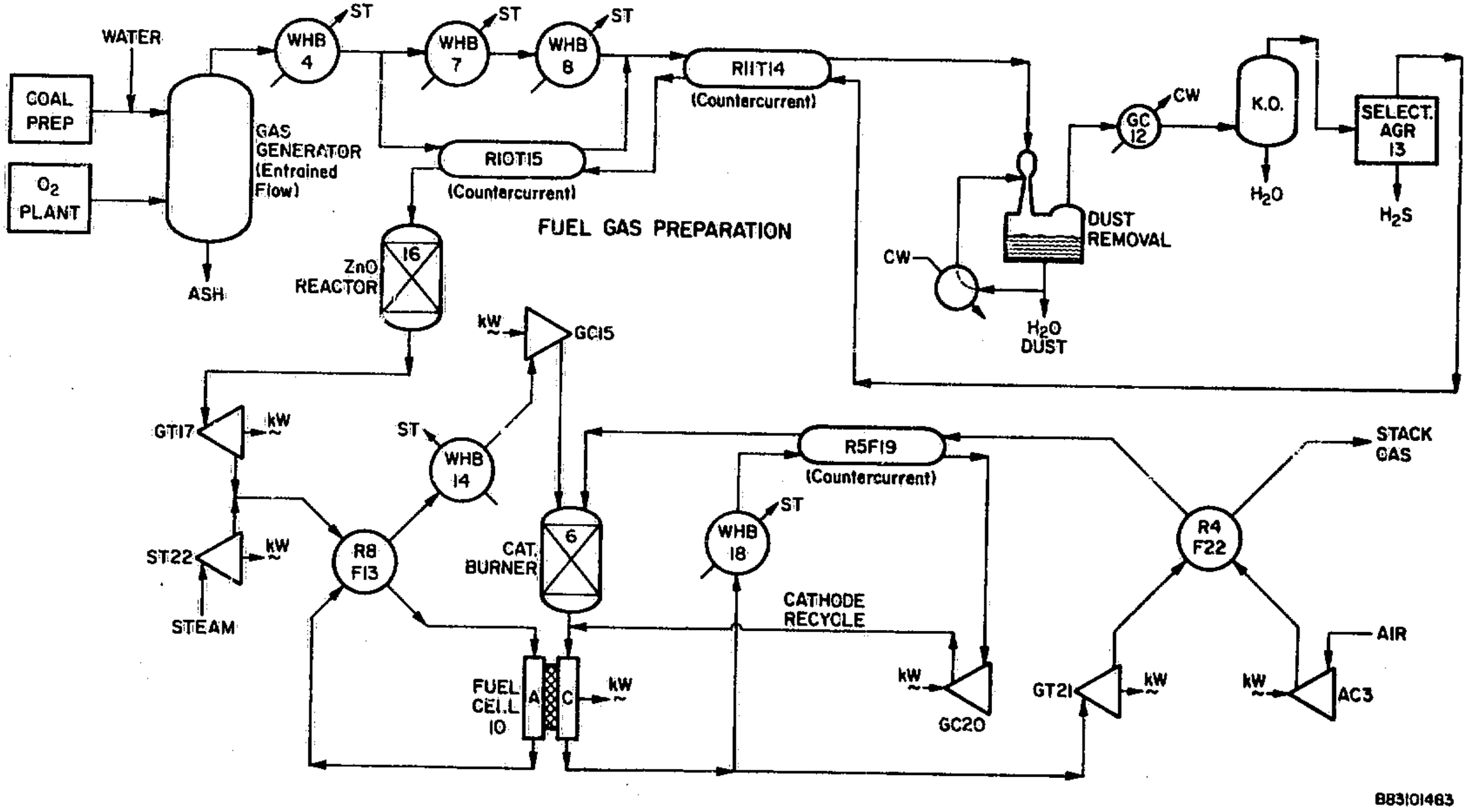

Figure 2. PROCESS FLOW DIAGRAM OF BASE CASE

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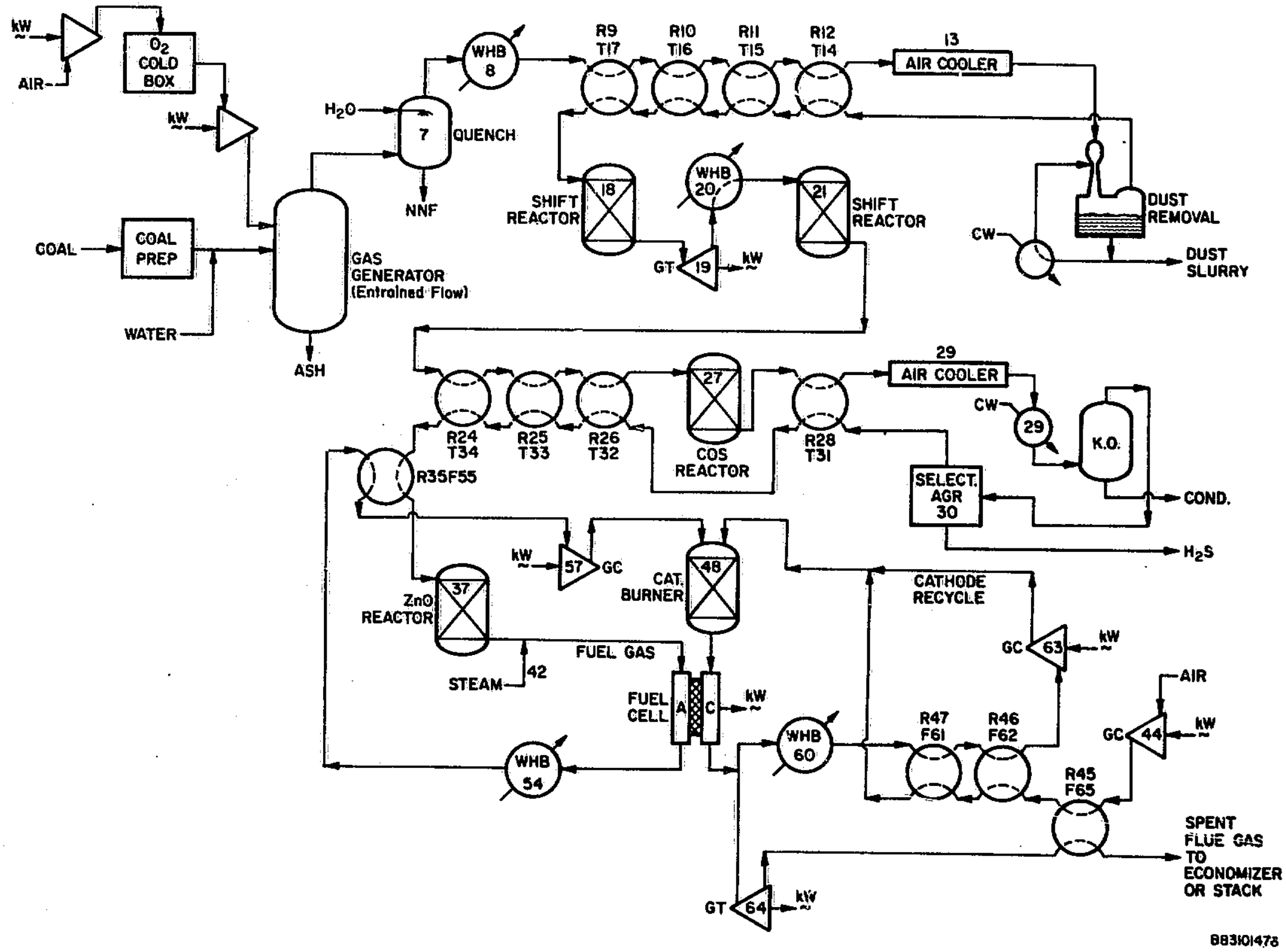

Figure 3. PROCESS FLOW DIAGRAM OF BASE CASE AFTER REDESIGN FOR OPERABILITY

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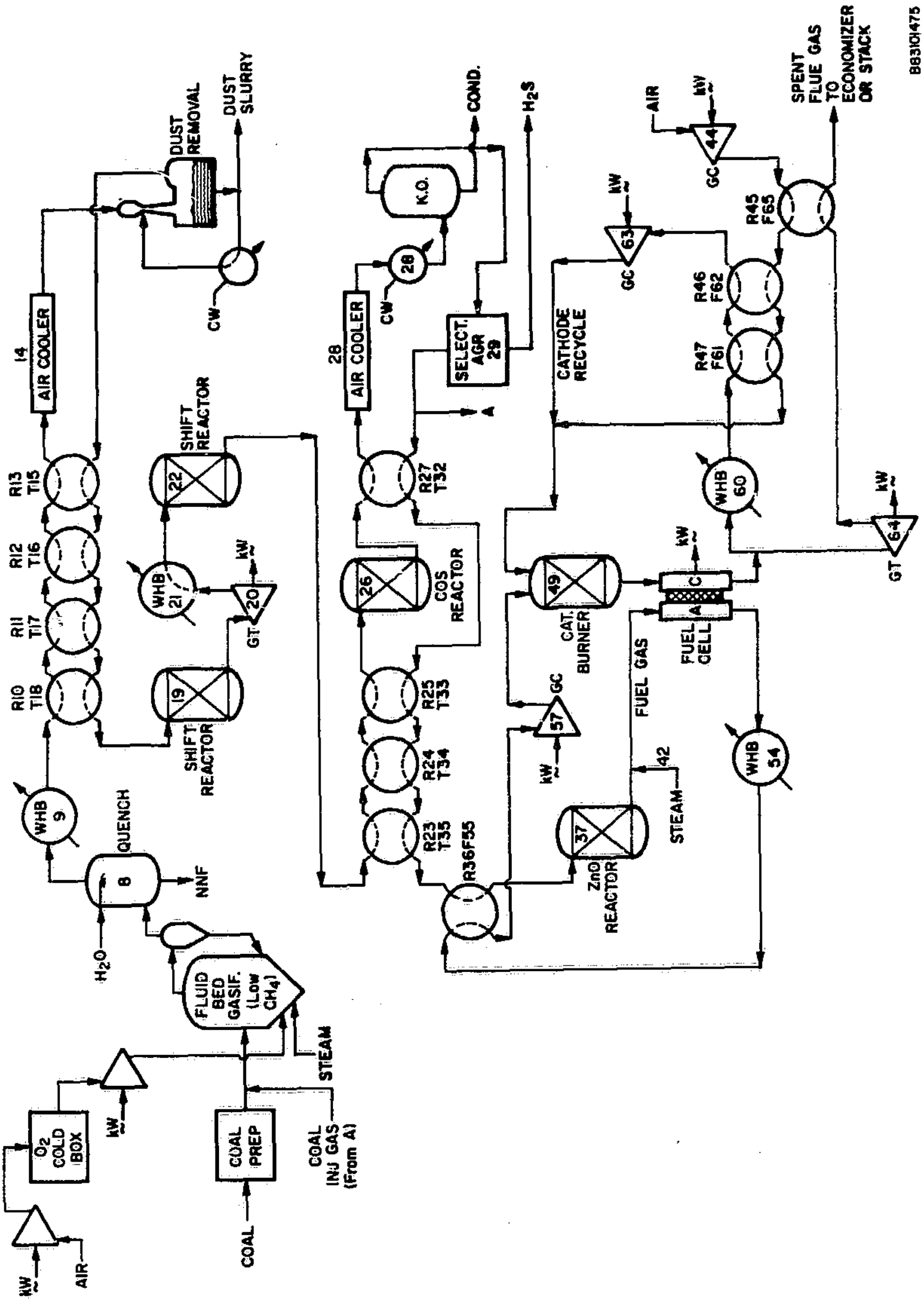

Figure 4. PROCESS FLOW DIAGRAM OF INITIAL EVALUATION WITH FLUIDIZED-BED GASIFICATION

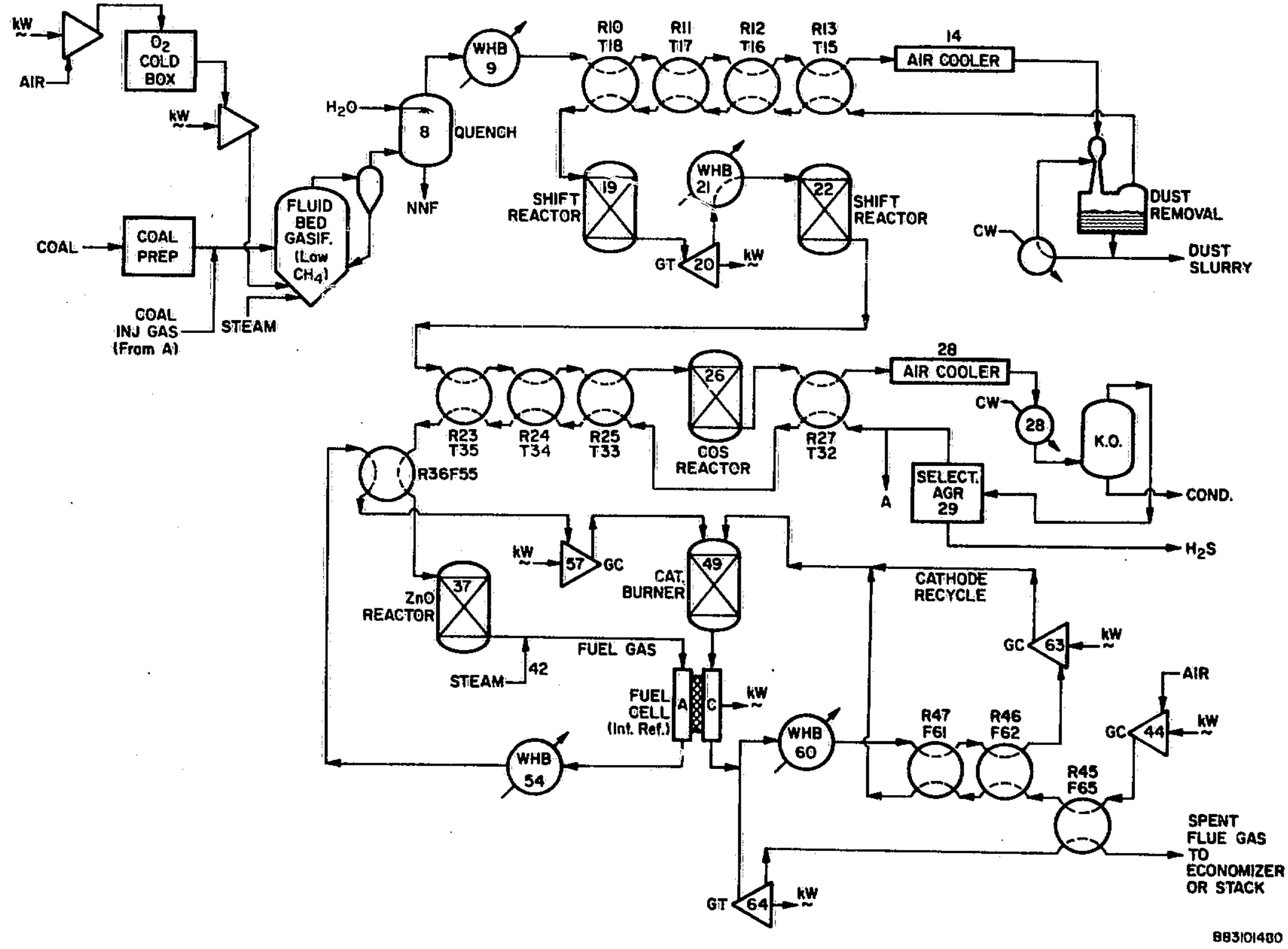

Figure 5. PROCESS FLOW DIAGRAM INCORPORATING FLUIDIZED-BED GASIFICATION AND STEAM-METHANE REFORMING INTEGRATED WITH THE FUEL CELL

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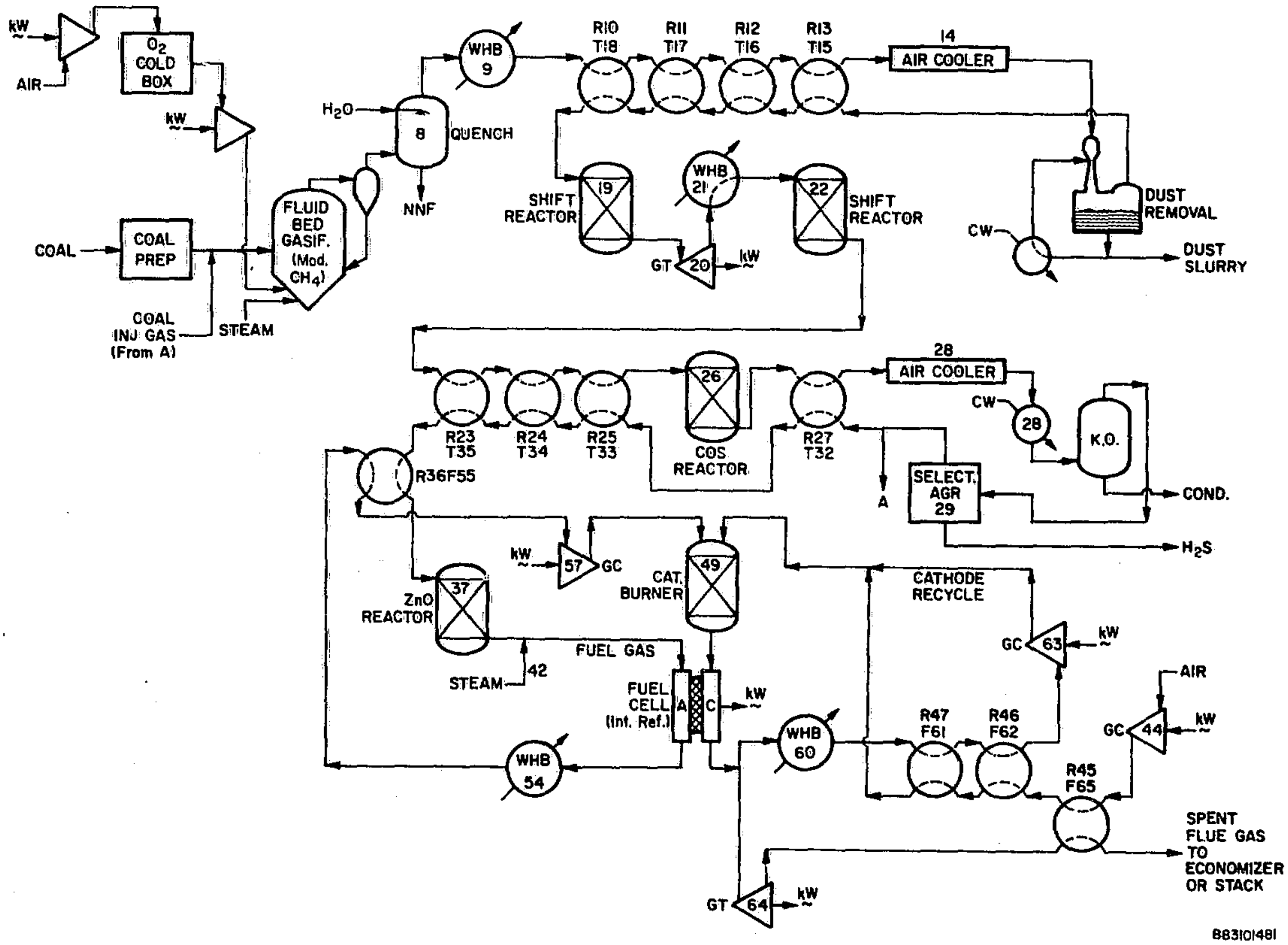

Figure 6. PROCESS FLOW DIAGRAM OF SYSTEM WITH INTERNALLY REFORMING FUEL CELL AND FLUIDIZED-BED GASIFICATION WITH MODERATE METHANE PRODUCTION

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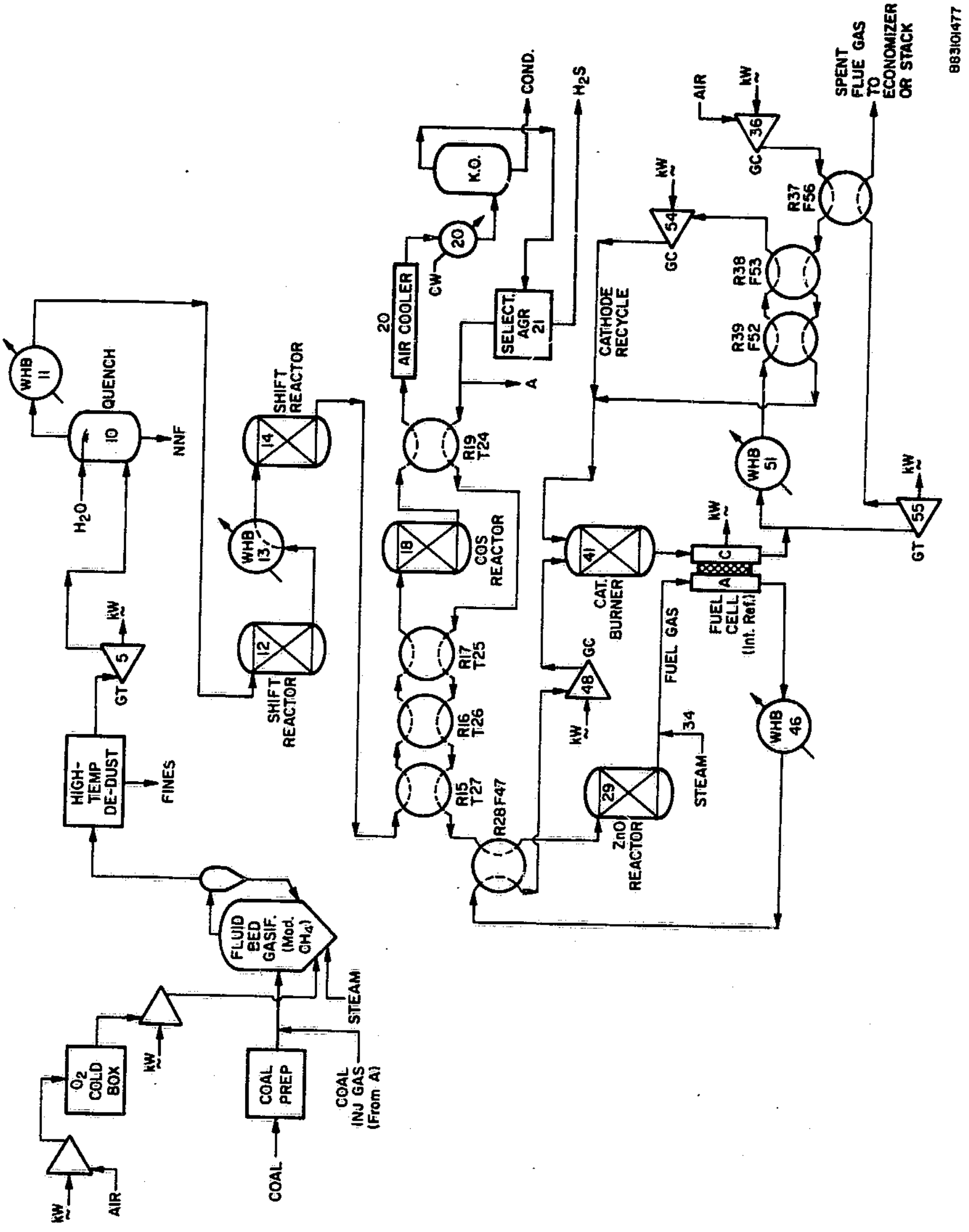

Figure 7. PROCESS FLOW DIAGRAM INCORPORATING HIGH-TEMPERATURE DUST REMOVAL

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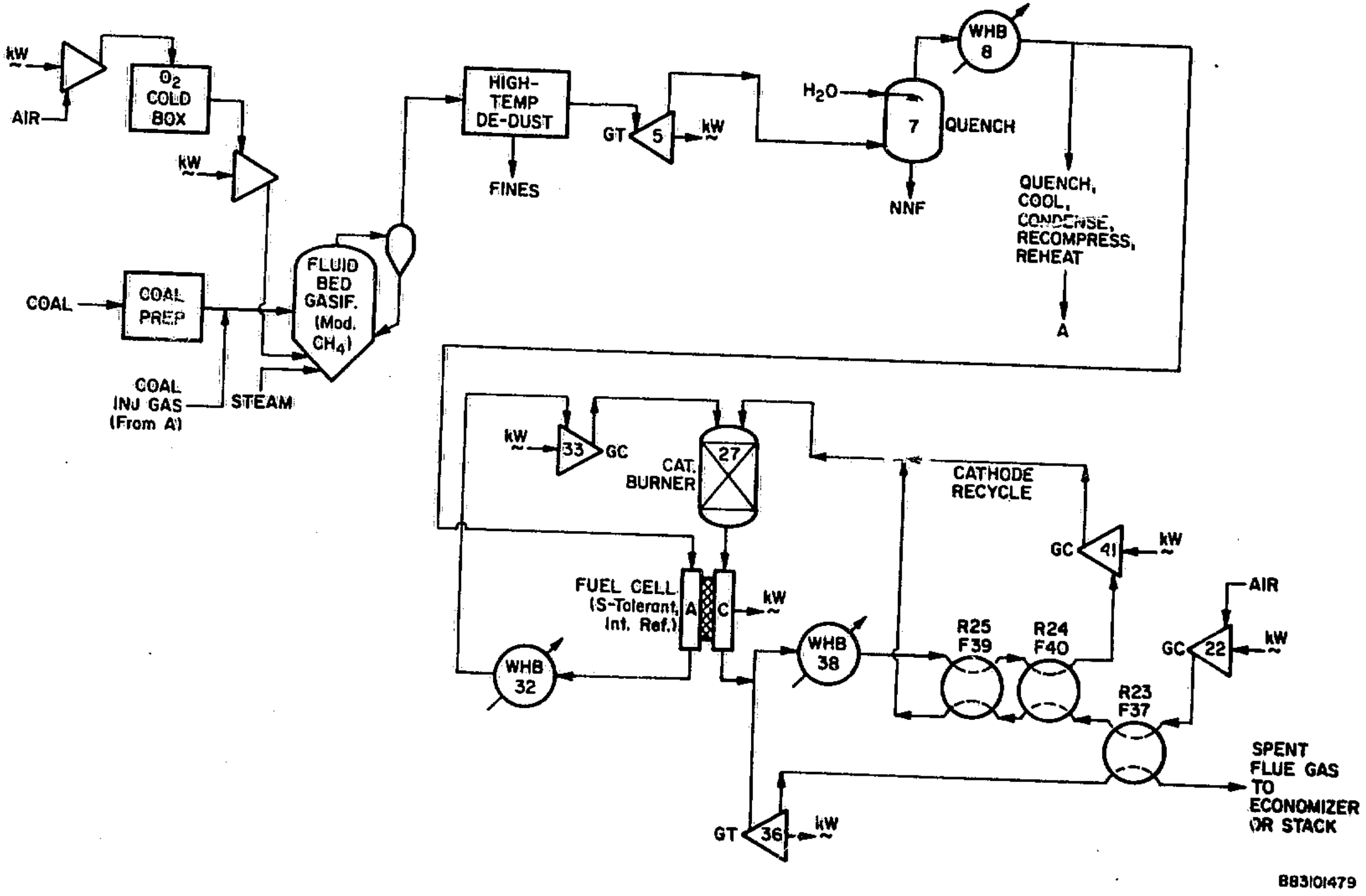

Figure 8. PROCESS FLOW DIAGRAM INCORPORATING SULFUR-TOLERANT FUEL CELL

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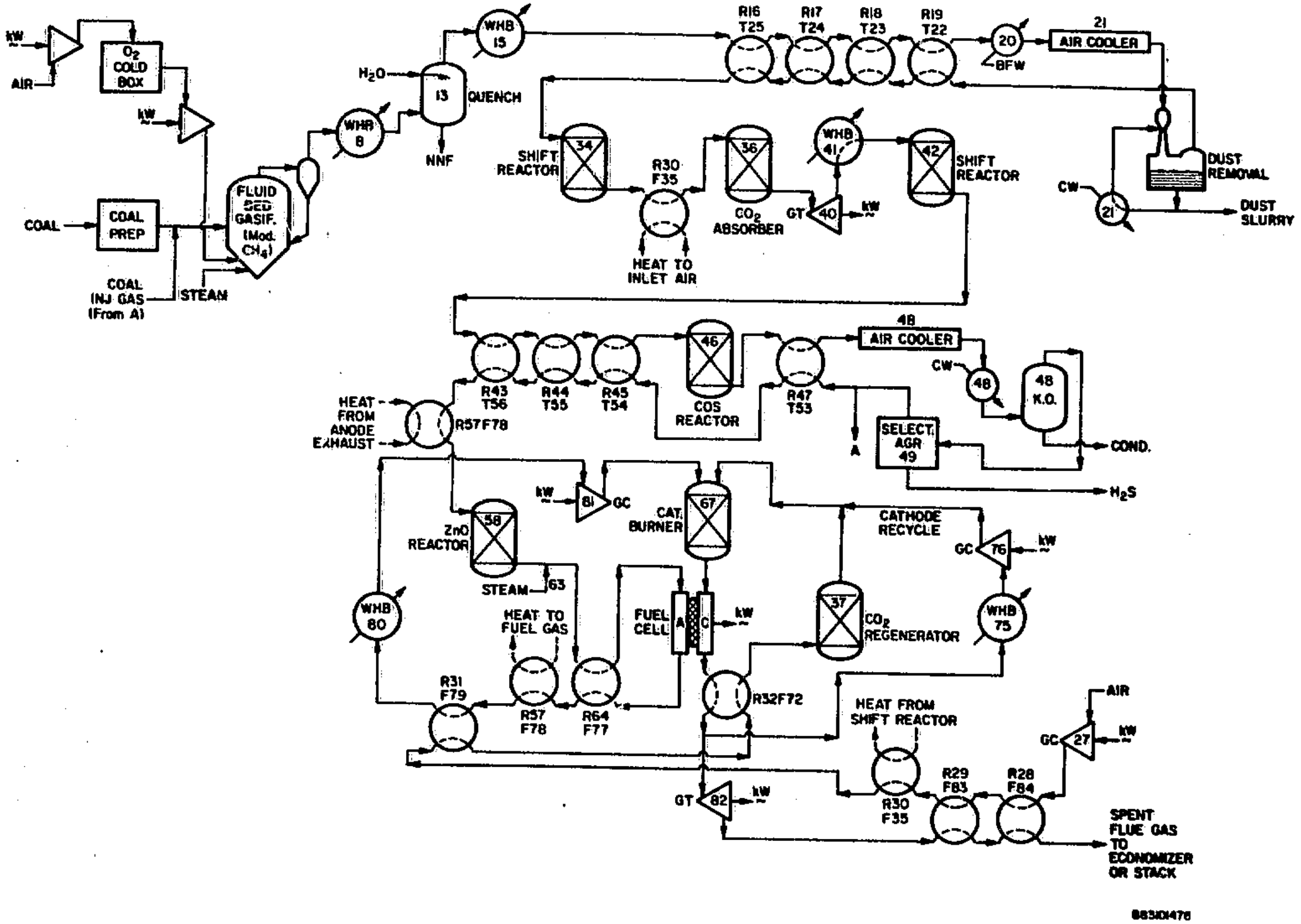

Figure 9. PROCESS FLOW DIAGRAM INCORPORATING HIGH-TEMPERATURE CO_2 TRANSFER

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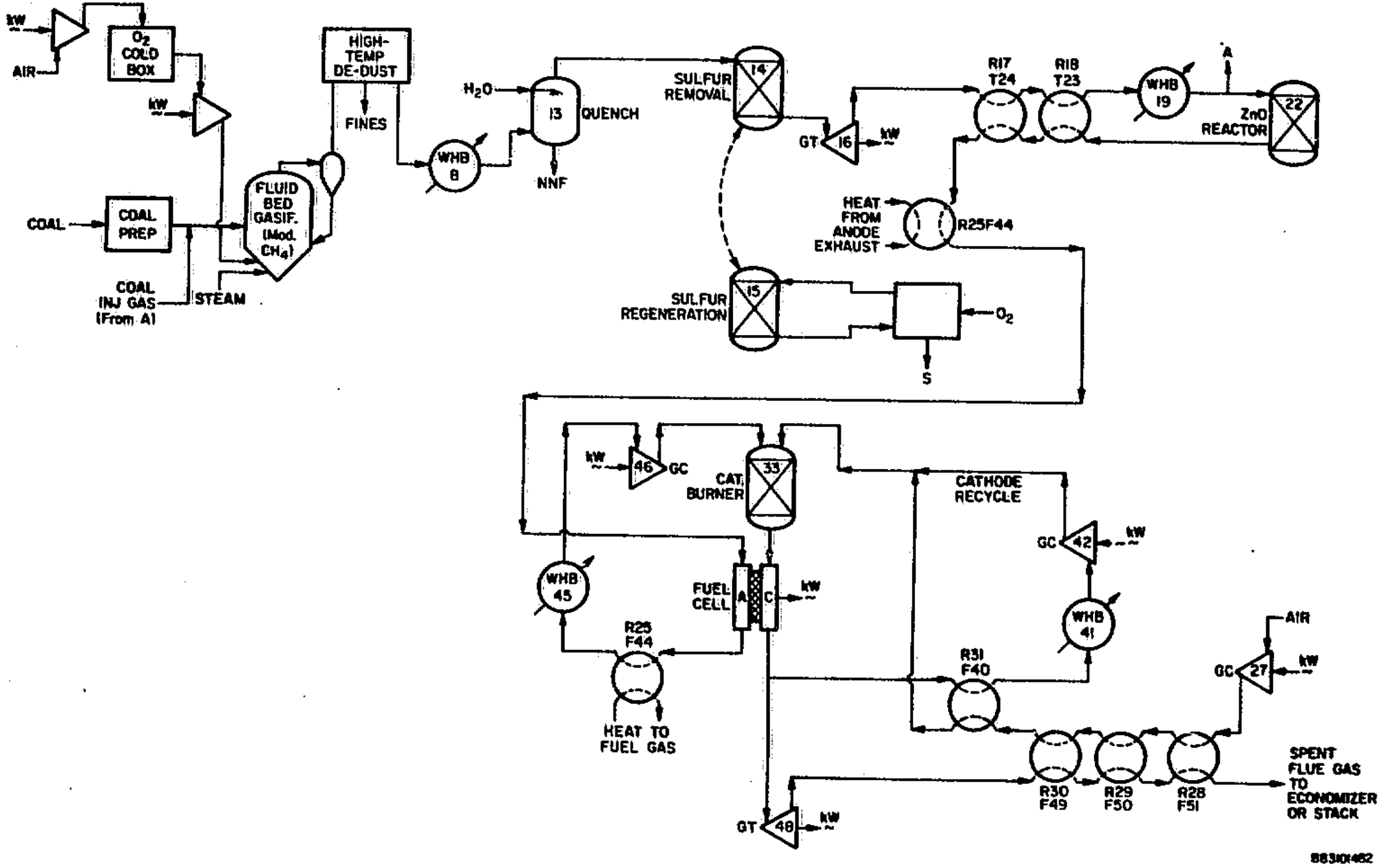

Figure 10. PROCESS FLOW DIAGRAM INCORPORATING HIGH-TEMPERATURE SULFUR REMOVAL

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 internalling 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 CO_2 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 CO_2-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-month 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.

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* 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 CO_2-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 CO_2 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 similator that would operate well at elevated pressures.

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6. "Technical and Economic Comparisons of the Westinghouse and IGT U-Gas Coal Gasification Processes," prepared by M. W. Kellogg Co. for DOE and GRI under Contract No. ET-78-C-01-2778, April 1981.

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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

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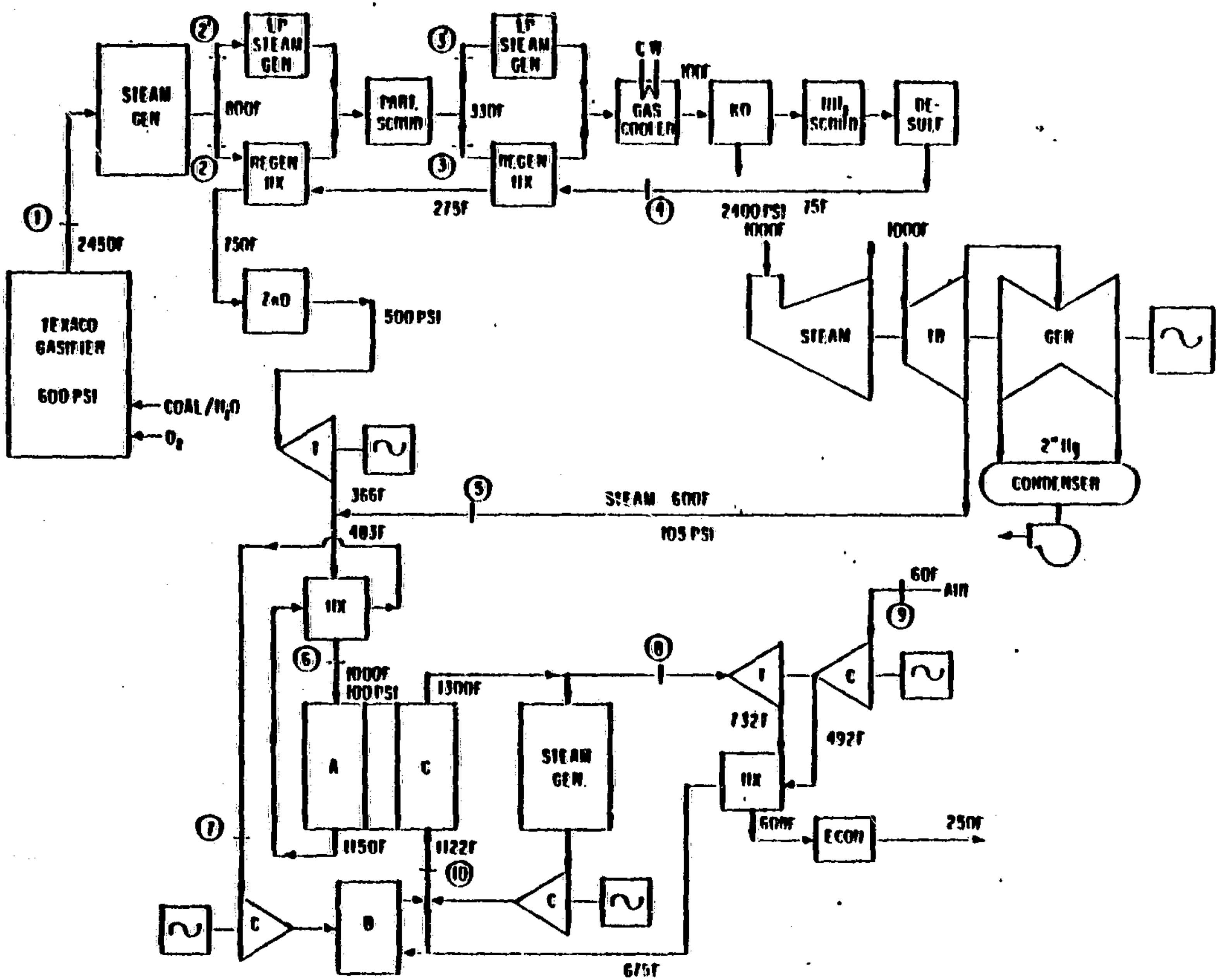

FIGURE 1. BASE CASE BLOCK FLOW DIAGRAM

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Table 1

BASE CASE FLOW STREAM GAS COMPOSITIONS

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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_2H_4 MF												
C_2H_6 MF												
CO_2 MF	0.087	→	→	→	→	0.096		0.219	0.509	0.110		0.137
H_2O 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_2S 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						105	100		99	14.7	
Temp (°F)	2450	800	800	330	330	75	600	1000	860	1300	60	1120

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_2S	0.0101
COS	0.0006
N_2	0.0078
H_2O	0.1788
NH_3	0.0019

FEEDS

O_2: 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 ($CO + H_2 + CH_4$) 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 H_2 in the fuel gas. Unreacted CO and H_2 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 CO_2 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 CO_2 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/cm^2). Fuel cell voltage this low may result in electrochemical corrosion phenomena within the fuel cell stack. Although improvements in molten carbonate fuel cell

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*Thermal cycle efficiency, coal pile to busbar.

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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_2O Oxidant Sensible Energy	0.008	
Injection Steam Energy	0.144	
Plant Compressors	0.02	
Total Energy Input to Gas Stream	1.172	
Energy Outputs		
Fuel Cell DC Output		0.305
Gas Tubine 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 = 0.46

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

- 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.

- 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.

- 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.

- Fuel Cell Average Current Density

  The design value of the fuel cell stack average current density is 160 mA/cm^2.

- 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.

- <u>Cell Polarization Factor</u>

  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.

- <u>Fuel Cell Stack Inverter</u>

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

2. PlantAuxiliary Power Requirements

Auxiliary power requirements are assumed to consist of the following:

- <u>Oxygen Plant</u>

  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.

- <u>Oxygen Compressors</u>

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

- <u>Plant Gas Recirculation Compressors</u>

  High efficiency electrically driven compressors are assumed.

- 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

   - <u>Compressors</u>

     Efficiency is assumed to be 0.88.

   - <u>Turbines</u>

     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

   - <u>Gas Purification</u>

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

- 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 N_2 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.

- CO_2 Recovery

  The Base Case employs a one stage acid gas removal process and 88% CO_2 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:

Fuel Processing Unit	Cost Base	Scaling Capacity Factor	Scaling Exponent C
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.

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* 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 cosed as follows:

State of Technology Development	Percentage of Installed Section Cost
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:

Year	Percent of Total Plant Cost
1	33 1/3
2	33 1/3
3	33 1/3

The capital estimates should be based on the following:

Year	Mid-1981
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:

- FIXED
  - Operating Labor
  - Maintenance Costs
  - Overhead Charges
- VARIABLE
  - Consumables
    - Fuel
    - Water
    - Ash Disposal
    - Catalysts and Chemicals
  - Variable Maintenance Charges (if any)
- 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

Process Unit Function	Percent/year. of Installed Plant Section Cost
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:

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

These items are discussed below.

The 1981 delivered coal cost is $1.65/10^6$ 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

Type	Illinois No. 6
PROXIMATE ANALYSIS (Wt %)	
Moisture	4.2
Ash	9.6
Fixed Carbon	52.0
Volatile Matter	34.2
	100.0
ULTIMATE ANALYSIS - DAF COAL (Wt %)	
Carbon	77.26
Hydrogen	5.92
Oxygen	11.14
Nitrogen	1.39
Sulfur	4.29
Other	-
	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:

		$MM
1.1	**Total Plant Cost (TPC)**	
	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	**Allowance for Funds During Construction (AFDC)**	
	AFDC = TPC (see TAG B-5-IV-1) = 945.35 (0.0373)	35.28
1.3	**Prepaid Royalties**	
	= 0.5% (TPC - Gen'l Facilities - Project Cont. on Gen'l. Facilities) = 0.005 (945.35 - 54.77 - 0.3 x 54.77)	4.37

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 x 0.31 = 1.7 MM 1.7

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

1.8 Total Capital Requirement (TCR)

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

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

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

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

3.0	Fixed Operating Costs		24.41
3.1	Operating Labor	3.74	
3.2	Maintenance Costs (60% Matls, 40% Labor)	17.45	
3.3	Overhead Charges (30% Oper + Maint. Labor	3.22	
		24.41	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:

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

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mid'78    mid'79      mid'80     mid'81
  |-------|-----------|---------|
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$$\frac{1}{3}\,\text{TPC} = \frac{1}{3}\,\text{TPC}$$

$$\frac{1}{3}\,\text{TPC}\ \text{o}\dashrightarrow\ = \frac{1}{3}\,\text{TPC.}\ {\frac{1+i}{1+e_i}}^{1} = \frac{1}{3}\,\text{TPC}\ {\frac{1.125}{1.085}}^{1}$$

$$\frac{1}{3}\,\text{TPC}\ \text{o}\dashrightarrow\ = \frac{1}{3}\,\text{TPC.}\ {\frac{1+i}{1+e_i}}^{2} = \frac{1}{3}\,\text{TPC}\ {\frac{1.125}{1.085}}^{2}$$

$$= \text{TPC}\ \frac{1}{3}\,(1 + 1.0751 + 1.0369)$$

$$= 1.0373 \times \text{TPC}$$

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

$$= .0373 \times 945.35$$

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

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

1.3 Prepaid Royalties:

= (0.005) x (T.P.C. - Gen'l Facilities - Project Cont. on Gen'l Facilities)

= (.005) x (945.35 - 54.77 - 0.3 x 54.77)

= $4.37 MM

1.4 Preproduction Or Start-up Costs:

a) 1 Month Fixed Oper. Costs

= 1/12 x $24.35 MM/Yr

= $2.03 MM

b) <u>1 month of variable operating costs at full capacity</u>

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

c) <u>25% of full capacity fuel cost for 1 month:</u>

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

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

d) <u>2% of the Total Plant Cost</u>

= .02 x $945.35

= $18.9 MM

Total Pre-Production or Start-Up Costs

$$\begin{array}{lr} = \text{a} & \$\ 2.03 \\ \text{b} & 6.1 \\ \text{c} & 1.5 \\ \text{d} & \underline{18.9} \\ & \$28.53 \text{ MM (1981 \$)} \end{array}$$

1.5 Inventory Capital:

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

<u>Coal</u>: $\frac{\$46.35}{.65} \times \frac{60}{365} =$ $11.7

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

$11.78 MM (1981 $)

1.6 Initial Chemical & Catalyst Charge:

Basis: Prorate based on AP-1543

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

Initial Charge = 5.4 x $0.31 MM/Yr

= $1.7 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 strorage area is proportional to power output.

Area = 105 + 675/1200 x 105 = 164 acres

Cost @ $5500/acre = $902,000

1.8 Total Capital Requirement

= $\sum$ 1.1 thru 1.8
= $ 1027.91 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:

$$403,\frac{199\text{ lbs}}{\text{hr}} \times \frac{12,235}{\text{lb}}\text{ Btu} \times 365\times24\times.65 \times \frac{\$1.65}{10^6\text{ Btu's}}$$

$$= \$46.35\text{ MM (1981 \$)}$$

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%

$$\frac{\text{Ratio of Condenser Duty}}{\text{Steam turbine power}} = \frac{1-.38}{.38}$$

Steam turbine input = 192 MW
= 655.3 MM Btu/hr

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

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

$$= 1,188\text{ MM Btu/hr}$$

$$\text{CTW make-up} = 1,188 \times 10^6\ \frac{\text{Btu}}{\text{hr}} \times \frac{\text{lb } H_2O\text{ M.U.}}{1000\text{ Btu's}}$$

$$= 1,188 \times 10^3\text{ lbs } H_2O/\text{hr}$$

$$= 142,600\text{ gal/hr}$$

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

Variable Operating Costs (Cont'd)

Costs per year @ 65% capacity

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

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

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

2.3 Ash Disposal Costs:

Assume Costs = $5/ton

Costs @ 65% Cap. =

$$4838 \ \overset{\text{TPD}}{\frac{\text{coal}}{\text{day}}} \times \overset{\text{Ash}}{.096} \times 365 \times .65 \times \$5$$

$$= \$0.55 \text{ MM/Yr} \quad (1981\ \$)$$

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:

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

$$= \underline{\underline{\$0.31}} \text{ MM/yr } 1981\ \$$$

C-2

3.0 FIXED OPERATING COSTS

Operating Labor = 3.7
Maintenance costs = 17.45
Overhead Charges = 3.2

$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	2.0		16.11
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.	1.5		36.66
balance sub-total		8.7	621.29
		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

4.0 FUEL CELL REPLACEMENT COST (as Operating Expense)

4.1 1981 Fuel Cell Stock Cost Caluculation

4.1.1 Ref. EPRI EM-1670:

$ 97.63/kw 1978 costs

= $133/kw 1981 scaled from CE plant costing index
- for 675 MW plant, 442 MW is MCFC

Cost = $58.8 MM 1981

4.1.2 Assume

- $\frac{\$13.60}{\text{ft.}^2 \text{ active area}}$ (1981 Costs w/out contingencies)
- $160 \, \frac{\text{milliamps}}{\text{cm}^2} \times \frac{\text{amps}}{1000 \text{ milliamps}} \times \frac{929 \text{ cm}^2}{\text{ft.}^2} = 149 \, \frac{\text{Amps}}{\text{ft.}^2}$
- $.69 \text{ volts} \times 149 \, \frac{\text{amps}}{\text{ft.}^2} = \frac{103 \text{ watts}}{\text{ft.}^2}$
- $\text{Total ft.}^2 = 442 \times 10^6 \text{ watts} \times \frac{\text{ft.}^2}{103 \text{ watts}}$

$$= 4.29 \times 10^6 \text{ ft.}^2$$

$\text{Cost} = 4.29 \times 10^6 \times \$13.60$

$= \$58 \text{ MM in 1981 } \$$

4.2 FUEL CELL REPLACEMENT COST

Use Level Annual Payment = x
Use Ten Year Period
Replacement @ 5 yrs & @ 10 Yrs
Base Cost @ Yr = 0 ; = $58 MM
Apparent Escalation = 8.5%/Yr = e_a $e_a = (1 + e_r)/(1 + e_i)$
Real Escalation = 0.0/Yr = e_r
Inflation Rate = 8.5%/Yr = e_i
Discount Rate = 12.5%/Yr = i

Cost to replace @ end of 5 yrs = $\$58 \, (1.085)^5$
= $87.21 MM

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

$x\,[(1+i)^5 + (1+i)^4 + (1+i)^3 + (1+i)^2 + (1+i)] = 7.21829X$
= 5 yr. Sinking Fund

Amount remaining after 1st replacement:
7.21829 X - 87.21

Amount @ end of 10 yr = \$131.14 mm =
Amount Remaining after 1st Payment + Accumulated 5 yr. Sinking Fund

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

$20.23X - 157.16 = 131.14$

$X =$ \$14.25 mm/yr

+ Process Cont. @ 50% = 7.13 mm/yr
+ Project Cont. @ 30% = 4.28 mm/yr

\$25.66 mm/yr

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

5.0 COST OF ELECTRICITY (COE) CALCULATION

5.1 Calculate short-term 10 yr, levelized COE:

1st 10 Years Levelized Revenue Requirement

$= (TCR)\ P_f + (Expenses) \times LF_f$ - Notes (1) & (2)

$= 1027.91 \times .180 + (Expenses) \times LF_f$

Expenses = Fixed + Coal + Other Variable + Fuel Cell Replacement

	1981 Exps	Int. Rate (i)	Real Escal. (e_r)	Inflat. Rate (e_i)	Apparent Esc. (e_a)	Level Factor (L_{10})	Level Exps
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	"	"	"	"	-	25.66
							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
25.7	8.0	• Fuel Cell Replacement Expense
$320.5	100.0	

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

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_2O	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_2S	440			-	-				-		440	-	-	-
COS	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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BASES FOR UNIT ESTIMATING

UNIT	Ref.	YR	Total Plant Inv. MM$	Process Cont.	Proj. Cont.	Constructed Cost w/out Cont. Note(1)	Plant Basis	Size	#TRAINS TOTAL	#TRAINS SPARE
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 H_2S)	3 (3-50% Units)	1 (14%
							Tail Gas Flow ------------->	8,227		
Fuel Cells	AP-1670; P. B-4	Mid-78					Fuel Cell AC Output (MW)		5 islands, 18 Fuel cells/ islands, 5 MW/Fuel Cell	
Fuel Cell Piping	"	"	158.7	49.6	9.9	99.2	(Fuel Cell Module @$97.63/kW w/out contingency)			
Remainder Fuel Cell Sub-Systems	"	"								
Steam Bottoming Cycle — Expansion Turbines +	AP-1670	"	82.4	-	7.5	74.9	Gross Shaft Output (MW)	125.7 +	5	
Steam Cycle	AP-1670	"					Gross ST Output (MW)	171.6	1	
Inverters	AP-1670	"	43.0	2.9	2.9	29.0	Fuel Cell AC Output (MW)	442	45 (1 island)	
Accessory Elect. Equip. Step Up & Transmission	"	"	20.1	-	1.8	18.3	Total Plant Output (MW)	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. Ovhd. + Fees Note (3)	Note (3) Process Cont. (1981)	Proj. Cont.	Total Plant Cost	$/kW
				$MM	$MM	$MM	$MM	$MM	$MM	$MM	$/kW
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	2,167 mph [Acid Gas=44% H_2S]	2/1	0.60	7.6	11.7			0.9	3.51	16.11	24.0
Fuel Processor Sub-Totals					240.2			11.8	72.06	324.06	480.0
Fuel Cells Fuel Cell Piping, etc.	432 MW		1.0	97.0	149.3			74.7	44.79	268.79	398.0
Steam+GT Expansion Turbines Bottoming + Cycles Steam Cycle	172 MW 192 MW		1.0	91.7	141.2			0	42.36	183.56	272.0
Inverters	432 MW		1.0	28.3	43.6			4.4	13.08	61.08	90.5
Accessory Elect. Equip. Step Up & Transmission	675		1.0	18.3	28.2			0	8.46	36.66	54.0
Power Cycles Sub-totals					362.3			79.1	108.69	550.09	814.5
Process (On-Site) Sub-Total					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			-	16.43	71.2	105.0
				TOTAL	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 (CE Plant Cost Indices); Mid'78 to Mid'81 = 1.36
3. E.H.O. Ovhd + Fees = Engineering and Home Office Overload 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%

APPENDIX B. Summary Printouts of Cases for Economic Analysis

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Case 1

The Base Case at 675 MW Output

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1  JPL.1  BASE CASE  85% CONV., NO QUENCH, NO PRE-SHIFT, ISO EFF: COMP 88% T 92%                CASE    1  JPL.1
                                                                                                     SUMMARY
                                                                                                18 MAR 83 10:20:52
            ORIGINAL BASE CASE AS SPECIFIED BY JET PROPULSION LABS.

  THIS IS THE FINAL SYNOPSIS OF THE ENERGY BUDGET OF AN INTEGRATED POWER PLANT USING COAL
GASIFICATION AND ELECTROCHEMICAL FUEL CELLS, ALONG WITH CONVENTIONAL POWER GENERATION MACHINERY.

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        *********      COAL CONSUMPTION IS   5137. TONS PER DAY.    *********
        *********                                            *********
        *********                                            *********
        *********                                            *********
        *********      NET DELIVERED POWER IS  675. MEGAWATTS.      *********
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---------------------------------------------DETAILS______________________________________

COAL CONSUMPTION IS  428. THOUSANDS POUNDS PER HOUR, OR ABOUT  5471. MILLIONS BTU.
THE OXYGEN PLANT SIZE IS  4337. TONS PER DAY.
PROCESS STEAM CONSUMPTION IS   391. THOUSANDS POUNDS PER HOUR.

PROCESS MACHINERY (COMPRESSORS AND MAJOR PUMPS) NEED                          -290289. KILOWATTS
PROCESS GAS TURBO-EXPANDERS & CLOSELY-ASSOCIATED STEAM TURBINES GIVE           297950. KILOWATTS
CONDENSING STEAM TURBINE (MULTI-STAGE W. REHEAT) GIVES                         113778. KILOWATTS
THE FUEL CELL ITSELF DELIVERS                                                  553352. KILOWATTS
                 (AFTER CONVERSION TO 13.8KV-AC @ 98% EFF.)
THE NET POWER DELIVERED TO BUSBAR IS                                           675000. KILOWATTS

THE VOLTAGE OF THE FUEL CELL IS          0.8232 VOLTS.
THE AREA OF THE FUEL CELL IS           4536647. SQ. FT.

OVERALL PLANT EFFICIENCY, COAL PILE TO BUSBAR, IS 42.1 PERCENT.
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```
1  JPL.1  BASE CASE  85% CONV., NO QUENCH, NO PRE-SHIFT, ISO EFF: COMP 88% T 92%              CASE    1  JPL.1
                                                                                                 POWER LIST
                                                                                              15 MAR 83 10:20:52

****************************************************************************************************************

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. LABS. BASE CASE.
  13    -1139.2 KILOWATTS      13 REMOVE SULFUR FROM RAW GAS BY SELEXOL SYSTEM.
  17    23530.6 KILOWATTS      17 EXPAND SWEETENED RAW GAS TO FUEL CELL PRESSURE.
  22    22066.0 KILOWATTS      22 BLOW STEAM DOWN POWER TURBINE. ADD TO PROCESS GAS (# 17).
   3 -152036.1 KILOWATTS      3 COMPRESS COMBUSTION AIR TO FUEL CELL PRESSURE.
  10  564644.5 KILOWATTS      10 THE FUEL CELL, 85% CONVERSION, SHIFT ONLY WITHIN ANODE.
  15    -9204.1 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  BASE CASE  85% CONV., NO QUENCH, NO PRE-SHIFT, ISO EFF: COMP 88% T 92%          CASE    1  JPL.1
                                                                                              STEAM LIST
                                                                                         13 MAR 83 10:20:52
*****************************************************************************************************

TABULATION OF STEAM NEEDED IN PROCESS.

13        65.0 PSIA           4887. MOLS STEAM  13 REMOVE SULFUR FROM RAW GAS BY SELEXOL SYSTEM.
22      1500.0 PSIA          16835. MOLS STEAM  22 BLOW STEAM DOWN POWER TURBINE. ADD TO PROCESS GAS (# 17).
```

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```
1  JPL.1  BASE CASE  85% CONV., NO QUENCH, NO PRE-SHIFT, ISO EFF: COMP 98% T 92%                CASE     1  JPL.1
                                                                                                  WASTE HEAT OUT
                                                                                                19 MAR 83 10:20:52

****************************************************************************************************************

TABULATION OF WASTE HEAT AVAILABLE FROM PROCESS BOILERS.

   4      900.0 DEGREES F.    658.49 MILLIONS BTU    4 COOL RAW GAS TO 900 DEGREES FAHRENHEIT.
   7      600.0 DEGREES F.     19.97 MILLIONS BTU    7 WASTE HEAT BOILER PARALLEL TO RAW GAS RECUP. EXCH. TRAIN.
  14      850.0 DEGREES F.    191.95 MILLIONS BTU    14 COOL ANODE EXIT GAS IN WHB - RAISE STEAM.
  18     1050.0 DEGREES F.    562.39 MILLIONS BTU    19 COOL CATHODE RECYCLE GAS IN WHB - RAISE STEAM.
```

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CASE 2

Add Quench, Shift, Crossflow Heat Exchange

```
L.JPL.1F  MODIFIED BASE CASE 85% CONV, QUENCH, PRE-SHIFT, COSX, TURBO-EXP MOVED.             CASE    L.JPL.1F
                                                                                                   SUMMARY
                                                                                             15 FEB 83 14:30:43
             70 TALLY UTILITIES; REPORT; CLOSE UP SHOP; GO TO NEXT CASE

  THIS IS THE FINAL SYNOPSIS OF THE ENERGY BUDGET OF AN INTEGRATED POWER PLANT USING COAL
GASIFICATION AND ELECTROCHEMICAL FUEL CELLS, ALONG WITH CONVENTIONAL POWER GENERATION MACHINERY.

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        *********                                                         *********
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        *********         COAL CONSUMPTION IS   5211. TONS PER DAY.       *********
        *********                                                         *********
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        *********                                                         *********
        *********         NET DELIVERED POWER IS  675. MEGAWATTS.         *********
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--------------------------------------------DETAILS_______________________________________
COAL CONSUMPTION IS  434. THOUSANDS POUNDS PER HOUR, OR ABOUT  5550. MILLIONS BTU.
THE OXYGEN PLANT SIZE IS  4399. TONS PER DAY.
PROCESS STEAM CONSUMPTION IS   320. THOUSANDS POUNDS PER HOUR.

PROCESS MACHINERY (COMPRESSORS AND MAJOR PUMPS) NEED                    -275167. KILOWATTS
PROCESS GAS TURBO-EXPANDERS & CLOSELY-ASSOCIATED STEAM TURBINES GIVE     314990. KILOWATTS
CONDENSING STEAM TURBINE (MULTI-STAGE W. REHEAT) GIVES                    65256. KILOWATTS
THE FUEL CELL ITSELF DELIVERS                                            569921. KILOWATTS
                  (AFTER CONVERSION TO 13.6KV-AC @ 98% EFF.)
THE NET POWER DELIVERED TO BUSBAR IS                                     675000. KILOWATTS

THE VOLTAGE OF THE FUEL CELL IS         0.8408 VOLTS.
THE AREA OF THE FUEL CELL IS          4653155. SQ. FT.

OVERALL PLANT EFFICIENCY, COAL PILE TO BUSBAR, IS 41.5 PERCENT.
```

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

*******************************************************************************************************************

TABULATION OF SHAFT ENERGY AVAILABLE OR NEEDED.

 102   -57995.0 KILOWATTS      AIR COMPRESSOR FOR SUPPLY TO CRYOGENIC OXYGEN PLANT
 104   -25303.2 KILOWATTS      CRYOGENIC OXYGEN COMPRESSOR FOR SUPPLY TO COAL GASIFIER
   2    -5992.8 KILOWATTS      2 TEXACO GASIFIER FOR JET PROP. LABS. 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.
```

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```
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  MODIFIED BASE CASE 85% CONV, QUENCH, PRE-SHIFT, COSX, TURBO-EXP MOVED.          CASE    L.JPL.1F
                                                                                            WASTE HEAT OUT
                                                                                          15 FEB 83 14:30:43
*****************************************************************************************************

TABULATION OF WASTE HEAT AVAILABLE FROM PROCESS BOILERS.

   8      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

```
                                                                                                   CASE    L.JPL.2
L.JPL.2    FLUID BED BASE CASE 85% CONV, QUENCH, PRE-SHIFT, COSX, 8.33% CH4 RAW                         SUMMARY
                                                                                                   15 FEB 83 14:41:31
              STANDARD U-GAS GASIFIER WITHOUT METHANE CREDIT IN FUEL CEL

  THIS IS THE FINAL SYNOPSIS CF THE ENERGY BUDGET OF AN INTEGRATED POWER PLANT USING COAL
GASIFICATION AND ELECTROCHEMICAL FUEL CELLS, ALONG WITH CONVENTIONAL POWER GENERATION MACHINERY.

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         *********                                                              *********
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         *********                                                              *********
         *********          COAL CONSUMPTION IS   5646. TONS PER DAY.           *********
         *********                                                              *********
         *********                                                              *********
         *********                                                              *********
         *********          NET DELIVERED POWER IS  675. MEGAWATTS.             *********
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----------------------------------------------_DETAILS_------------------------------------------

COAL CONSUMPTION IS  470. THOUSANDS POUNDS PER HOUR, OR ABOUT  6013. MILLIONS BTU.
THE OXYGEN PLANT SIZE IS  3690. TONS PER DAY.
PROCESS STEAM CONSUMPTION IS   887. THOUSANDS POUNDS PER HOUR.

PROCESS MACHINERY (COMPRESSORS AND MAJOR PUMPS) NEED                         -346231. KILOWATTS
PROCESS GAS TURBO-EXPANDERS & CLOSELY-ASSOCIATED STEAM TURBINES GIVE          373660. KILOWATTS
CONDENSING STEAM TURBINE (MULTI-STAGE W. REHEAT) GIVES                         89488. KILOWATTS
THE FUEL CELL ITSELF DELIVERS                                                 558084. KILOWATTS
                   (AFTER CONVERSION TO 13.6KV-AC @ 99% EFF.)
THE NET POWER DELIVERED TO BUSBAR IS                                          675000. KILOWATTS

THE VOLTAGE OF THE FUEL CELL IS           0.8263 VOLTS.
THE AREA OF THE FUEL CELL IS            4636815. SQ. FT.

OVERALL PLANT EFFICIENCY, COAL PILE TO BUSBAR, IS 38.3 PERCENT.
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L.JPL.2   FLUID BED BASE CASE 85% CONV, QUENCH, PRE-SHIFT, COS%, 8.33% CH4 RAW        CASE    L.JPL.2
                                                                                         POWER LIST
                                                                                    15 FEB 83 14:41:31

**************************************************************************************************************

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 -194211.1 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 14:41:31
***************************************************************************************************************

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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L.JPL.2   FLUID BED BASE CASE 85% CONV, QUENCH, PRE-SHIFT, COSX, 8.33% CH4 RAW              CASE    L.JPL.2
                                                                                                 WASTE HEAT OUT
                                                                                                15 FEB 83 14:41:31

****************************************************************************************************************

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

```
L.JPL.3   FLUID BED, REFORM, 95% CONV, QUENCH, PRE-SHIFT, COSX, 8.33% CH4 RAW        CASE    L.JPL.3
                                                                                            SUMMARY
                                                                                       15 FEB 83 15:00:37
          STANDARD U-GAS GASIFIER WITH METHANE REFORMED IN FUEL CELL

  THIS IS THE FINAL SYNOPSIS OF THE ENERGY BUDGET OF AN INTEGRATED POWER PLANT USING COAL
GASIFICATION AND ELECTROCHEMICAL FUEL CELLS, ALONG WITH CONVENTIONAL POWER GENERATION MACHINERY.

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     *********                                                    *********
     *********        COAL CONSUMPTION IS   4823. TONS PER DAY.   *********
     *********                                                    *********
     *********                                                    *********
     *********                                                    *********
     *********        NET DELIVERED POWER IS  675. MEGAWATTS.     *********
     *********                                                    *********
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-------------------------------------------DETAILS_____________________________________

COAL CONSUMPTION IS  402. THOUSANDS POUNDS PER HOUR, OR ABOUT  5137. MILLIONS BTU.
THE OXYGEN PLANT SIZE IS  3152. TONS PER DAY.
PROCESS STEAM CONSUMPTION IS   758. THOUSANDS POUNDS PER HOUR.

PROCESS MACHINERY (COPPRESSORS AND MAJOR PUMPS) NEED                  -264590. KILOWATTS
PROCESS GAS TURBO-EXPANDERS & CLOSELY-ASSOCIATED STEAM TURBINES GIVE   311091. KILOWATTS
CONDENSING STEAM TURBINE (MULTI-STAGE W. REHEAT) GIVES                  11191. KILOWATTS
THE FUEL CELL ITSELF DELIVERS                                          617317. KILOWATTS
             (AFTER CONVERSION TO 13.6KV-AC @ 98% EFF.)
THE NET POWER DELIVERED TO BUSBAR IS                                   675000. KILOWATTS

THE VOLTAGE OF THE FUEL CELL IS         0.8411 VOLTS.
THE AREA OF THE FUEL CELL IS          5374565. SQ. FT.

OVERALL PLANT EFFICIENCY, COAL PILE TO BUSBAR, IS 44.8 PERCENT.
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L.JPL.3   FLUID BED, REFORM, 85% CONV, QUENCH, PRE-SHIFT, COSX, 8.33% CH4 RAW                         CASE    L.JPL.3
                                                                                                          POWER LIST
                                                                                                      15 FEB 83 15:00:37

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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   -1856.7 KILOWATTS     70 COMPRESS RECYCLE GAS TO INJECT FEED COAL TO GASIFIER.
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L.JPL.3   FLUID BED, REFORM, 85% CONV, QUENCH, PRE-SHIFT, COSX, 8.33% CH4 RAW                CASE    L.JPL.3
                                                                                               STEAM LIST
                                                                                           15 FEB 83 15:00:37
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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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L.JPL.3   FLUID BED, REFORM, 95% CONV, QUENCH, PRE-SHIFT, COSX, 8.33% CH4 RAW          CASE    L.JPL.3
                                                                                     WASTE HEAT OUT
                                                                                  15 FEB 83 15:00:37

**********************************************************************************************************

TABULATION OF WASTE HEAT AVAILABLE FROM PROCESS BOILERS.

   9     700.0 DEGREES F.    491.51 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.C 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

```
L.AIR.4E  FLUID BED, REFORM, 85% CONV, QUENCH, PRE-SHIFT, COSX, 13.5% CH4 RAW          CASE    L.AIR.4E
                                                                                          SUMMARY
             METH-PICH U-GAS WITH ANODE METHANE REFORMING, 13.5% C TO C              15 FEB 83 15:36:37

  THIS IS THE FINAL SYNOPSIS OF THE ENERGY BUDGET OF AN INTEGRATED POWER PLANT USING COAL
 GASIFICATION AND ELECTROCHEMICAL FUEL CELLS, ALONG WITH CONVENTIONAL POWER GENERATION MACHINERY.

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         *********         COAL CONSUMPTION IS   4726. TONS PER DAY.        *********
         *********                                                          *********
         *********                                                          *********
         *********                                                          *********
         *********         NET DELIVERED POWER IS  675. MEGAWATTS.          *********
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 ---------------------------------------------DETAILS---------------------------------------

 COAL CONSUMPTION IS  394. THOUSANDS POUNDS PER HOUR, OR ABOUT  5032. MILLIONS BTU.
 THE OXYGEN PLANT SIZE IS  2862. TONS PER DAY.
 PROCESS STEAM CONSUMPTION IS   750. THOUSANDS POUNDS PER HOUR.

 PROCESS MACHINERY (COMPRESSORS AND MAJOR PUMPS) NEED                       -255350. KILOWATTS
 PROCESS GAS TURBO-EXPANDERS & CLOSELY-ASSOCIATED STEAM TURBINES GIVE        306179. KILOWATTS
 CONDENSING STEAM TURBINE (MULTI-STAGE W. REHEAT) GIVES                        3912. KILOWATTS
 THE FUEL CELL ITSELF DELIVERS                                               620260. KILOWATTS
                 (AFTER CONVERSION TO 13.6KV-AC @ 98% EFF.)
 THE NET POWER DELIVERED TO BUSBAR IS                                        675000. KILOWATTS

 THE VOLTAGE OF THE FUEL CELL IS          0.8350 VOLTS.
 THE AREA OF THE FUEL CELL IS          5439707. SQ. FT.

 OVERALL PLANT EFFICIENCY, COAL PILE TO BUSBAR, IS 45.8 PERCENT.
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                                                                                   CASE    L.AIR.4E
L.AIR.4E  FLUID BED, REFORM, 85% CONV, QUENCH, PRE-SHIFT, COSX, 13.5% CH4 RAW            POWER LIST
                                                                                  15 FEB 83 15:36:37

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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 -162012.4 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  266004.1 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           7590. MOLS STEAM  42 ADD STEAM VIA TURBINE; PREVENT CARBON DROP IN FUEL CELL.
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L.AIR.4E  FLUID BED, REFORM, 95% CONV, QUENCH, PRE-SHIFT, COSX, 13.5% CH4 RAW                          CASE    L.AIR.4E
                                                                                                          WASTE HEAT OUT
                                                                                                        15 FEB 83 15:36:37

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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

```
L.QEX.2   HIGH TEMPERATURE DE-DUSTING, RAW GAS LIKE JPL4, QUENCH AFTER TURBINE.              CASE    L.QEX.2
                                                                                                 SUMMARY
                                                                                           17 FEB 83   7:33:26
                 HI-TEMP DE-DUST, LIKE JPL.4, ANODE REFORMING, 13.5% C TO C

   THIS IS THE FINAL SYNOPSIS OF THE ENERGY BUDGET OF AN INTEGRATED POWER PLANT USING COAL
 GASIFICATION AND ELECTROCHEMICAL FUEL CELLS, ALONG WITH CONVENTIONAL POWER GENERATION MACHINERY.

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            *********         COAL CONSUMPTION IS   4578. TONS PER DAY.        *********
            *********                                                          *********
            *********                                                          *********
            *********                                                          *********
            *********         NET DELIVERED POWER IS  675. MEGAWATTS.          *********
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-------------------------------------------DETAILS-------------------------------------------

COAL CONSUMPTION IS  381. THOUSANDS POUNDS PER HOUR, OR ABOUT  4874. MILLIONS BTU.
THE OXYGEN PLANT SIZE IS  2772. TONS PER DAY.
PROCESS STEAM CONSUMPTION IS   726. THOUSANDS POUNDS PER HOUR.

PROCESS MACHINERY (COMPRESSORS AND MAJOR PUMPS) NEED                        -247364. KILOWATTS
PROCESS GAS TURBO-EXPANDERS & CLOSELY-ASSOCIATED STEAM TURBINES GIVE         318425. KILOWATTS
CONDENSING STEAM TURBINE (MULTI-STAGE W. REHEAT) GIVES                         3114. KILOWATTS
THE FUEL CELL ITSELF DELIVERS                                                600826. KILOWATTS
                   (AFTER CONVERSION TO 13.6KV-AC @ 98% EFF.)
THE NET POWER DELIVERED TO BUSBAR IS                                         675000. KILOWATTS

THE VOLTAGE OF THE FUEL CELL IS         0.8350 VOLTS.
THE AREA OF THE FUEL CELL IS          5268847. SQ. FT.

OVERALL PLANT EFFICIENCY, COAL PILE TO BUSBAR, IS 47.3 PERCENT.
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TABULATION OF SHAFT ENERGY AVAILABLE OR NEEDED.

 102  -36546.9 KILOWATTS     AIR COMPRESSOR FOR SUPPLY TO CRYOGENIC OXYGEN PLANT
 104  -15066.7 KILOWATTS     CRYOGENIC OXYGEN COMPRESSOR FOR SUPPLY TO COAL GASIFIER
   2   -4806.6 KILOWATTS     2 U-GAS GASIFIER NORMAL OPERATION. STREAM #2 = NET MAKE.
   5   50794.8 KILOWATTS     5 EXPAND HOT RAW GAS VIA TURBINE; COLLECT KILOWATTS.
  21   -6097.0 KILOWATTS     21 REMOVE SULFUR FROM RAW GAS BY SELEXOL SYSTEM.
  34    9746.6 KILOWATTS     34 ADD STEAM VIA TURBINE; PREVENT CARBON DROP IN FUEL CELL.
  36 -156923.6 KILOWATTS     36 COMPRESS COMBUSTION AIR TO FUEL CELL PRESSURE.
  43  613087.3 KILOWATTS     43 THE FUEL CELL, MAXIMUM CONVERSION, PEFORM WITHIN ANODE.
  48   -7636.0 KILOWATTS     48 ANODE EXIT GAS BOOSTER FAN, TO FEED CATALYTIC BURNER.
  54  -13637.0 KILOWATTS     54 CATHODE RECYCLE GAS CIRCULATION FAN.
  55  257883.3 KILOWATTS     55 EXPAND STACK GAS THROUGH TURBINE TO RECOVER POWER.
  63   -1649.9 KILOWATTS     63 COMPRESS RECYCLE GAS TO INJECT FEED COAL TO GASIFIER.
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                                                                                        CASE    L.QEX.2
L.QEX.2   HIGH TEMPERATURE DE-DUSTING, RAW GAS LIKE JPL4, QUENCH AFTER TURBINE.               STEAM LIST
                                                                                        17 FEB 83  7:33:26

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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.QEX.2   HIGH TEMPERATURE DE-DUSTING, RAW GAS LIKE JPL4, QUENCH AFTER TURBINE.                    CASE    L.QEX.2
                                                                                                      WASTE HEAT OUT
                                                                                                   17 FEB 83  7:33:26

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TABULATION OF WASTE HEAT AVAILABLE FROM PROCESS BOILERS.

  11      600.0 DEGREES F.     310.85 MILLIONS BTU    11 COOL SHIFT FEED GAS TO 600F, RAISE STEAM.
  13      600.0 DEGREES F.     109.45 MILLIONS BTU    13 COOL INTER-STAGE SHIFT GAS IN WHB; RAISE STEAM.
  46      925.0 DEGREES F.     307.26 MILLIONS BTU    46 COOL ANODE EXIT GAS IN WHB - RAISE STEAM.
  51     1115.0 DEGREES F.     171.96 MILLIONS BTU    51 COOL CATHODE RECYCLE GAS IN WHB - RAISE STEAM.
```

CASE 7

Employ Sulfur-Tolerant Fuel Cell

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                                                                                                  CASE    L.THICEL
L.THICEL  SULFUR-TOLERANT CELL, 85X CONV, INT. REFORM, NO PRECOOL, 13.5X CH4 RAW                        SUMMARY
                                                                                                  23 FEB 83 10:11:20
             UTILITIES TABULATION FOR SOUR GAS FUEL CELL.

  THIS IS THE FINAL SYNOPSIS OF THE ENERGY BUDGET OF AN INTEGRATED POWER PLANT USING COAL
GASIFICATION AND ELECTROCHEMICAL FUEL CELLS, ALONG WITH CONVENTIONAL POWER GENERATION MACHINERY.

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        *********                                                         *********
        *********         COAL CONSUMPTION IS   4334. TONS PER DAY.       *********
        *********                                                         *********
        *********                                                         *********
        *********                                                         *********
        *********         NET DELIVERED POWER IS  675. MEGAWATTS.         *********
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---------------------------------------------DETAILS---------------------------------------
COAL CONSUMPTION IS  361. THOUSANDS POUNDS PER HOUR, OR ABOUT  4615. MILLIONS BTU.
THE OXYGEN PLANT SIZE IS  2625. TONS PER DAY.
PROCESS STEAM CONSUMPTION IS   377. THOUSANDS POUNDS PER HOUR.

PROCESS MACHINERY (COMPRESSORS AND MAJOR PUMPS) NEED                         -260697. KILOWATTS
PROCESS GAS TURBO-EXPANDERS & CLOSELY-ASSOCIATED STEAM TURBINES GIVE          307277. KILOWATTS
CONDENSING STEAM TURBINE (MULTI-STAGE W. REHEAT) GIVES                         73213. KILOWATTS
THE FUEL CELL ITSELF DELIVERS                                                 555206. KILOWATTS
                    (AFTER CONVERSION TO 13.6KV-AC & 98X EFF.)
THE NET POWER DELIVERED TO BUSBAR IS                                          675000. KILOWATTS

THE VOLTAGE OF THE FUEL CELL IS           0.8148 VOLTS.
THE AREA OF THE FUEL CELL IS            4989556. SQ. FT.

OVERALL PLANT EFFICIENCY, COAL PILE TO BUSBAR, IS 49.9 PERCENT.
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L.THICEL  SULFUR-TOLERANT CELL, 85% CONV, INT. REFORM, NO PRECOOL, 13.5% CH4 RAW                CASE    L.THICEL
                                                                                                   POWER LIST
                                                                                                 23 FEB 83 10:11:20

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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    60251.5 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   566537.0 KILOWATTS     29 THE FUEL CELL, MAXIMUM CONVERSION, REFORM WITHIN ANODE.
  33    -6568.2 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.THICEL  SULFUR-TOLERANT CELL, 85% CONV, INT. REFORM, NO PRECOOL, 13.5% CH4 RAW              CASE     L.THICEL
                                                                                                 STEAM LIST
                                                                                              23 FEB 83 10:11:20

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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    1010.0 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 CO_2 Transfer

```
                                                                                            CASE    L.RUN.D2
L.RUN.D2  HIGH TEMP. CO2 PASS ADDED TO BASE CASE, COMPARE L.AIR.4E                               SUMMARY
                                                                                          1 MAR 83 14:03:06
             HIGH-TEMPERATURE CO2 TRANSFER ADDED TO BASE CASE.

  THIS IS THE FINAL SYNOPSIS OF THE ENERGY BUDGET OF AN INTEGRATED POWER PLANT USING COAL
GASIFICATION AND ELECTROCHEMICAL FUEL CELLS, ALONG WITH CONVENTIONAL POWER GENERATION MACHINERY.

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        *********                                                         *********
        *********                                                         *********
        *********                                                         *********
        *********                                                         *********
        *********         COAL CONSUMPTION IS   4564. TONS PER DAY.       *********
        *********                                                         *********
        *********                                                         *********
        *********                                                         *********
        *********         NET DELIVERED POWER IS  675. MEGAWATTS.         *********
        *********                                                         *********
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---------------------------------------------DETAILS_______________________________________

COAL CONSUMPTION IS  380. THOUSANDS POUNDS PER HOUR, OR ABOUT  4860. MILLIONS BTU.
THE OXYGEN PLANT SIZE IS  2764. TONS PER DAY.
PROCESS STEAM CONSUMPTION IS   683. THOUSANDS POUNDS PER HOUR.

PROCESS MACHINERY (COMPRESSORS AND MAJOR PUMPS) NEED                    -244513. KILOWATTS
PROCESS GAS TURBO-EXPANDERS & CLOSELY-ASSOCIATED STEAM TURBINES GIVE     252006. KILOWATTS
CONDENSING STEAM TURBINE (MULTI-STAGE W. REHEAT) GIVES                   32528. KILOWATTS
THE FUEL CELL ITSELF DELIVERS                                           634979. KILOWATTS
                  (AFTER CONVERSION TO 13.6KV-AC @ 98% EFF.)
THE NET POWER DELIVERED TO BUSBAR IS                                    675000. KILOWATTS

THE VOLTAGE OF THE FUEL CELL IS          0.8851 VOLTS.
THE AREA OF THE FUEL CELL IS          5253598. SQ. FT.

OVERALL PLANT EFFICIENCY, COAL PILE TO BUSBAR, IS 47.4 PERCENT.
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                                                                                          CASE     L.RUN.D2
L.RUN.D2  HIGH TEMP. CO2 PASS ADDED TO BASE CASE, COMPARE L.AIR.4E                          POWER LIST
                                                                                         1 MAR 83 14:03:06

*******************************************************************************************************

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   647938.1 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     81 ANODE EXIT GAS BOOSTER FAN, TO FEED CATALYTIC BURNER.
  82   224662.9 KILOWATTS     82 EXPAND CATHODE EXIT STACK GAS TO RECOVER POWER.
  86    -1371.3 KILOWATTS     86 COMPRESS COAL FEED INJECTION GAS, A SMALL RECYCLE.
```

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                                                                                                 CASE    L.RUN.D2
L.RUN.D2  HIGH TEMP. CO2 PASS ADDED TO BASE CASE, COMPARE L.AIR.4E                                   STEAM LIST
                                                                                                1 MAR 83 14:03:06

*******************************************************************************************************************

TABULATION OF STEAM NEEDED IN PROCESS.

 2       550.0 PSIA          22016. MOLS STEAM  2 U-GAS GASIFIER 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.
```

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L.RUN.D2  HIGH TEMP. CO2 PASS ADDED TO BASE CASE, COMPARE L.AIR.4E                                    CASE     L.RUN.D2
                                                                                                        WASTE HEAT OUT
                                                                                                      1 MAR 83 14:03:06

*********************************************************************************************************************

TABULATION OF WASTE HEAT AVAILABLE FROM PROCESS BOILERS.

   8     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.      33.91 MILLIONS BTU    20 COOL RAW GAS; PRE-HEAT BOILER FEED WATER.
  36      855.9 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.     226.39 MILLIONS BTU    75 COOL CATHODE RECYCLE GAS IN WHB - RAISE STEAM.
  80      880.0 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

```
                                                                                             CASE    L.RUN.C3
L.RUN.C3  HIGH TEMP. H2S & DUST REMOVAL, MAKE MOLTEN SULFUR FROM OXYGEN.                          SUMMARY
                                                                                           9 MAR 83 13:31:39
            HIGH-TEMPERATURE H2S TRANSFER ADDED TO BASE CASE.

  THIS IS THE FINAL SYNOPSIS OF THE ENERGY BUDGET OF AN INTEGRATED POWER PLANT USING COAL
GASIFICATION AND ELECTROCHEMICAL FUEL CELLS, ALONG WITH CONVENTIONAL POWER GENERATION MACHINERY.

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          *********                                                    *********
          *********                                                    *********
          *********                                                    *********
          *********                                                    *********
          *********        COAL CONSUMPTION IS   4330. TONS PER DAY.   *********
          *********                                                    *********
          *********                                                    *********
          *********                                                    *********
          *********        NET DELIVERED POWER IS  675. MEGAWATTS.     *********
          *********                                                    *********
          *********                                                    *********
          *********                                                    *********
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------------------------------------------------DETAILS------------------------------------------------

COAL CONSUMPTION IS  361. THOUSANDS POUNDS PER HOUR, OR ABOUT  4611. MILLIONS BTU.
THE OXYGEN PLANT SIZE IS  2705. TONS PER DAY.
PROCESS STEAM CONSUMPTION IS   330. THOUSANDS POUNDS PER HOUR.

PROCESS MACHINERY (COMPRESSORS AND MAJOR PUMPS) NEED                       -236765. KILOWATTS
PROCESS GAS TURBO-EXPANDERS & CLOSELY-ASSOCIATED STEAM TURBINES GIVE        285756. KILOWATTS
CONDENSING STEAM TURBINE (MULTI-STAGE W. REHEAT) GIVES                       72054. KILOWATTS
THE FUEL CELL ITSELF DELIVERS                                               553945. KILOWATTS
                (AFTER CONVERSION TO 13.6KV-AC @ 98% EFF.)
THE NET POWER DELIVERED TO BUSBAR IS                                        675000. KILOWATTS

THE VOLTAGE OF THE FUEL CELL IS          0.8136 VOLTS.
THE AREA OF THE FUEL CELL IS           4935361. SQ. FT.

OVERALL PLANT EFFICIENCY, COAL PILE TO BUSBAR, IS 50.0 PERCENT.
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L.RUN.C3  HIGH TEMP. H2S & DUST REMOVAL, MAKE MOLTEN SULFUR FROM OXYGEN.                                          CASE     L.RUN.C3
                                                                                                                     POWER LIST
                                                                                                                  9 MAR 83 13:31:33

****************************************************************************************************************************

TABULATION OF SHAFT ENERGY AVAILABLE OR NEEDED.

 102   -34573.0 KILOWATTS      AIR COMPRESSOR FOR SUPPLY TO CRYOGENIC OXYGEN PLANT
 104   -14253.5 KILOWATTS      CRYOGENIC OXYGEN COMPRESSOR FOR SUPPLY TO COAL GASIFIER
   2    -4544.3 KILOWATTS      2 U-GAS GASIFIER NORMAL OPERATION. STREAM #2 = NET MAKE.
 112      -59.6 KILOWATTS      OXYGEN COMPRESSOR FOR REGENERATING SPINELS FOR HOT H2S DELETE
  15    -1375.9 KILOWATTS      SUPPLY OF  33. T/D EXTRA OXYGEN FOR HOT H2S REMOVAL.
 113     -567.1 KILOWATTS      COMPRESSOR FOR RECYCLING DILUENT GAS; FOR HOT H2S REMOVAL.
  16    41814.2 KILOWATTS      16 EXPAND SULFUR-LEAN RAW GAS TO CELL PRESSURE; GET POWER.
  27 -140553.3 KILOWATTS       27 COMPRESS COMBUSTION AIR TO FUEL CELL PRESSURE.
  35   565359.2 KILOWATTS      NORMAL FUEL CELL, 1200 F INPUT SETS VOLTAGE, 1260 F EXIT.
  42   -25010.1 KILOWATTS      42 CATHODE RECYCLE GAS CIRCULATION FAN.
  46    -6611.8 KILOWATTS      46 ANODE EXIT GAS BOOSTER FAN, TO FEED CATALYTIC BURNER.
  43   243942.0 KILOWATTS      43 EXPAND CATHODE EXIT STACK GAS TO RECOVER POWER.
  55    -1410.6 KILOWATTS      55 COMPRESS COAL FEED INJECTION GAS, A SMALL RECYCLE.
```

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                                                                                           CASE     L.RUN.C3
L.RUN.C3  HIGH TEMP. H2S & DUST REMOVAL, MAKE MOLTON SULFUR FROM OXYGEN.                      STEAM LIST
                                                                                         9 MAR 93 13:31:33

********************************************************************************************************

TABULATION OF STEAM NEEDED IN PROCESS.

 2       550.0 PSIA          20[illegible]. MOLS STEAM  2 U-GAS GASIFIER NORMAL OPERATION. STREAM #2 = NET MAKE.
14      1200.0 PSIA            1[illegible]7. MOLS STEAM  14 ADSORB H2S ONTO ZINC SPINEL; A NEW IGT DEVELOPMENT.
15      1200.0 PSIA             50. MOLS STEAM  15 RECLAIM ZN SPINEL (OXIDE) BY OXYGEN; MAKE SULFUR.
```

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L.RUN.C3  HIGH TEMP. H2S & DUST REMOVAL. MAKE MOLTON SULFUR FROM OXYGEN.                    CASE    L.RUN.C3
                                                                                             WASTE HEAT OUT
                                                                                            9 MAR 83 13:31:38

*****************************************************************************************************

TABULATION OF WASTE HEAT AVAILABLE FROM PROCESS BOILERS.

   8    1500.0 DEGREES F.    192.10 MILLIONS BTU    8 COOL HOT SOUR RAW GAS IN WHB; RAISE STEAM.
  14    1200.0 DEGREES F.    -27.77 MILLIONS BTU   14 ADSORB H2S ONTO ZINC SPINEL; A NEW IGT DEVELOPMENT.
  15    1400.0 DEGREES F.     -3.51 MILLIONS BTU                     COOLING COIL IN FLUID BED SPINEL REGENERATOR.
  15     250.0 DEGREES F.     13.22 MILLIONS BTU                       SULFUR CONDENSER WHICH GENERATES 15 # STEAM.
  19     750.0 DEGREES F.     32.93 MILLIONS BTU   19 COOL SULFUR-LEAN RAW GAS; MAKE STEAM.
  41     350.0 DEGREES F.    602.62 MILLIONS BTU   41 COOL CATHODE RECYCLE GAS IN WHB; RAISE STEAM.
  45     350.0 DEGREES F.    254.31 MILLIONS BTU   45 COOL ANODE EXIT GAS IN WHB; RAISE STEAM.
  53     360.0 DEGREES F.     13.20 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	Steam	Net		
		MW					
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 H_2S 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 tuboexpanded 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

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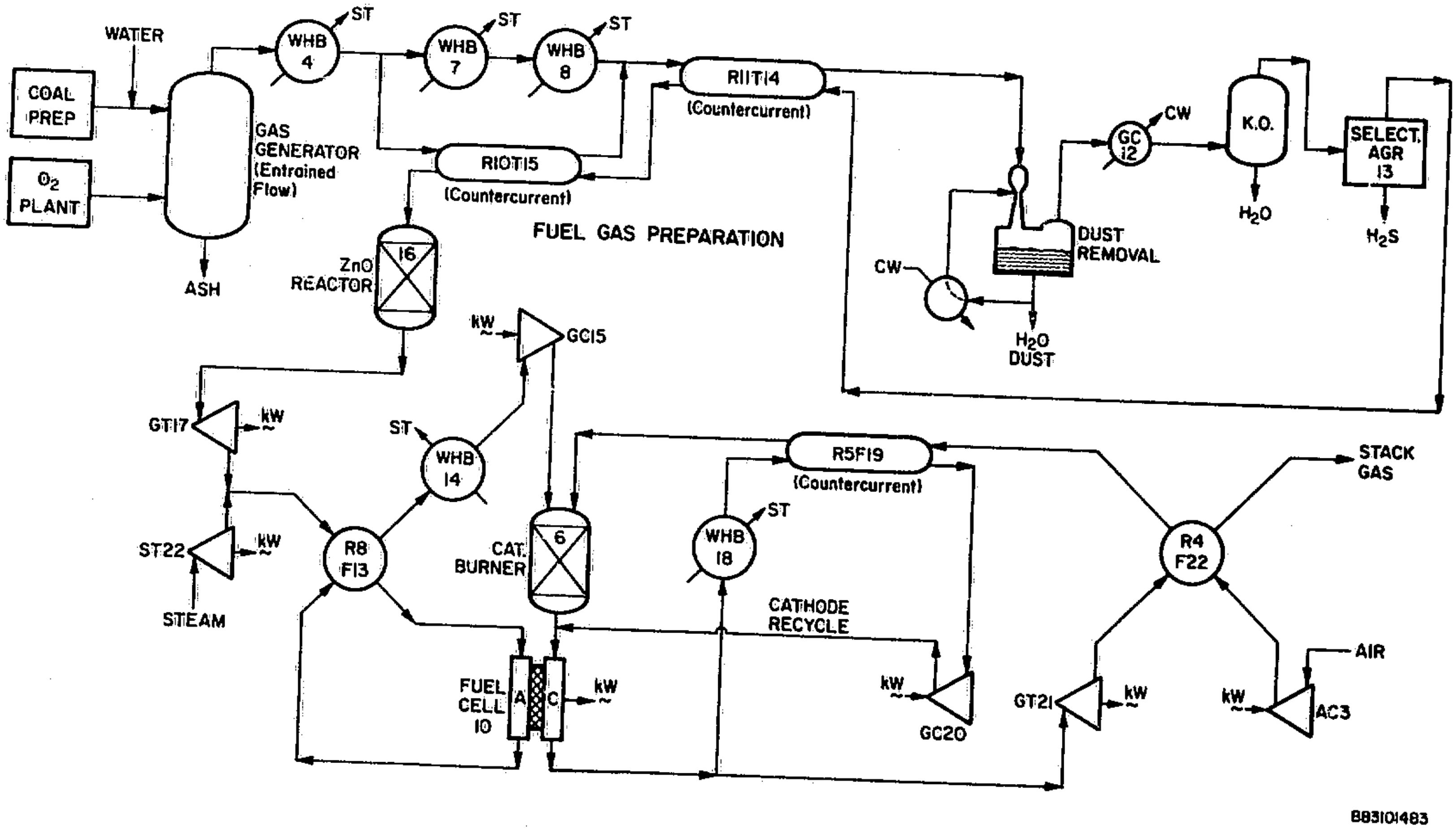

Figure Ga-1. PROCESS FLOW DIAGRAM OF BASE CASE the incoming air. Isentropic compressor efficiencies are 88% and expander efficiencies are 92%. Other assumptions used in the analysis were —

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 stock at approximately 500°F to avoid potential plume formation.

3. The gas purification by a selective Selexol system was specified for 19% H_2S 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 Seloxol 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 is 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 Ie 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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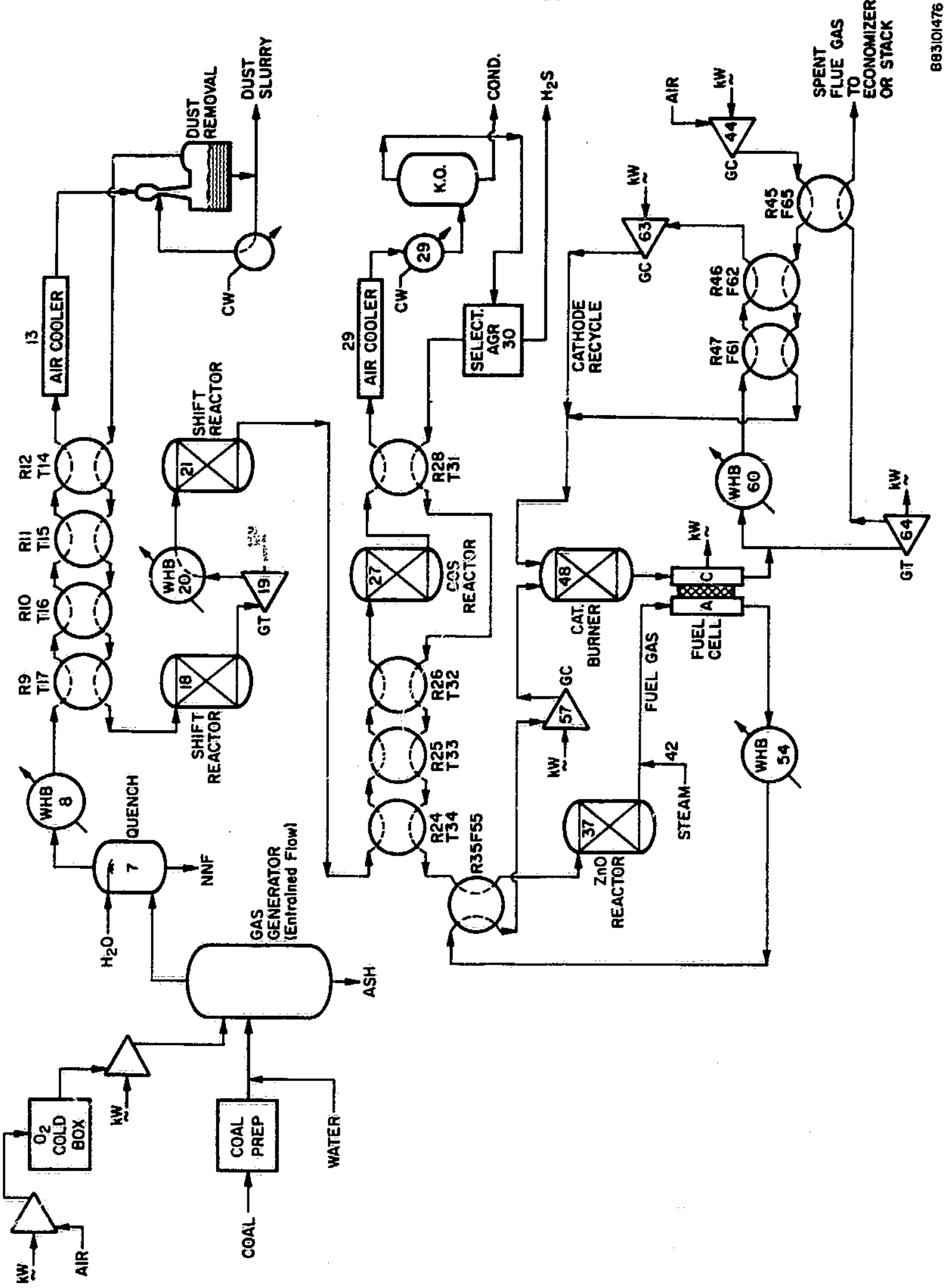

Figure Ca-2. PROCESS FLOW DIAGRAM OF BASE CASE AFTER REDESIGN FOR OPERABILITY

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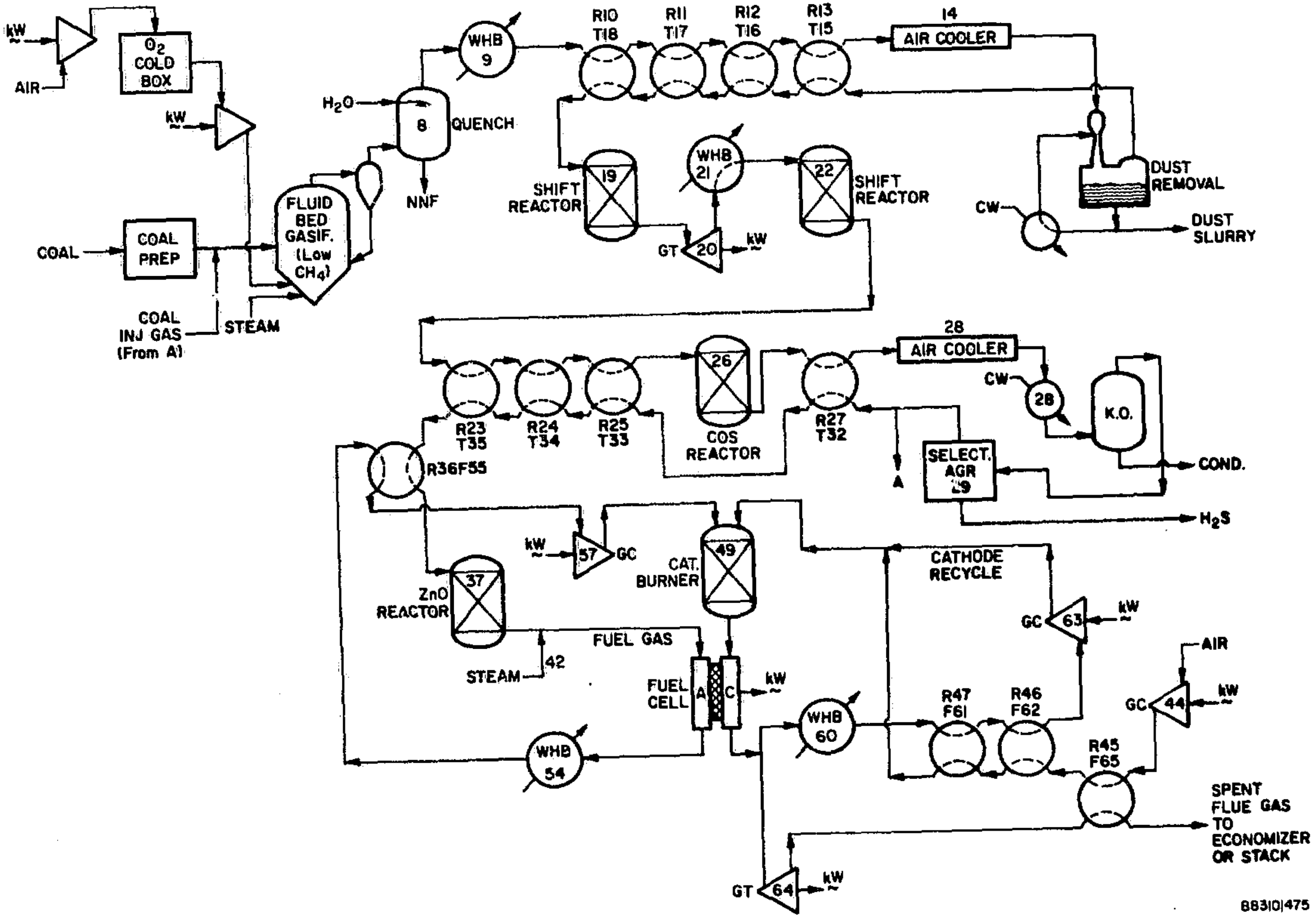

Figure Ca-3. PROCESS FLOW DIAGRAM OF INITIAL EVALUATION WITH FLUIDIZED-BED GASIFICATION technology. Coal injection gas requirements were identified, as well as dust losses and carbon content in the ash, to assure a realistic design.

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 occuring 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

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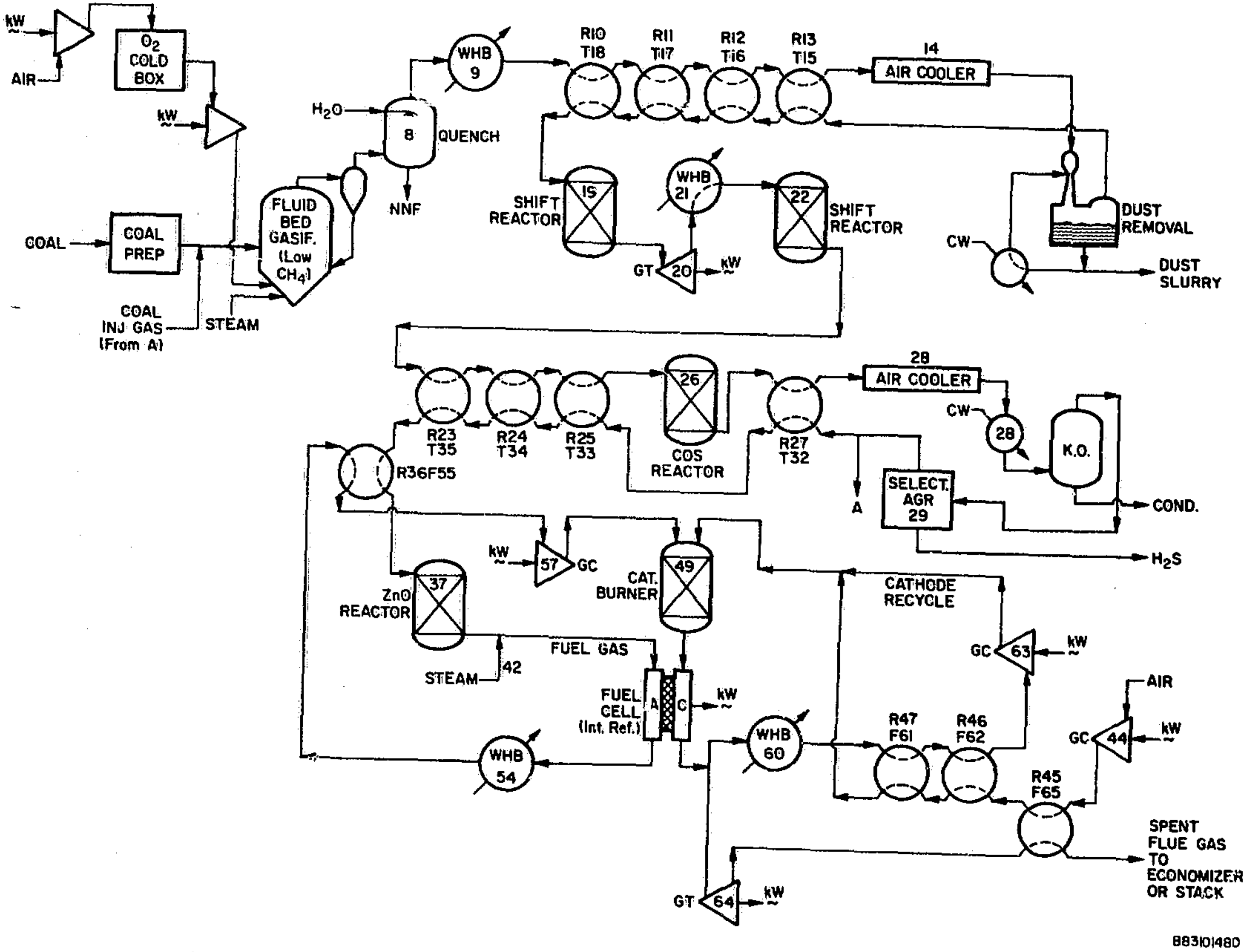

Figure Ca-4. PROCESS FLOW DIAGRAM INCORPORATING FLUIDIZED-BED GASIFICATION AND STEAM-METHANE REFORMING INTEGRATED WITH THE FUEL CELL operating advantage. Therefore, the three gasifiers discussed at this point are sumamrized below in Table Ca-2.

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 If, 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.

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Table Ca-2. GASIFIER SUMMARIES

	Entrained Bed	Fluidized Bed	
		Low Methane	High Methane
Product Gas			
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_2O	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 H_2 = (HHV H_2) (H_2 + CO + $4CH_4$)/HHV Coal.

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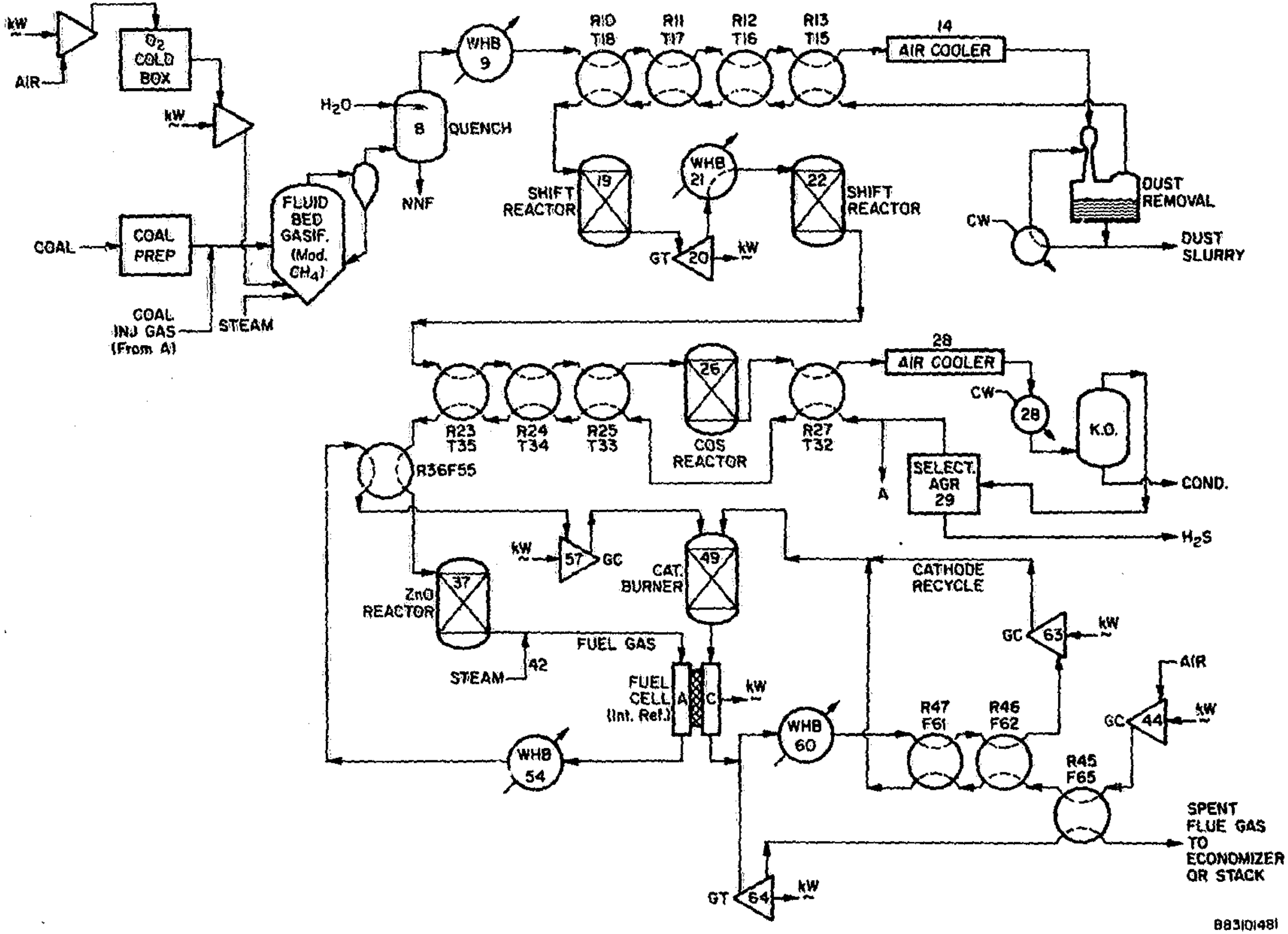

Figure Ca-5. PROCESS FLOW DIAGRAM OF SYSTEM WITH INTERNALLY REFORMING FUEL CELL AND FLUIDIZED-BED GASIFICATION WITH MODERATE METHANE PRODUCTION

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 conceputal 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 CO_2 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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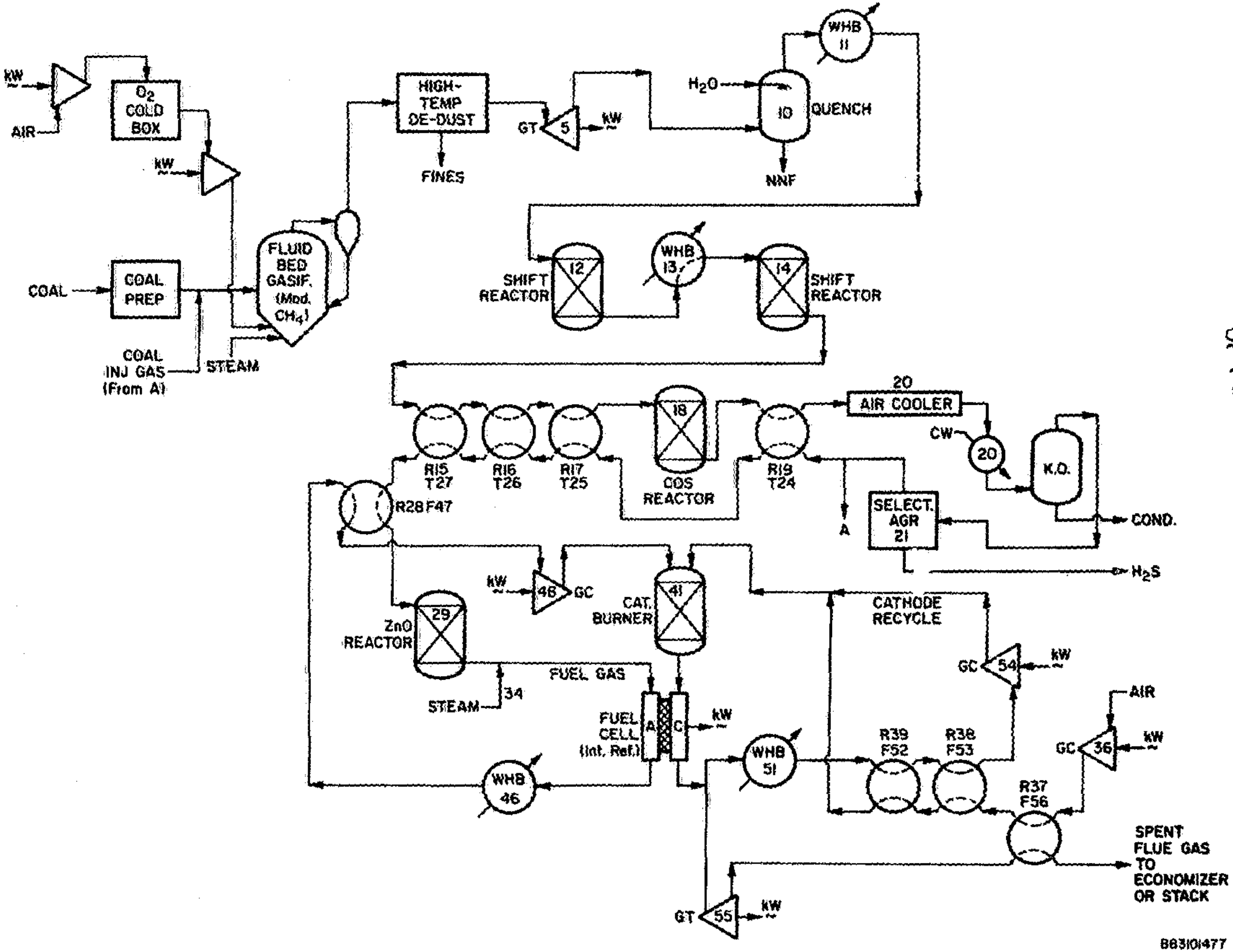

Figure Ca-6. PROCESS FLOW DIAGRAM INCORPORATING HIGH-TEMPERATURE DUST REMOVAL and acid-gas removal before reheating through the single heat exchanger train. The fuel cell incorporates the internal reforming concept.

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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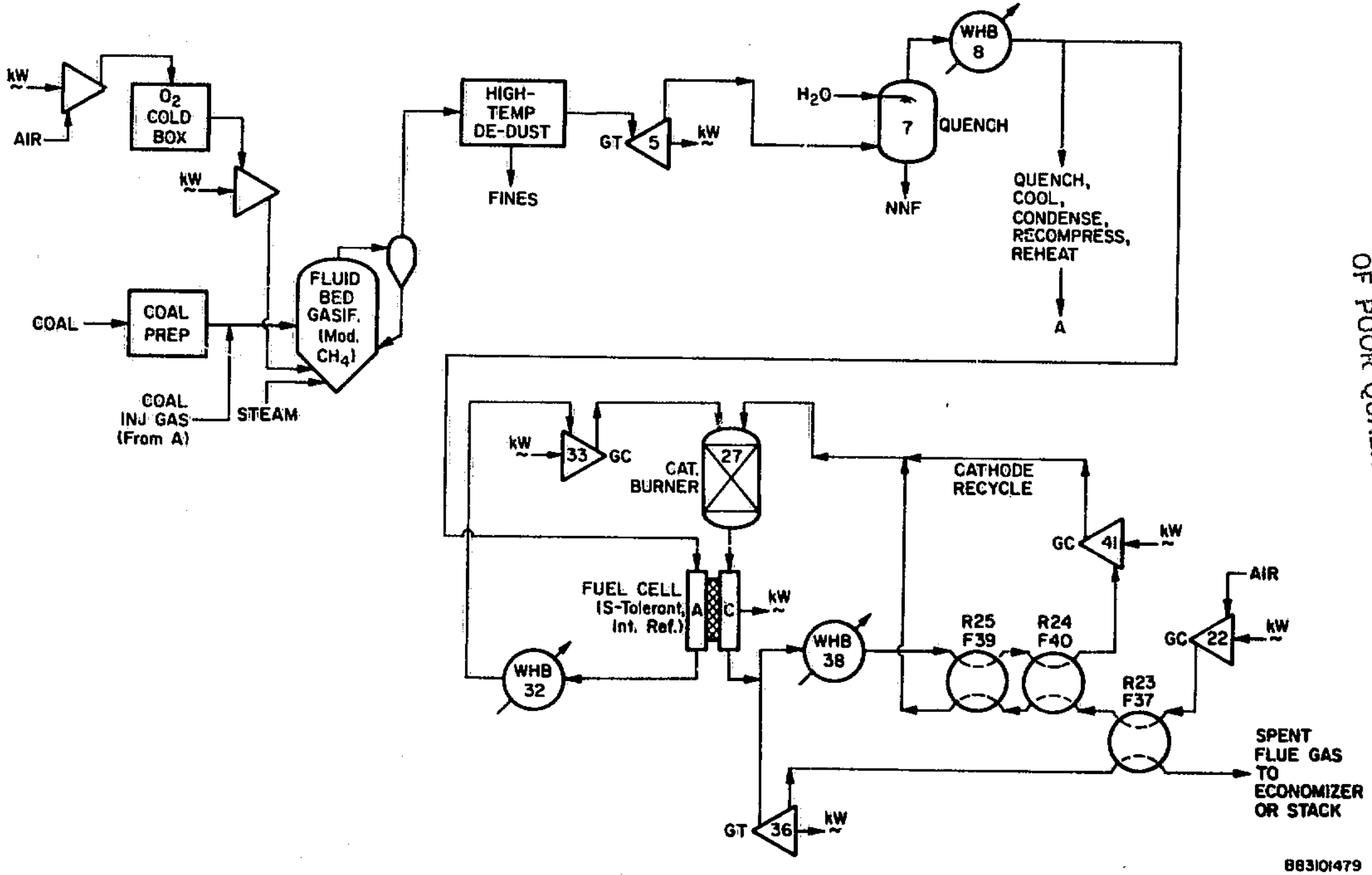

Figure Ca-7. PROCESS FLOW DIAGRAM INCORPORATING SULFUR-TOLERANT FUEL CELL

One note of caution should be considered in the energy analysis of this system: The debits and credits for SO_2 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. Similarily, the carbon dioxide is transferred directly to the cathode where it is required for the chemical reactions.

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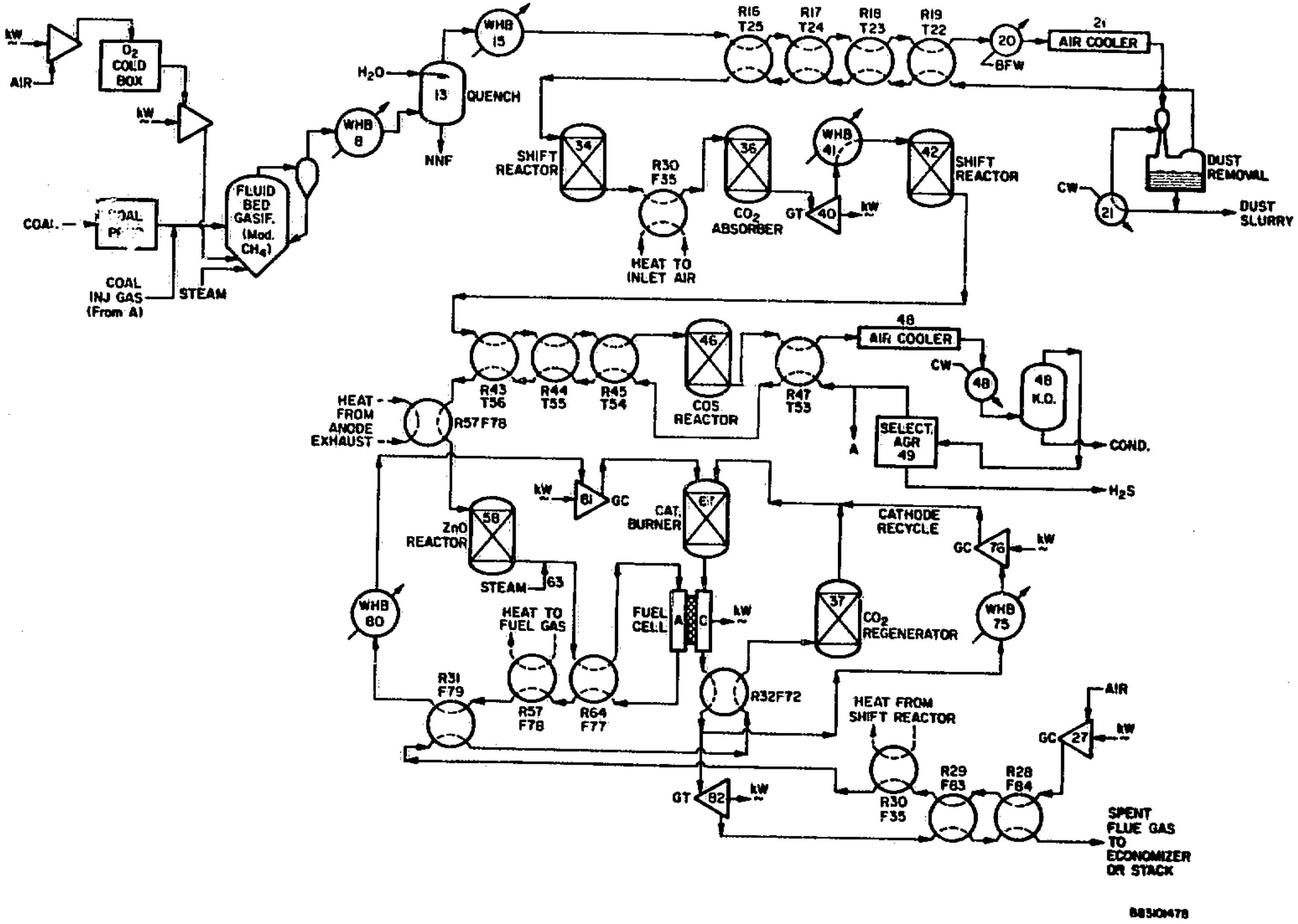

Figure Ca-8. PROCESS FLOW DIAGRAM INCORPORATING HIGH-TEMPERATURE CO_2 TRANSFER

The system with CO_2 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 SO_2 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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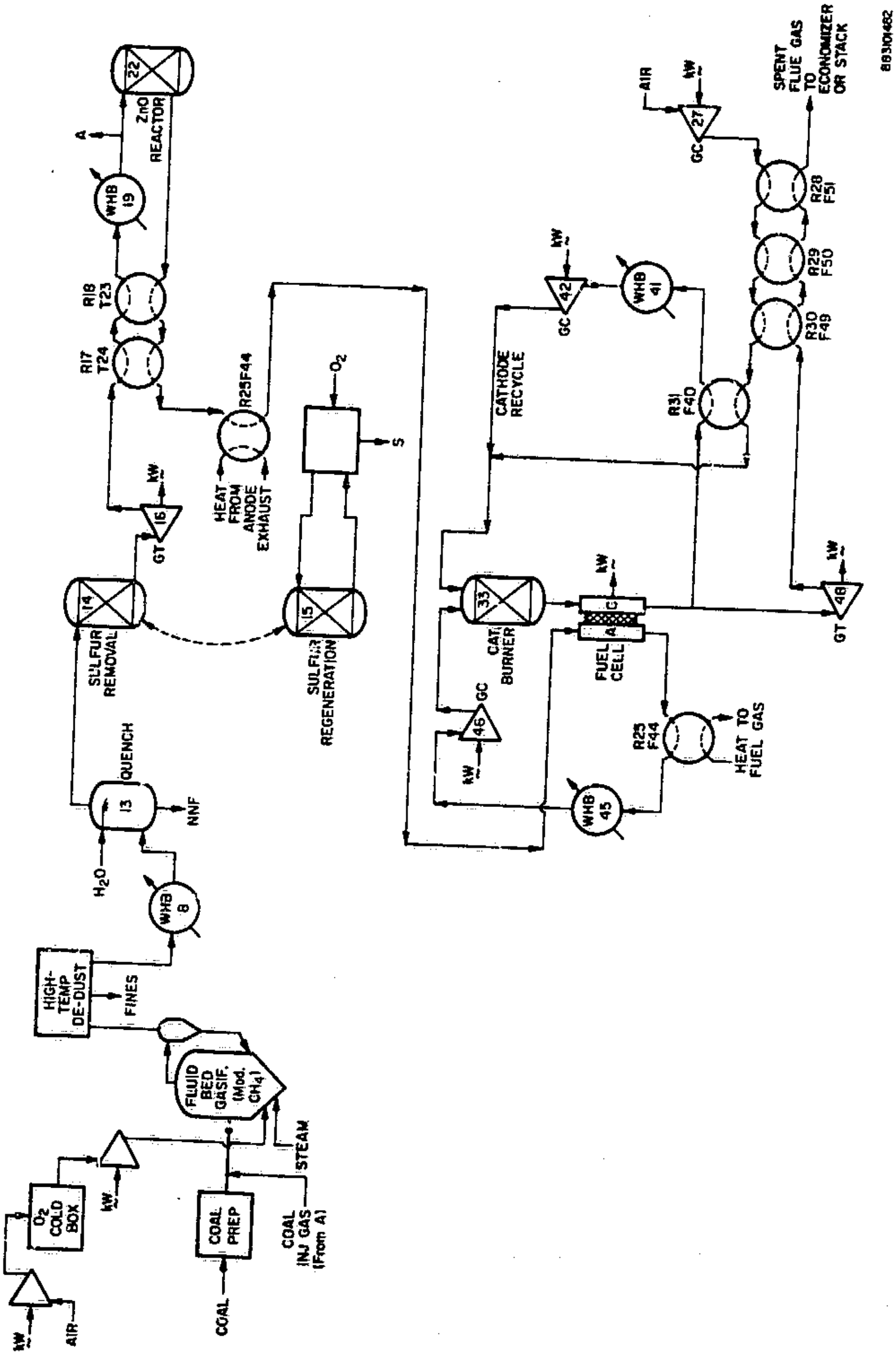

Figure Ca-9. PROCESS FLOW DIAGRAM INCORPORATING HIGH-TEMPERATURE SULFUR REMOVAL

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 considerally 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:

$$H_2 \rightarrow 2H^+ + 2e \qquad \text{(Anode)}$$

$$2H^+ + 1/2\ O_2 + 2e \rightarrow H_2O \qquad \text{(Cathode)}$$

$$H_2 + 1/2\ O_2 \rightarrow H_2O \qquad \text{(Overall)}$$

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:

$$H_2 + CO_3^{--} \rightarrow H_2O + CO_2 + 2e \qquad \text{(Anode)}$$

$$\underline{1/2\ O_2 + CO_2 + 2e \rightarrow CO_3^{--}} \qquad \text{(Cathode)}$$

$$H_2 + 1/2\ O_2 \rightarrow H_2O \qquad \text{(Overall)}$$

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 $(H_2)/(H_2O) \cdot (CO_2)$. 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 counterbalanced 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/cm^2.

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/cm^2 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 CO_2 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 CO_2-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 <u>not</u> 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:

$$C + H_2O \rightarrow H_2 + CO$$

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:

$$C + O_2 \rightarrow CO_2$$

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. Similarily, 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.

$$C + 2H_2 \rightarrow CH_4$$

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:

$$2CO \rightarrow C + CO_2$$

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 —

$$2CO \rightarrow C + CO_2$$

$$H_2 + CO \rightarrow C + H_2O$$

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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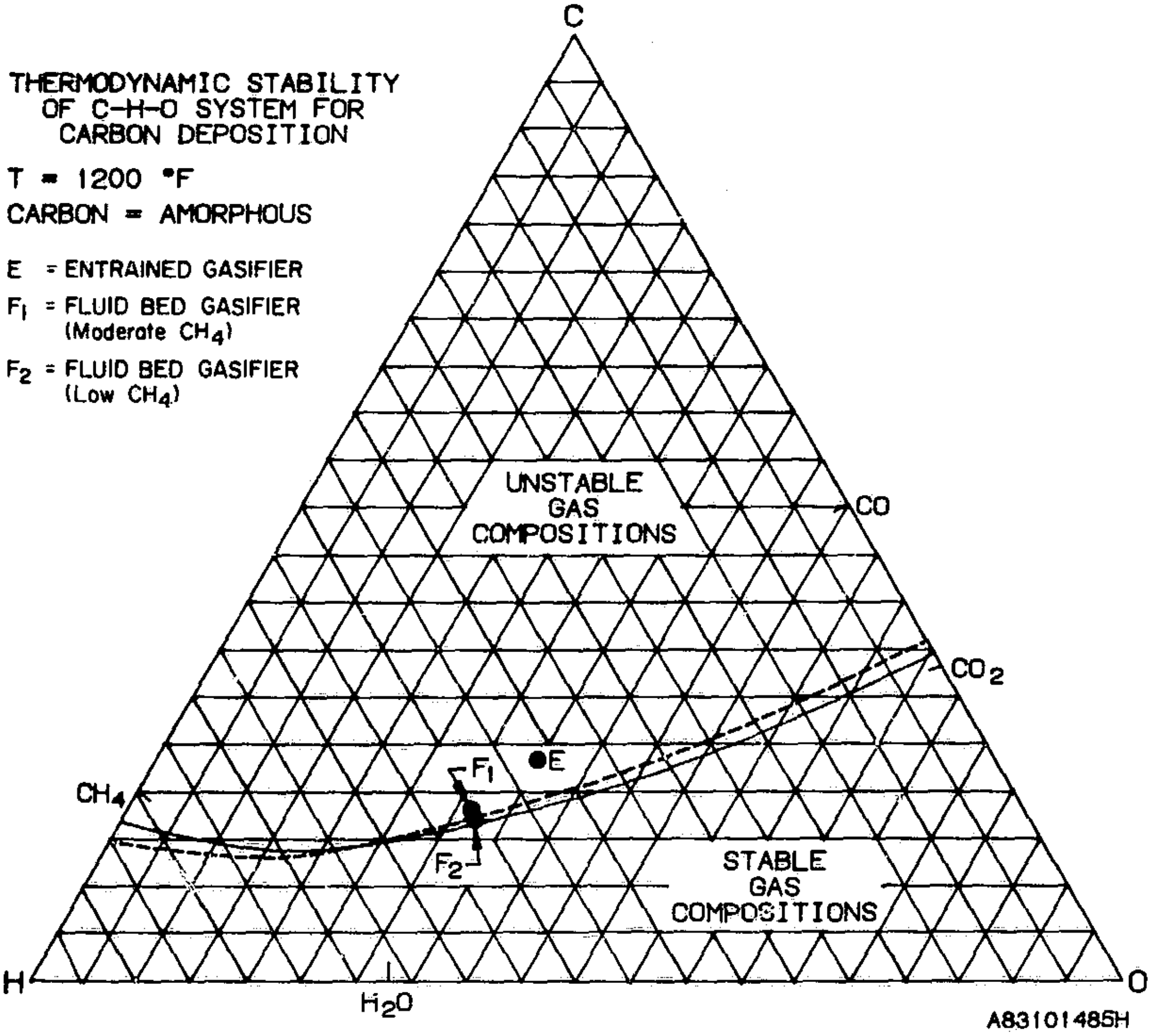

Figure Cb-1. THERMODYNAMIC STABILITY OF FUEL GAS MIXTURES

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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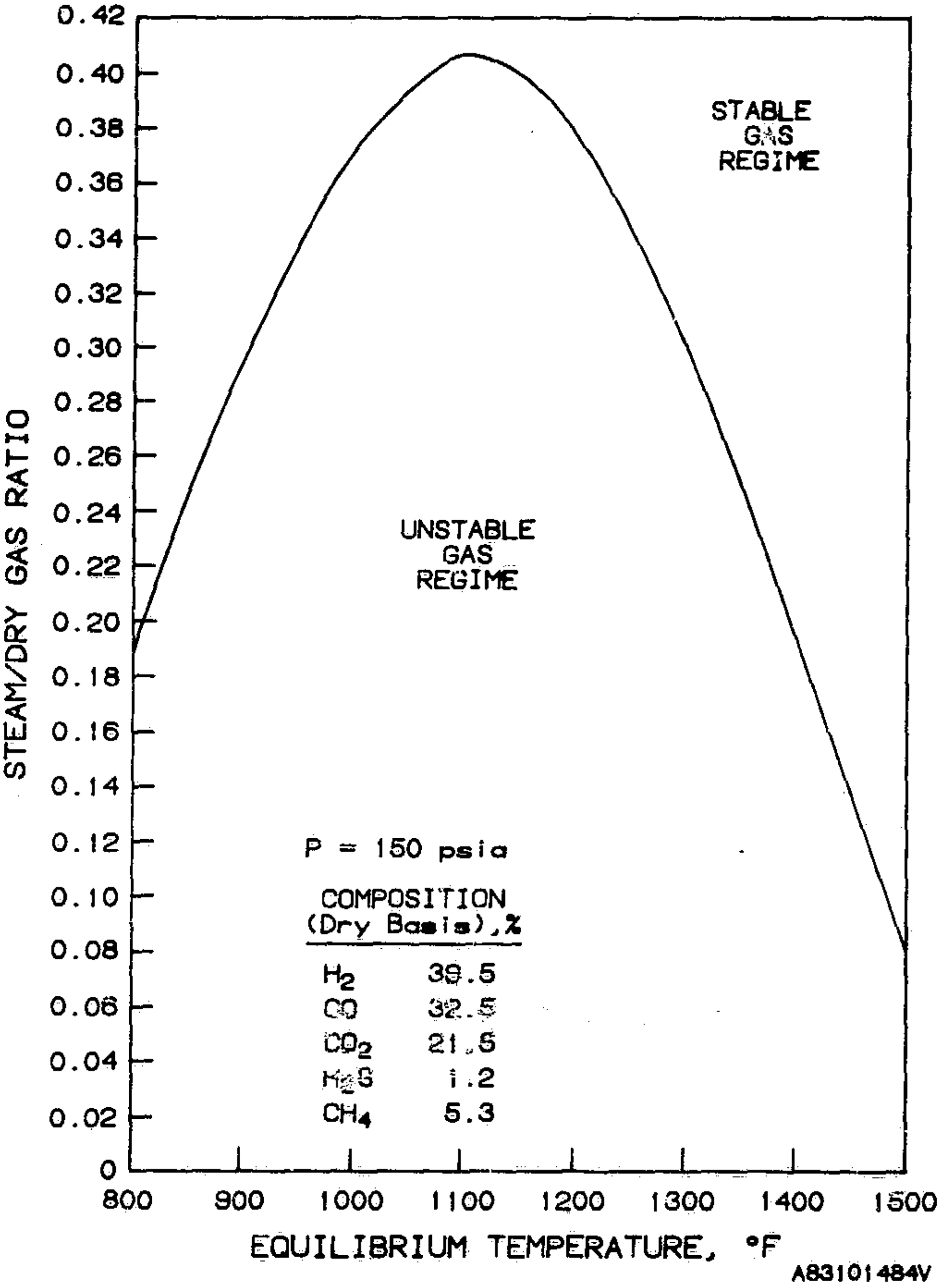

Figure Cb-2. EFFECT OF WATER ADDITION ON STABILITY OF A FUEL GAS MIXTURE

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:

$$CO + H_2O \rightarrow H_2 + CO_2$$

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.

C-3

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:

$$CH_4 + H_2O \rightarrow 3H_2 + CO$$

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:

$$\begin{array}{lr}
C + H_2O \rightarrow CO + H_2 & \text{(gasification)} \\
CO + H_2O \rightarrow H_2 + CO_2 & \text{(water-gas shift)} \\
2H_2 + 2CO_3^{--} \rightarrow 2H_2O + 2CO_2 + 4e & \text{(anode)} \\
\underline{O_2 + 2CO_2 + 4e \rightarrow 2CO_3^{--}} & \text{(cathode)} \\
C + O_2 \rightarrow CO_2 & \text{(overall)}
\end{array}$$

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:

$$\begin{array}{lr}
C + 2H_2 \rightarrow CH_4 & \text{(hydrogasfication)} \\
CH_4 + H_2O \rightarrow 3H_2 + CO & \text{(steam-methane reforming)} \\
CO + H_2O \rightarrow CO_2 + H_2 & \text{(water-gas shift)} \\
2H_2 + 2CO_3^{--} \rightarrow 2H_2O + 2CO_2 + 4e & \text{(anode)} \\
\underline{O_2 + 2CO_2 + 4e \rightarrow 2CO_3^{--}} & \text{(cathode)} \\
C + O_2 \rightarrow CO_2 & \text{(overall)}
\end{array}$$

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°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 sweeten- 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 order 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 order 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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                                                                                    CASE  1
1  JPL.1  BASE CASE  65% CONV., NO QUENCH, NO PRE-SHIFT, ISO EFF: COMP 98% T 92%          SUMMARY
                                                                                     18 MAR 83  9:46:11
           ORIGINAL BASE CASE AS SPECIFIED BY JET PROPULSION LABS.

  THIS IS THE FINAL SYNOPSIS OF THE ENERGY BUDGET OF AN INTEGRATED POWER PLANT USING COAL
GASIFICATION AND ELECTROCHEMICAL FUEL CELLS, ALONG WITH CONVENTIONAL POWER GENERATION MACHINERY.

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        ***********************************************************************
        ***********************************************************************
        ***********************************************************************
        *********                                                     *********
        *********                                                     *********
        *********                                                     *********
        *********                                                     *********
        *********        COAL CONSUMPTION IS   4989. TONS PER DAY.    *********
        *********                                                     *********
        *********                                                     *********
        *********                                                     *********
        *********        NET DELIVERED POWER IS  656. MEGAWATTS.      *********
        *********                                                     *********
        *********                                                     *********
        *********                                                     *********
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------------------------------------------DETAILS--------------------------------------
COAL CONSUMPTION IS  416. THOUSANDS POUNDS PER HOUR, OR ABOUT  5313. MILLIONS BTU.
THE OXYGEN PLANT SIZE IS  4212. TONS PER DAY.
PROCESS STEAM CONSUMPTION IS   330. THOUSANDS POUNDS PER HOUR.

PROCESS MACHINERY (COMPRESSORS AND MAJOR PUMPS) NEED                -281946. KILOWATTS
PROCESS GAS TURBO-EXPANDERS & CLOSELY-ASSOCIATED STEAM TURBINES GIVE  289396. KILOWATTS
CONDENSING STEAM TURBINE (MULTI-STAGE W. REHEAT) GIVES                110701. KILOWATTS
THE FUEL CELL ITSELF DELIVERS                                         537447. KILOWATTS
                 (AFTER CONVERSION TO 13.8KV-AC @ 98% EFF.)
THE NET POWER DELIVERED TO BUSBAR IS                                  655598. KILOWATTS

THE VOLTAGE OF THE FUEL CELL IS          0.8232 VOLTS.
THE AREA OF THE FUEL CELL IS           4454817. SQ. FT.

OVERALL PLANT EFFICIENCY, COAL PILE TO BUSBAR, IS 42.1 PERCENT.
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1  JPL.1  BASE CASE  85% CONV., NO QUENCH, NO PRE-SHIFT, ISO EFF: COMP 88% T 92%           CASE    1  JPL.1
                                                                                            POWER LIST
                                                                                          19 MAR 83  9:46:11

*********************************************************************************************************************

TABULATION OF SHAFT ENERGY AVAILABLE OR 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. LABS. BASE CASE.
  13    -1106.4 KILOWATTS     13 REMOVE SULFUR FROM RAW GAS BY SELEXOL SYSTEM.
  17    22954.3 KILOWATTS     17 EXPAND SWEETENED RAW GAS TO FUEL CELL PRESSURE.
  22    21431.5 KILOWATTS     22 BLOW STEAM DOWN POWER TURBINE. ADD TO PROCESS GAS (# 17).
   3 -147714.5 KILOWATTS     3 COMPRESS COMBUSTION AIR TO FUEL CELL PRESSURE.
  10   548415.4 KILOWATTS     10 THE FUEL CELL, 85% CONVERSION, SHIFT ONLY WITHIN ANODE.
  15    -8937.6 KILOWATTS     15 ANODE EXIT GAS BOOSTER FAN
  20   -32677.0 KILOWATTS     20 CATHODE RECYCLE GAS CIRCULATION FAN.
  21   245109.6 KILOWATTS     21 EXPAND STACK GAS THROUGH TURBINE TO RECOVER POWER.
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1  JPL.1  BASE CASE  85% CONV., NO QUENCH, NO PRE-SHIFT, ISO EFF: COMP 88% T 92%            CASE     1  JPL.1
                                                                                               STEAM LIST
                                                                                           18 MAR 83  9:45:11

******************************************************************************************************************

TABULATION OF STEAM NEEDED IN PROCESS.

13          65.0 PSIA             4747. MOLS STEAM  13 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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                                                                                          CASE     1  JPL.1
1  JPL.1  BASE CASE  85% CONV., NO QUENCH, NO PRE-SHIFT, ISO EFF: COMP 89% T 92%         WASTE HEAT OUT
                                                                                          18 MAR 83  9:46:11

*************************************************************************************************************

TABULATION OF WASTE HEAT AVAILABLE FROM PROCESS BOILERS.

   4     800.0 DEGREES F.    443.44 MILLIONS BTU    4 COOL RAW GAS TO 800 DEGREES FAHRENHEIT.
   7     600.0 DEGREES F.     19.39 MILLIONS BTU    7 WASTE HEAT BOILER PARALLEL TO RAW GAS RECUP. EXCH. TRAIN.
  14     850.0 DEGREES F.    196.43 MILLIONS BTU    14 COOL ANODE EXIT GAS IN WHB - RAISE STEAM.
  18    1050.0 DEGREES F.    546.37 MILLIONS BTU    18 COOL CATHODE RECYCLE GAS IN WHB - RAISE STEAM.
```

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CASE Id

Conservative Heat Recuperation

```
JPL.TX.1D MODIFIED BASE CASE 85% CONV, NO QUENCH, NO PRE-SHIFT, TRUE HEAT RECUP.                    CASE  JPL.TX.1D
                                                                                                          SUMMARY
                 23 TALLY UTILITIES; REPORT; CLOSE UP SHOP; GO TO NEXT CASE                         5 JAN 83 16:35:31

   THIS IS THE FINAL SYNOPSIS OF THE ENERGY BUDSET OF AN INTEGRATED POWER PLANT USING COAL
  GASIFICATION AND ELECTROCHEMICAL FUEL CELLS, ALONG WITH CONVENTIONAL POWER GENERATION MACHINERY.

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            **********           COAL CONSUMPTION IS   4989. TONS PER DAY.                **********
            **********                                                                    **********
            **********                                                                    **********
            **********                                                                    **********
            **********           NET DELIVERED POWER IS  650. MEGAWATTS.                  **********
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  *****************************************************DETAILS*****************************************************

  COAL CONSUMPTION IS  416. THOUSANDS POUNDS PER HOUR, OR ABOUT  5313. MILLIONS BTU.
  THE OXYGEN PLANT SIZE IS  4212. TONS PER DAY.
  PROCESS STEAM CONSUMPTION IS   379. THOUSANDS POUNDS PER HOUR.

  PROCESS MACHINERY (COMPRESSORS AND MAJOR PUMPS) NEED                    -285126. KILOWATTS
  PROCESS GAS TURBO-EXPANDERS & CLOSELY-ASSOCIATED STEAM TURBINES GIVE     297005. KILOWATTS
  CONDENSING STEAM TURBINE (MULTI-STAGE W. REHEAT) GIVES                   104636. KILOWATTS
  THE FUEL CELL ITSELF DELIVERS                                            533837. KILOWATTS
             (AFTER CONVERSION TO 13.6KV-AC @ 98% EFF.)
  THE NET POWER DELIVERED TO BUSBAR IS                                     650352. KILOWATTS

  THE VOLTAGE OF THE FUEL CELL IS          0.8227 VOLTS.
  THE AREA OF THE FUEL CELL IS           4454817. SQ. FT.

  OVERALL PLANT EFFICIENCY, COAL PILE TO BUSBAR, IS 41.8 PERCENT.
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JPL.TX.1D MODIFIED BASE CASE 85% CONV, NO QUENCH, NO PRE-SHIFT, TRUE HEAT RECUP.            CASE    JPL.TX.1D
                                                                                                POWER LIST
                                                                                            5 JAN 83 16:35:31

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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.
  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   21431.5 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  544731.2 KILOWATTS     10 THE FUEL CELL, 85% CONVERSION, SHIFT ONLY WITHIN ANODE.
  15   -9939.6 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 RECUP.          CASE    JPL.TX.1D
                                                                                            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.    613.09 MILLIONS BTU    4 COOL RAW GAS TO 900 DEGREES FAHRENHEIT, MAKE STEAM.
  14      950.0 DEGREES F.    185.43 MILLIONS BTU   14 COOL ANODE EXIT GAS IN WHB - RAISE STEAM.
  18     1050.0 DEGREES F.    663.98 MILLIONS BTU   18 COOL CATHODE RECYCLE GAS IN WHB - RAISE STEAM.
```

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CASE Ie

Carbon Control and Shift

```
JPL.TX.1E MODIFIED BASE CASE 85% CONV, WITH QUENCH, PRE-SHIFT, & COS DESTRUCTION          CASE  JPL.TX.1E
                                                                                            SUMMARY
                                                                                     5 JAN 83 16:49:02
            23 TALLY UTILITIES; REPORT; CLOSE UP SHOP; GO TO NEXT CASE

  THIS IS THE FINAL SYNOPSIS OF THE ENERGY BUDGET OF AN INTEGRATED POWER PLANT USING COAL
GASIFICATION AND ELECTROCHEMICAL FUEL CELLS, ALONG WITH CONVENTIONAL POWER GENERATION MACHINERY.

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         *********          COAL CONSUMPTION IS   4989. TONS PER DAY.          *********
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         *********          NET DELIVERED POWER IS  620. MEGAWATTS.            *********
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----------------------------------------------------DETAILS------------------------------------------

COAL CONSUMPTION IS  416. THOUSANDS POUNDS PER HOUR, OR ABOUT  5313. MILLIONS BTU.
THE OXYGEN PLANT SIZE IS  4212. TONS PER DAY.
PROCESS STEAM CONSUMPTION IS   279. THOUSANDS POUNDS PER HOUR.

PROCESS MACHINERY (COMPRESSORS AND MAJOR PUMPS) NEED                    -271325. KILOWATTS
PROCESS GAS TURBO-EXPANDERS & CLOSELY-ASSOCIATED STEAM TURBINES GIVE     289268. KILOWATTS
CONDENSING STEAM TURBINE (MULTI-STAGE W. REHEAT) GIVES                    64126. KILOWATTS
THE FUEL CELL ITSELF DELIVERS                                            538093. KILOWATTS
                (AFTER CONVERSION TO 13.6KV-AC & 98% EFF.)
THE NET POWER DELIVERED TO BUSBAR IS                                     620161. KILOWATTS

THE VOLTAGE OF THE FUEL CELL IS          0.8292 VOLTS.
THE AREA OF THE FUEL CELL IS          4454930. SQ. FT.

OVERALL PLANT EFFICIENCY, COAL PILE TO BUSBAR, IS 39.5 PERCENT.
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JPL.TX.1E MODIFIED BASE CASE 85% CONV, WITH QUENCH, PRE-SHIFT, & COS DESTRUCTION          CASE    JPL.TX.1E
                                                                                             POWER LIST
                                                                                          5 JAN 83 16:48:02

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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.
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                                                                                              CASE    JPL.TX.1E
JPL.TX.1E MODIFIED BASE CASE 85% CONV, WITH QUENCH, PRE-SHIFT, 3 COS DESTRUCTION             STEAM LIST
                                                                                           5 JAN 83 16:49:02

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TABULATION OF STEAM NEEDED IN PROCESS.

27        65.0 PSIA           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
                                                                                                5 JAN 83 16:48:02
*****************************************************************************************************************

TABULATION OF WASTE HEAT AVAILABLE FROM PROCESS BOILERS.

   8      900.0 DEGREES F.     198.21 MILLIONS BTU     8 COOL RAW GAS TO 900 DEGREES FAHRENHEIT, MAKE STEAM.
  20      600.0 DEGREES F.     160.55 MILLIONS BTU    20 COOL SHIFTED EXIT GAS; RAISE STEAM.
  14      850.0 DEGREES F.     192.92 MILLIONS BTU    14 COOL ANODE EXIT GAS IN WHE - RAISE STEAM.
  18     1050.0 DEGREES F.     356.04 MILLIONS BTU    18 COOL CATHODE RECYCLE GAS IN WHB - RAISE STEAM.
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CASE If

Optimized by Moving Expander

```
                                                                                          CASE    JPL.TX.1F
JPL.TX.1F MODIFIED BASE CASE 85% CONV, QUENCH, PRE-SHIFT, COSX, TURBO-EXP MOVED.             SUMMARY
                                                                                        5 JAN 83 16:57:51
                  70 TALLY UTILITIES; REPORT; CLOSE UP SHOP; GO TO NEXT CASE

   THIS IS THE FINAL SYNOPSIS OF THE ENERGY BUDGET OF AN INTEGRATED POWER PLANT USING COAL
GASIFICATION AND ELECTROCHEMICAL FUEL CELLS, ALONG WITH CONVENTIONAL POWER GENERATION MACHINERY.

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            **********        COAL CONSUMPTION IS    4991. TONS PER DAY.      **********
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            **********        NET DELIVERED POWER IS  637. MEGAWATTS.         **********
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                                         DETAILS

 COAL CONSUMPTION IS  416. THOUSANDS POUNDS PER HOUR, OR ABOUT  5315. MILLIONS BTU.
 THE OXYGEN PLANT SIZE IS  4213. TONS PER DAY.
 PROCESS STEAM CONSUMPTION IS   307. THOUSANDS POUNDS PER HOUR.

 PROCESS MACHINERY (COMPRESSORS AND MAJOR PUMPS) NEED                    -263830. KILOWATTS
 PROCESS GAS TURBO-EXPANDERS & CLOSELY-ASSOCIATED STEAM TURBINES GIVE     302375. KILOWATTS
 CONDENSING STEAM TURBINE (MULTI-STAGE W. REHEAT) GIVES                    56186. KILOWATTS
 THE FUEL CELL ITSELF DELIVERS                                            542202. KILOWATTS
                   (AFTER CONVERSION TO 13.6KV-AC @ 98% EFF.)
 THE NET POWER DELIVERED TO BUSBAR IS                                     636933. KILOWATTS

 THE VOLTAGE OF THE FUEL CELL IS          0.8352 VOLTS.
 THE AREA OF THE FUEL CELL IS          4456395. SQ. FT.

 OVERALL PLANT EFFICIENCY, COAL PILE TO BUSBAR, IS 40.9 PERCENT.
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JPL.TX.1F MODIFIED BASE CASE 85% CONV, QUENCH, PRE-SHIFT, COSX, TURBO-EXP MOVED.                        CASE     JPL.TX.1F
                                                                                                           POWER LIST
                                                                                                        5 JAN 83 16:57:51
*********************************************************************************************************************

TABULATION OF SHAFT ENERGY AVAILABLE OR 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 FOR JET PROP. LABS. BASE CASE.
  19    43603.7 KILOWATTS     19 EXPAND SHIFT (I) 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 85% CONV, QUENCH, PRE-SHIFT, COSX, TURBO-EXP MOVED.                CASE    JPL.TX.IF
                                                                                                 STEAM LIST
                                                                                               5 JAN 83 16:57:51
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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 MODIFIED BASE CASE 85% CONV, QUENCH, PRE-SHIFT, COSX, TURBO-EXP MOVED.       CASE    JPL.TX.IF
                                                                                          WASTE HEAT OUT
                                                                                         5 JAN 83 16:57:51
*****************************************************************************************************

TABULATION OF WASTE HEAT AVAILABLE FROM PROCESS BOILERS.

   8     700.0 DEGREES F.     264.96 MILLIONS BTU    8 COOL RAW GAS TO 700 DEGREES FAHRENHEIT, MAKE STEAM.
  20     600.0 DEGREES F.      49.27 MILLIONS BTU   20 COOL INTER-STAGE SHIFT GAS IN WHB: RAISE STEAM
  54     920.0 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

```
                                                                                    CASE    JPL.UG.2
JPL.UG.2  FLUID BED BASE CASE 35% CONV, QUENCH, PRE-SHIFT, COSX, 8.33% CH4 RAW          SUMMARY
                                                                                  4 JAN 83 13:36:09
              STANDARD U-GAS GASIFIER WITHOUT METHANE CREDIT IN FUEL CEL

  THIS IS THE FINAL SYNOPSIS OF THE ENERGY BUDGET OF AN INTEGRATED POWER PLANT USING COAL
 GASIFICATION AND ELECTROCHEMICAL FUEL CELLS, ALONG WITH CONVENTIONAL POWER GENERATION MACHINERY.

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          *********          COAL CONSUMPTION IS   4991. TONS PER DAY.  *********
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          *********          NET DELIVERED POWER IS  585. MEGAWATTS.    *********
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                                       DETAILS

 COAL CONSUMPTION IS  416. THOUSANDS POUNDS PER HOUR, OR ABOUT  5315. MILLIONS BTU.
 THE OXYGEN PLANT SIZE IS  3242. TONS PER DAY.
 PROCESS STEAM CONSUMPTION IS   734. THOUSANDS POUNDS PER HOUR.

 PROCESS MACHINERY (COMPRESSORS AND MAJOR PUMPS) NEED                    -306390. KILOWATTS
 PROCESS GAS TURBO-EXPANDERS & CLOSELY-ASSOCIATED STEAM TURBINES GIVE     330730. KILOWATTS
 CONDENSING STEAM TURBINE (MULTI-STAGE W. REHEAT) GIVES                    70822. KILOWATTS
 THE FUEL CELL ITSELF DELIVERS                                           490008. KILOWATTS
                 (AFTER CONVERSION TO 13.8KV-AC @ 98% EFF.)
 THE NET POWER DELIVERED TO BUSBAR IS                                    585171. KILOWATTS

 THE VOLTAGE OF THE FUEL CELL IS            0.8207 VOLTS.
 THE AREA OF THE FUEL CELL IS          4399368. SQ. FT.

 OVERALL PLANT EFFICIENCY, COAL PILE TO BUSBAR, IS 37.6 PERCENT.
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JPL.UG.2  FLUID BED BASE CASE 85% CONV, QUENCH, PRE-SHIFT, COSX, 8.33% CH4 RAW                  CASE    JPL.UG.2
                                                                                                   POWER LIST
                                                                                               4 JAN 83 13:36:09

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TABULATION OF SHAFT 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 #2 = NET MAKE.
  20   32755.2 KILOWATTS      20 EXPAND SHIFT (I) EXIT GAS THROUGH POWER RECOVERY TURBINE.
  29   -7457.1 KILOWATTS      29 REMOVE SULFUR FROM RAW GAS BY SELEXOL SYSTEM.
  42    9064.9 KILOWATTS      42 ADD STEAM VIA TURBINE; PREVENT CARBON DROP IN FUEL CELL.
  44 -171579.4 KILOWATTS      44 COMPRESS COMBUSTION AIR TO FUEL CELL PRESSURE.
  51  500008.1 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   -1932.9 KILOWATTS      70 COMPRESS RECYCLE GAS TO INJECT FEED COAL TO GASIFIER.
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                                                                                      CASE    JPL.UG.2
JPL.UG.2  FLUID BED BASE CASE 85% CONV, QUENCH, PRE-SHIFT, COSX, 9.33% CH4 RAW              STEAM LIST
                                                                                     4 JAN 83 13:36:09

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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          12652. MOLS STEAM  29 REMOVE SULFUR FROM RAW GAS BY SELEXOL SYSTEM.
42      1500.0 PSIA           5811. MOLS STEAM  42 ADD STEAM VIA TURBINE; PREVENT CARBON DROP IN FUEL CELL.
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JPL.UG.2  FLUID BED BASE CASE 85% CONV, QUENCH, PRE-SHIFT, COSX, 9.73% CH4 RAW        CASE    JPL.UG.2
                                                                                       WASTE HEAT OUT
                                                                                      4 JAN 83 13:36:09

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TABULATION OF WASTE HEAT AVAILABLE FROM PROCESS BOILERS.

    9      700.0 DEGREES F.     493.21 MILLIONS BTU     9 COOL RAW GAS TO 700 DEGREES FAHRENHEIT, MAKE STEAM.
   21      800.0 DEGREES F.       9.54 MILLIONS BTU    21 COOL INTER-STAGE SHIFT GAS IN WHB, RAISE STEAM
   54      920.0 DEGREES F.     334.77 MILLIONS BTU    54 COOL ANODE EXIT GAS IN WHB - RAISE STEAM.
   60     1040.0 DEGREES F.     736.54 MILLIONS BTU    60 COOL CATHODE RECYCLE GAS IN WHB - RAISE STEAM.
```

CASE III

Internally Reforming Fuel Cell

```
JPL.UG.3  FLUID BED, HIFORY, 25% CONV, QUENCH, PRE-SHIFT, COSX, 9.37% CH4 RAW          CASE    JPL.UG.3
                                                                                         SUMMARY
              STANDARD U-GAS GASIFIER WITH METHANE REFORMED IN FUEL CELL            4 JAN 83 14:34:31

  THIS IS THE FINAL SYNOPSIS OF THE ENERGY BUDGET OF AN INTEGRATED POWER PLANT USING COAL
 GASIFICATION AND ELECTROCHEMICAL FUEL CELLS, ALONG WITH CONVENTIONAL POWER GENERATION MACHINERY.

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          *********          COAL CONSUMPTION IS  4991. TONS PER DAY.           *********
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          *********          NET DELIVERED POWER IS  695. MEGAWATTS.            *********
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                                          DETAILS

 COAL CONSUMPTION IS  416. THOUSANDS POUNDS PER HOUR, OR ABOUT  5315. MILLIONS BTU.
 THE OXYGEN PLANT SIZE IS  3262. TONS PER DAY.
 PROCESS STEAM CONSUMPTION IS   784. THOUSANDS POUNDS PER HOUR.

 PROCESS MACHINERY (COMPRESSORS AND MAJOR PUMPS) NEED                       -274141. KILOWATTS
 PROCESS GAS TURBO-EXPANDERS & CLOSELY-ASSOCIATED STEAM TURBINES GIVE        322722. KILOWATTS
 CONDENSING STEAM TURBINE (MULTI-STAGE W. REHEAT) GIVES                       11460. KILOWATTS
 THE FUEL CELL ITSELF DELIVERS                                               634517. KILOWATTS
                   (AFTER CONVERSION TO 13.8KV-AC @ 95% EFF.)
 THE NET POWER DELIVERED TO BUSBAR IS                                        694557. KILOWATTS

 THE VOLTAGE OF THE FUEL CELL IS          0.6355 VOLTS.
 THE AREA OF THE FUEL CELL IS           5501176. SQ. FT.

 OVERALL PLANT EFFICIENCY, COAL PILE TO BUSBAR, IS 44.6 PERCENT.
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JPL.UG.3  FLUID BED, REFORM, 95% CONV, QUENCH, PRE-SHIFT, CCSX, 9.33% CH4 RAW                    CASE    JPL.UG.3
                                                                                                      POWER LIST
                                                                                                 4 JAN 81 14:34:31
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TABULATION OF SHAFT 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   -5240.1 KILOWATTS      2 U-GAS GASIFIER NORMAL OPERATION. STREAM 52 = NET MAKE.
  20   33791.4 KILOWATTS      20 EXPAND SHIFT (I) EXIT GAS THROUGH POWER RECOVERY TURBINE.
  29   -7467.1 KILOWATTS      29 REMOVE SULFUR FROM RAW GAS BY SELEXOL SYSTEM.
  42    9064.9 KILOWATTS      42 ADD STEAM VIA TURBINE; PREVENT CARBON DROP IN FUEL CELL.
  44 -170942.1 KILOWATTS      44 COMPRESS COMBUSTION AIR TO FUEL CELL PRESSURE.
  51  647655.3 KILOWATTS      51 THE FUEL CELL, MAXIMUM CONVERSION, REFORM WITHIN ANODE.
  57   -8168.4 KILOWATTS      57 ANODE EXIT GAS BOOSTER FAN, TO FEED CATALYTIC BURNER.
  63  -19551.2 KILOWATTS      63 CATHODE RECYCLE GAS CIRCULATION FAN.
  64  210355.7 KILOWATTS      64 EXPAND STACK GAS THROUGH TURBINE TO RECOVER POWER.
  70   -1921.3 KILOWATTS      70 COMPRESS RECYCLE GAS TO INJECT FEED COAL TO GASIFIER.
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JPL.UG.3  FLUID BED, REFORM, 35% CONV, QUENCH, PRE-SHIFT, COOL, 3.33% CH4 RAW                CASE    JPL.UG.3
                                                                                                  STEAM LIST
                                                                                              4 JAN 83 14:04:31

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TABULATION OF STEAM NEEDED IN PROCESS.

  2      500.0 PSIA          24272. MOLS STEAM  2 U-GAS GASIFIER NORMAL OPERATION. STREAM #2 = NET MAKE.
 29       65.0 PSIA          12552. MOLS STEAM  29 REMOVE SULFUR FROM RAW GAS BY SELEXOL SYSTEM.
 42     1500.0 PSIA           4811. MOLS STEAM  42 ADD STEAM VIA TURBINE; PREVENT CARBON DROP IN FUEL CELL.
```

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JPL.UG.3  FLUID BED, R-FORM, 3ST CONV, QUENCH, PRE-SHIFT, CO3X, 9.33% CH4 RAW                CASE    JPL.UG.3
                                                                                              WASTE HEAT OUT
                                                                                             4 JAN 83 14:04:31
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TABULATION OF WASTE HEAT AVAILABLE FROM PROCESS BOILERS.

   9     700.0 DEGREES F.    495.23 MILLIONS BTU    9 COOL RAW GAS TO 700 DEGREES FAHRENHEIT, MAKE STEAM.
  21     600.0 DEGREES F.      5.73 MILLIONS BTU   21 COOL INTER-STAGE SHIFT GAS IN WH5; RAISE STEAM.
  54     925.0 DEGREES F.    327.70 MILLIONS BTU   54 COOL ANODE EXIT GAS IN WH3 - RAISE STEAM.
  60    1090.0 DEGREES F.    215.92 MILLIONS BTU   60 COOL CATHODE RECYCLE GAS IN WH6 - RAISE STEAM.
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CASE IV

High-Methane Gasifier

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JPL.UG.4  FLUID BED, REFORM, 95% CONV, QUENCH, PRE-SHIFT, COSA, 13.5% CH4 RAW          CASE   JPL.UG.4
                                                                                           SUMMARY
                                                                                      4 JAN 83 14:35:19
              METH-RICH U-GAS WITH ANODE METHANE REFORMING, 13.5% C TO C

  THIS IS THE FINAL SYNOPSIS OF THE ENERGY BUDGET OF AN INTEGRATED POWER PLANT USING COAL
 GASIFICATION AND ELECTROCHEMICAL FUEL CELLS, ALONG WITH CONVENTIONAL POWER GENERATION MACHINERY.

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         *********                                                          *********
         *********         COAL CONSUMPTION IS   4991. TONS PER DAY.        *********
         *********                                                          *********
         *********                                                          *********
         *********                                                          *********
         *********         NET DELIVERED POWER IS  706. MEGAWATTS.          *********
         *********                                                          *********
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                                         DETAILS

 COAL CONSUMPTION IS  416. THOUSANDS POUNDS PER HOUR, OR ABOUT  5314. MILLIONS BTU.
 THE OXYGEN PLANT SIZE IS  3023. TONS PER DAY.
 PROCESS STEAM CONSUMPTION IS   772. THOUSANDS POUNDS PER HOUR.

 PROCESS MACHINERY (COMPRESSORS AND MAJOR PUMPS) NEED                      -269030. KILOWATTS
 PROCESS GAS TURBO-EXPANDERS & CLOSELY-ASSOCIATED STEAM TURBINES GIVE       321056. KILOWATTS
 CONDENSING STEAM TURBINE (MULTI-STAGE W. REHEAT) GIVES                       3653. KILOWATTS
 THE FUEL CELL ITSELF DELIVERS                                              650614. KILOWATTS
                    (AFTER CONVERSION TO 13.6KV-AC & 98% EFF.)
 THE NET POWER DELIVERED TO BUSBAR IS                                       706294. KILOWATTS

 THE VOLTAGE OF THE FUEL CELL IS          0.8294 VOLTS.
 THE AREA OF THE FUEL CELL IS           5744276. SQ. FT.

 OVERALL PLANT EFFICIENCY, COAL PILE TO BUSBAR, IS 45.4 PERCENT.
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JPL.UG.4  FLUID BED, REFORM, 85% CONV, QUENCH, PRE-SHIFT, COSX, 13.5% CH4 RAW                    CASE    JPL.UG.4
                                                                                                   POWER LIST
                                                                                                4 JAN 83 14:35:19
**********************************************************************************************************

TABULATION OF SHAFT ENERGY AVAILABLE OR NEEDED.

  97   -39644.7 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 (T) EXIT GAS THROUGH POWER RECOVERY TURBINE.
  29    -6957.4 KILOWATTS      29 REMOVE SULFUR FROM RAW GAS BY SELEXOL SYSTEM.
  42    10667.5 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   603892.0 KILOWATTS      51 THE FUEL CELL. MAXIMUM CONVERSION. REFORM WITHIN ANODE.
  57    -6324.6 KILOWATTS      57 ANODE EXIT GAS BOOSTER FAN. TO FEED CATALYTIC BURNER.
  63   -19496.7 KILOWATTS      63 CATHODE RECYCLE GAS CIRCULATION FAN.
  64   270032.7 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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JPL.UG.4  FLUID BED. REFORM. HSW CONV. QUENCH. PRE-SHIFT. COSX. 13.5% CH4 RAW            CASE   JPL.UG.4
                                                                                          STEAM LIST
                                                                                       4 JAN 83 14:35:19
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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        11689. 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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                                                                                 CASE      JPL.UG.4
  JPL.UG.4  FLUID BED, REFORM, 85% CONV, QUENCH, PRE-SHIFT, COSA, 13.5% CH4 RAW     WASTE HEAT OUT
                                                                                  4 JAN 83 14:35:19

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  TABULATION OF WASTE HEAT AVAILABLE FROM PROCESS BOILERS.

     9     700.0 DEGREES F.    438.66 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.    319.06 MILLIONS BTU    54 COOL ANODE EXIT GAS IN WHB - RAISE STEAM.
    60    1090.0 DEGREES F.    199.62 MILLIONS BTU    60 COOL CATHODE RECYCLE GAS IN WHB - RAISE STEAM.
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CASE IVa

Conceptual Gasifier

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                                                                              CASE    JPL.UG.4A
JPL.UG.4A FLUID BED, ??????, ??? COND, ??????, PRE-SHIFT, COSX, ??.?% CH4 RAW     SUMMARY
                                                     20%                        5 JAN 83  6:56:52
         METH-RICH U-GAS WITH ANODE METHANE REFORMING, ~~???~~ C TO C

  THIS IS THE FINAL SYNOPSIS OF THE ENERGY BUDGET OF AN INTEGRATED POWER PLANT USING COAL
GASIFICATION AND ELECTROCHEMICAL FUEL CELLS, ALONG WITH CONVENTIONAL POWER GENERATION MACHINERY.

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        *********         COAL CONSUMPTION IS   4991. TONS PER DAY.           *********
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        *********                                                             *********
        *********         NET DELIVERED POWER IS   745. MEGAWATTS.            *********
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-------------------------------------------------DETAILS-----------------------------------------

COAL CONSUMPTION IS  416. THOUSANDS POUNDS PER HOUR, OR ABOUT  5314. MILLIONS BTU.
THE OXYGEN PLANT SIZE IS  2439. TONS PER DAY.
PROCESS STEAM CONSUMPTION IS   611. THOUSANDS POUNDS PER HOUR.

PROCESS MACHINERY (COMPRESSORS AND MAJOR PUMPS) NEED                   -259049. KILOWATTS
PROCESS GAS TURBO-EXPANDERS & CLOSELY-ASSOCIATED STEAM TURBINES GIVE    325352. KILOWATTS
DEBIT POWER VALUE OF IMPORTED STEAM,  0.1692E+03 THOUSAND LBS/HR        -19199. KILOWATTS
THE FUEL CELL ITSELF DELIVERS                                           698479. KILOWATTS
                  (AFTER CONVERSION TO 13.8KV-AC @ 93% EFF.)
THE NET POWER DELIVERED TO BUS BAR IS                                   745354. KILOWATTS

THE VOLTAGE OF THE FUEL CELL IS            0.???? VOLTS.
THE AREA OF THE FUEL CELL IS          6215517. SQ. FT.

OVERALL PLANT EFFICIENCY, COAL PILE TO BUSBAR, IS 47.9 PERCENT.
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                                                                              CASE    JPL.UG.4A
JPL.UG.4A FLUID BED, REFORM, 85% CONV, QUENCH, PRE-SHIFT, COSX, 20.0% CH4 RAW       POWER LIST
                                                                               5 JAN 83  9:56:52

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TABULATION OF SHAFT ENERGY AVAILABLE OR NEEDED.

  97   -31753.4 KILOWATTS      AIR COMPRESSOR FOR SUPPLY TO CRYOGENIC OXYGEN PLANT
  99   -13190.0 KILOWATTS      CRYOGENIC OXYGEN COMPRESSOR FOR SUPPLY TO COAL GASIFIER
   2    -5240.7 KILOWATTS      2 U-GAS GASIFIER NORMAL OPERATION. STREAM #2 = NET MAKE.
  20    30343.2 KILOWATTS      20 EXPAND SHIFT (I) EXIT GAS THROUGH POWER RECOVERY TURBINE.
  29    -6474.5 KILOWATTS      29 REMOVE SULFUR FROM RAW GAS BY SELEXOL SYSTEM.
  42    12331.1 KILOWATTS      42 ADD STEAM VIA TURBINE; PREVENT CARBON DROP IN FUEL CELL.
  44 -170742.3 KILOWATTS      44 COMPRESS COMBUSTION AIR TO FUEL CELL PRESSURE.
  51   712774.2 KILOWATTS      51 THE FUEL CELL, MAXIMUM CONVERSION, REFORM WITHIN ANODE.
  57    -5783.4 KILOWATTS      57 ANODE EXIT GAS BOOSTER FAN, TO FEED CATALYTIC BURNER.
  63   -23167.7 KILOWATTS      63 CATHODE RECYCLE GAS CIRCULATION FAN.
  64   231971.8 KILOWATTS      64 EXPAND STACK GAS THROUGH TURBINE TO RECOVER POWER.
  70    -1699.1 KILOWATTS      70 COMPRESS RECYCLE GAS TO INJECT FEED COAL TO GASIFIER.
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                                                                                           CASE     JPL.UG.4A
JPL.UG.4A FLUID BED, REFORM, 15% CONV, QUENCH, PRE-SHIFT, COSY, 20.0% CH4 RAW                STEAM LIST
                                                                                           5 JAN 83  8:54:52

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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           11242. MOLS STEAM 29 REMOVE SULFUR FROM RAW GAS BY SELEXOL SYSTEM.
42      1500.0 PSIA            9645. MOLS STEAM 42 ADD STEAM VIA TURBINE; PREVENT CARBON DROP IN FUEL CELL.
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                                                                                                  CASE    JPL.UG.4A
JPL.UG.4A FLUID BED, REFORM, REV CONV, QUENCH, PRE-SHIFT, COSX, 20.0% CH4 RAW                      WASTE HEAT OUT
                                                                                                 5 JAN 83  3:56:52

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TABULATION OF WASTE HEAT AVAILABLE FROM PROCESS BOILERS.

   9     700.0 DEGREES F.     247.42 MILLIONS BTU     9 COOL RAW GAS TO 700 DEGREES FAHRENHEIT, MAKE STEAM.
  21     600.0 DEGREES F.      -3.74 MILLIONS BTU    21 COOL INTER-STAGE SHIFT GAS IN WHB; RAISE STEAM.
  54     925.0 DEGREES F.     350.15 MILLIONS BTU    54 COOL ANODE EXIT GAS IN WHB - RAISE STEAM.
  60    1115.0 DEGREES F.  [illegible] MILLIONS BTU    60 COOL CATHODE RECYCLE GAS IN WHB - RAISE STEAM.
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CASES IVb, c, and d

Vary Fuel Cell Conversion

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CONV70    FLUID BED, REFORM, 85% CONV, QUENCH, PRE-SHIFT, COS%, 13.5% CH4 RAW          CASE    CONV70
                                                                                          SUMMARY
                                                                                     5 JAN 83 11:34:14
          METH-RICH U-GAS WITH ANODE METHANE REFORMING, 13.5% C TO C

  THIS IS THE FINAL SYNOPSIS OF THE ENERGY BUDGET OF AN INTEGRATED POWER PLANT USING COAL
GASIFICATION AND ELECTROCHEMICAL FUEL CELLS, ALONG WITH CONVENTIONAL POWER GENERATION MACHINERY.

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        *********           COAL CONSUMPTION IS   4991. TONS PER DAY.            *********
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        *********           NET DELIVERED POWER IS  651. MEGAWATTS.              *********
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------------------------------------------------DETAILS--------------------------------------------
COAL CONSUMPTION IS  416. THOUSANDS POUNDS PER HOUR, OR ABOUT  5314. MILLIONS BTU.
THE OXYGEN PLANT SIZE IS  3023. TONS PER DAY.
PROCESS STEAM CONSUMPTION IS   792. THOUSANDS POUNDS PER HOUR.

PROCESS MACHINERY (COMPRESSORS AND MAJOR PUMPS) NEED                     -270882. KILOWATTS
PROCESS GAS TURBO-EXPANDERS & CLOSELY-ASSOCIATED STEAM TURBINES GIVE      345973. KILOWATTS
CONDENSING STEAM TURBINE (MULTI-STAGE W. REHEAT) GIVES                     33398. KILOWATTS
THE FUEL CELL ITSELF DELIVERS                                             542658. KILOWATTS
                  (AFTER CONVERSION TO 13.6KV-AC @ 98% EFF.)
THE NET POWER DELIVERED TO BUSBAR IS                                      651147. KILOWATTS

THE VOLTAGE OF THE FUEL CELL IS          0.8400 VOLTS.
THE AREA OF THE FUEL CELL IS           4730581. SQ. FT.

OVERALL PLANT EFFICIENCY, COAL PILE TO BUSBAR, IS 41.8 PERCENT.
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CONV70    FLUID BED, REFORM, 85% CONV, QUENCH, PRE-SHIFT, COSX, 13.5% CH4 RAW                 CASE    CONV70
                                                                                              WASTE HEAT OUT
                                                                                             5 JAN 83 11:34:14

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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, 85X CONV, QUENCH, PRE-SHIFT, COSX, 13.5X CH4 RAW                    CASE    CONV70
                                                                                                  STEAM LIST
                                                                                               5 JAN 83 11:34:14
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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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CONV70     FLUID BED, REFORM, 85% CONV, QUENCH, PRE-SHIFT, COSX, 13.5% CH4 RAW            CASE    CONV70
                                                                                          POWER LIST
                                                                                       5 JAN 83 11:34:14
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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  553732.9 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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                                                                                            SE      ONV
CONV80   FLUID BED, REFORM, 85% CONV, QUENCH, PRE-SHIFT, COSX, 13.5% CH4 RAW                  SUMMARY
                                                                                            5 JAN 83 11:40:08
            METH-RICH U-GAS WITH ANODE METHANE REFORMING, 13.5% C TO C

 THIS IS THE FINAL SYNOPSIS CF THE ENERGY BUDGET OF AN INTEGRATED POWER PLANT USING COAL
GASIFICATION AND ELECTROCHEMICAL FUEL CELLS, ALONG WITH CONVENTIONAL POWER GENERATION MACHINERY.

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        *********        COAL CONSUMPTION IS   4991. TONS PER DAY.       *********
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        *********                                                        *********
        *********        NET DELIVERED POWER IS  688. MEGAWATTS.         *********
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                                            DETAILS
COAL CONSUMPTION IS  416. THOUSANDS POUNDS PER HOUR, OR ABOUT  5314. MILLIONS BTU.
THE OXYGEN PLANT SIZE IS  3023. TONS PER DAY.
PROCESS STEAM CONSUMPTION IS   792. THOUSANDS POUNDS PER HOUR.

PROCESS MACHINERY (COMPRESSORS AND MAJOR PUMPS) NEED                    -269868. KILOWATTS
PROCESS GAS TURBO-EXPANDERS & CLOSELY-ASSOCIATED STEAM TURBINES GIVE     330047. KILOWATTS
CONDENSING STEAM TURBINE (MULTI-STAGE W. REHEAT) GIVES                    12576. KILOWATTS
THE FUEL CELL ITSELF DELIVERS                                            615459. KILOWATTS
                  (AFTER CONVERSION TO 13.6KV-AC @ 98% EFF.)
THE NET POWER DELIVERED TO BUSBAR IS                                     688213. KILOWATTS

THE VOLTAGE OF THE FUEL CELL IS           0.8336 VOLTS.
THE AREA OF THE FUEL CELL IS         5406378. SQ. FT.

OVERALL PLANT EFFICIENCY, COAL PILE TO BUSBAR, IS 44.2 PERCENT.
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                                                                                                   CASE    CONV80
  CONV80     FLUID BED, REFORM, 85% CONV, QUENCH, PRE-SHIFT, COS%, 13.5% CH4 RAW                      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           11839. 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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                                                                                                   CASE    CONVB0
  CONVB0    FLUID BED, REFORM, 85% CONV, QUENCH, PRE-SHIFT, COSX, 13.5% CH4 RAW                    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.
```

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                                                                                                 CASE  _ CONV80
CONV80    FLUID BED, REFORM, 85% CONV, QUENCH, PRE-SHIFT, COS%, 13.5% CH4 RAW                        POWER LIST
                                                                                                 5 JAN 83 11:40:08

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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  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                CASE    CONV90
                                                                                                   SUMMARY
                                                                                              5 JAN 83 11:44:57
             METH-RICH U-GAS WITH ANODE METHANE REFORMING, 13.5% C TO C

   THIS IS THE FINAL SYNOPSIS OF THE ENERGY BUDGET OF AN INTEGRATED POWER PLANT USING COAL
 GASIFICATION AND ELECTROCHEMICAL FUEL CELLS, ALONG WITH CONVENTIONAL POWER GENERATION MACHINERY.

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         *********         COAL CONSUMPTION IS   4991. TONS PER DAY.         *********
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         *********         NET DELIVERED POWER IS  724. MEGAWATTS.           *********
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 ---------------------------------------------DETAILS------------------------------------------

 COAL CONSUMPTION IS  416. THOUSANDS POUNDS PER HOUR, OR ABOUT  5314. MILLIONS BTU.
 THE OXYGEN PLANT SIZE IS  3023. TONS PER DAY.
 PROCESS STEAM CONSUMPTION IS   792. THOUSANDS POUNDS PER HOUR.

 PROCESS MACHINERY (COMPRESSORS AND MAJOR PUMPS) NEED                      -268951. KILOWATTS
 PROCESS GAS TURBO-EXPANDERS & CLOSELY-ASSOCIATED STEAM TURBINES GIVE       315413. KILOWATTS
 DEBIT POWER VALUE OF IMPORTED STEAM,  0.5872E+02 THOUSAND LBS/HR            -7295. KILOWATTS
 THE FUEL CELL ITSELF DELIVERS                                              684413. KILOWATTS
                   (AFTER CONVERSION TO 13.6KV-AC @ 98% EFF.)
 THE NET POWER DELIVERED TO BUSBAR IS                                       723580. KILOWATTS

 THE VOLTAGE OF THE FUEL CELL IS          0.8240 VOLTS.
 THE AREA OF THE FUEL CELL IS          6092175. SQ. FT.

 OVERALL PLANT EFFICIENCY, COAL PILE TO BUSBAR, IS 46.5 PERCENT.
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CONV90     FLUID BED, REFORM, 85% CONV, QUENCH, PRE-SHIFT, COS%, 13.5% CH4 RAW                    CASE     CONV90
                                                                                                      POWER LIST
                                                                                                 5 JAN 83 11:44:57

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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   698380.5 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    -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, COS%, 13.5% 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.
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CONV90    FLUID BED, REFORM, 85% CONV, QUENCH, PRE-SHIFT, COSX, 13.5% CH4 RAW                    CASE    CONV90
                                                                                                  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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               METH-RICH U-GAS WITH ANODE METHANE REFORMING. 13.5% C TO C

   THIS IS THE FINAL SYNOPSIS OF THE ENERGY BUDGET OF AN INTEGRATED POWER PLANT USING COAL
GASIFICATION AND ELECTROCHEMICAL FUEL CELLS. ALONG WITH CONVENTIONAL POWER GENERATION MACHINERY.

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          *********          COAL CONSUMPTION IS   4991. TONS PER DAY.           *********
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          *********          NET DELIVERED POWER IS  712. MEGAWATTS.             *********
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-----------------------------------------------------DETAILS----------------------------------------------
COAL CONSUMPTION IS  416. THOUSANDS POUNDS PER HOUR. OR ABOUT  5314. MILLIONS BTU.
THE OXYGEN PLANT SIZE IS  3023. TONS PER DAY.
PROCESS STEAM CONSUMPTION IS   792. THOUSANDS POUNDS PER HOUR.

PROCESS MACHINERY (COMPRESSORS AND MAJOR PUMPS) NEED                    -268075. KILOWATTS
PROCESS GAS TURBO-EXPANDERS & CLOSELY-ASSOCIATED STEAM TURBINES GIVE     327910. KILOWATTS
CONDENSING STEAM TURBINE (MULTI-STAGE W. REHEAT) GIVES                     1290. KILOWATTS
THE FUEL CELL ITSELF DELIVERS                                            651336. KILOWATTS
                     (AFTER CONVERSION TO 13.6KV-AC @ 98% EFF.)
THE NET POWER DELIVERED TO BUSBAR IS                                     712461. KILOWATTS

THE VOLTAGE OF THE FUEL CELL IS          0.8303 VOLTS.
THE AREA OF THE FUEL CELL IS          5744276. SQ. FT.

OVERALL PLANT EFFICIENCY. COAL PILE TO BUSBAR. IS 45.8 PERCENT.
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AIR4.160  JPL4, 2-STAGE AIR COMPRESSOR, WITH AIR FEED @ 160.0% OF STOICHIOMETRIC          CASE    AIR4.160
                                                                                             POWER LIST
                                                                                          14 JAN 83  9:21:14

*****************************************************************************************************************

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

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                                                                                   CASE    JPL.25.4
 JPL.25.4  FLUID BED, REFORM, 85% CONV, QUENCH, PRE-SHIFT, COSX, 13.5% CH4 RAW           SUMMARY
                                                                                  28 JAN 83 14143100
               METH-RICH U-GAS WITH ANODE METHANE REFORMING, 13.5% C TO C

   THIS IS THE FINAL SYNOPSIS OF THE ENERGY BUDGET OF AN INTEGRATED POWER PLANT USING COAL
 GASIFICATION AND ELECTROCHEMICAL FUEL CELLS, ALONG WITH CONVENTIONAL POWER GENERATION MACHINERY.

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         *********        COAL CONSUMPTION IS   4991. TONS PER DAY.       *********
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         *********        NET DELIVERED POWER IS  720. MEGAWATTS.         *********
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                                        DETAILS

 COAL CONSUMPTION IS  416. THOUSANDS POUNDS PER HOUR, OR ABOUT  5314. MILLIONS BTU.
 THE OXYGEN PLANT SIZE IS  3023. TONS PER DAY.
 PROCESS STEAM CONSUMPTION IS   767. THOUSANDS POUNDS PER HOUR.

 PROCESS MACHINERY (COMPRESSORS AND MAJOR PUMPS) NEED                       -269122. KILOWATTS
 PROCESS GAS TURBO-EXPANDERS & CLOSELY-ASSOCIATED STEAM TURBINES GIVE        318475. KILOWATTS
 CONDENSING STEAM TURBINE (MULTI-STAGE W. REHEAT) GIVES                       12411. KILOWATTS
 THE FUEL CELL ITSELF DELIVERS                                               657775. KILOWATTS
                 (AFTER CONVERSION TO 13.6KV-AC @ 98% EFF.)
 THE NET POWER DELIVERED TO BUSBAR IS                                        719539. KILOWATTS

 THE VOLTAGE OF THE FUEL CELL IS         0.8385 VOLTS.
 THE AREA OF THE FUEL CELL IS         5744273. SQ. FT.

 OVERALL PLANT EFFICIENCY, COAL PILE TO BUSBAR, IS 46.2 PERCENT.
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                                                                                CASE     JPL.25.4
JPL.25.4  FLUID BED, REFORM, 85% CONV, QUENCH, PRE-SHIFT, COSX, 13.5% CH4 RAW      POWER LIST
                                                                                28 JAN 83 14143100

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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    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 14:43:00
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TABULATION OF STEAM NEEDED IN PROCESS.

  2       550.0 PSIA             24072. MOLS STEAM   2 U-GAS GASIFIER NORMAL OPERATION. STREAM #2 = NET MAKE.
 29        65.0 PSIA             11804. MOLS STEAM  29 REMOVE SULFUR FROM RAW GAS BY SELEXOL SYSTEM.
 42      1500.0 PSIA              6692. MOLS STEAM  42 ADD STEAM VIA TURBINE; PREVENT CARBON DROP IN FUEL CELL.
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JPL.25.4  FLUID BED, REFORM, 85% CONV, QUENCH, PRE-SHIFT, COSX, 13.5% CH4 RAW                        CASE     JPL.25.4
                                                                                                        WASTE HEAT OUT
                                                                                                     28 JAN 83 14:43:00
***********************************************************************************************************************

TABULATION OF WASTE HEAT AVAILABLE FROM PROCESS BOILERS.

   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 WHB: 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

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                                                                                               CASE    DKF.4
DKF.4     COMPETITIVE, REFORM, 85% CONV, NO QUENCH, PRE-SHIFT, COSX, 13.5% CH4                  SUMMARY
                                                                                          12 JAN 83 15:18:23
              METH-RICH U-GAS WITH ANODE METHANE REFORMING, 13.5% C TO C

  THIS IS THE FINAL SYNOPSIS OF THE ENERGY BUDGET OF AN INTEGRATED POWER PLANT USING COAL
GASIFICATION AND ELECTROCHEMICAL FUEL CELLS, ALONG WITH CONVENTIONAL POWER GENERATION MACHINERY.

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         *********         COAL CONSUMPTION IS   4991. TONS PER DAY.      *********
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         *********         NET DELIVERED POWER IS  729. MEGAWATTS.        *********
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---------------------------------------------DETAILS----------------------------------------

COAL CONSUMPTION IS  416. THOUSANDS POUNDS PER HOUR, OR ABOUT  5314. MILLIONS BTU.
THE OXYGEN PLANT SIZE IS  3023. TONS PER DAY.
PROCESS STEAM CONSUMPTION IS   745. THOUSANDS POUNDS PER HOUR.

PROCESS MACHINERY (COMPRESSORS AND MAJOR PUMPS) NEED                      -268949. KILOWATTS
PROCESS GAS TURBO-EXPANDERS & CLOSELY-ASSOCIATED STEAM TURBINES GIVE       314211. KILOWATTS
CONDENSING STEAM TURBINE (MULTI-STAGE W. REHEAT) GIVES                      27727. KILOWATTS
THE FUEL CELL ITSELF DELIVERS                                              655950. KILOWATTS
                 (AFTER CONVERSION TO 13.6KV-AC @ 98% EFF.)
THE NET POWER DELIVERED TO BUSBAR IS                                       728939. KILOWATTS

THE VOLTAGE OF THE FUEL CELL IS          0.8362 VOLTS.
THE AREA OF THE FUEL CELL IS          5744265. SQ. FT.

OVERALL PLANT EFFICIENCY, COAL PILE TO BUSBAR, IS 46.8 PERCENT.
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                                                                                                  CASE    DKF.4
DKF.4     COMPETITIVE, REFORM, 85% CONV, NO QUENCH, PRE-SHIFT, COSX, 13.5% CH4                      POWER LIST
                                                                                                  12 JAN 83 15:18:23

**********************************************************************************************************

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 (I) 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
DKF.4     COMPETITIVE, REFORM, 85% CONV, NO QUENCH, PRE-SHIFT, COSX, 13.5% CH4           STEAM LIST
                                                                                    12 JAN 83 15:18:23

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

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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.     422.40 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 WHB; RAISE STEAM.
   54     925.0 DEGREES F.     337.21 MILLIONS BTU    54 COOL ANODE EXIT GAS IN WHB - RAISE STEAM.
   60    1115.0 DEGREES F.     200.80 MILLIONS BTU    60 COOL CATHODE RECYCLE GAS IN WHB - RAISE STEAM.
```

CASE V

Dust-Tolerant Shift Reactor

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             DUSTY SHIFT, WITH PRE-QUENCH RAW GAS HEAT RECOVERY.

  THIS IS THE FINAL SYNOPSIS OF THE ENERGY BUDGET OF AN INTEGRATED POWER PLANT USING COAL
GASIFICATION AND ELECTROCHEMICAL FUEL CELLS, ALONG WITH CONVENTIONAL POWER GENERATION MACHINERY.

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        *********         COAL CONSUMPTION IS   4991. TONS PER DAY.       *********
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        *********         NET DELIVERED POWER IS  739. MEGAWATTS.         *********
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_____________________________________________DETAILS_______________________________________

COAL CONSUMPTION IS  416. THOUSANDS POUNDS PER HOUR, OR ABOUT  5314. MILLIONS BTU.
THE OXYGEN PLANT SIZE IS  3023. TONS PER DAY.
PROCESS STEAM CONSUMPTION IS   709. THOUSANDS POUNDS PER HOUR.

PROCESS MACHINERY (COMPRESSORS AND MAJOR PUMPS) NEED                    -278648. KILOWATTS
PROCESS GAS TURBO-EXPANDERS & CLOSELY-ASSOCIATED STEAM TURBINES GIVE     334380. KILOWATTS
CONDENSING STEAM TURBINE (MULTI-STAGE W. REHEAT) GIVES                    29546. KILOWATTS
THE FUEL CELL ITSELF DELIVERS                                            653268. KILOWATTS
                  (AFTER CONVERSION TO 13.6KV-AC @ 98% EFF.)
THE NET POWER DELIVERED TO BUSBAR IS                                     738545. KILOWATTS

THE VOLTAGE OF THE FUEL CELL IS         0.8328 VOLTS.
THE AREA OF THE FUEL CELL IS          5744278. SQ. FT.

OVERALL PLANT EFFICIENCY, COAL PILE TO BUSBAR, IS 47.4 PERCENT.
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RUN.81    DUSTY SHIFT BASE CASE. FC REFORM. 13.5% CARB EX COAL TO CH4                              CASE     RUN.81
                                                                                                      POWER LIST
                                                                                                    5 FEB 83 10:20:02

****************************************************************************************************

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.
  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   41792.8 KILOWATTS     41 EXPAND SWEET FUEL GAS THROUGH POWER RECOVERY TURBINE.
  46 -179574.5 KILOWATTS     46 COMPRESS COMBUSTION AIR TO FUEL CELL PRESSURE.
  54  666599.8 KILOWATTS     NORMAL FUEL CELL. 1200 F INPUT SETS VOLTAGE. 1260 F EXIT.
  61  -22106.3 KILOWATTS     61 CATHODE RECYCLE GAS CIRCULATION FAN.
  63  286304.0 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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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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                                                                                       CASE     RUN.81
RUN.81    DUSTY SHIFT BASE CASE, FC REFORM. 13.5% CARB EX COAL TO CH4                    WASTE HEAT OUT
                                                                                       5 FEB 83 10:20:02

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TABULATION OF WASTE HEAT AVAILABLE FROM PROCESS BOILERS.

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

High-Temperature Dust Removal

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                                                                                   CASE    QEX.2
QEX.2     HIGH TEMPERATURE DE-DUSTING, RAW GAS LIKE JPL4, QUENCH AFTER TURBINE.             SUMMARY
                                                                                  16 FEB 83 17:10:29
               HI-TEMP DE-DUST, LIKE JPL.4, ANODE REFORMING, 13.5% C TO C

   THIS IS THE FINAL SYNOPSIS OF THE ENERGY BUDGET OF AN INTEGRATED POWER PLANT USING COAL
 GASIFICATION AND ELECTROCHEMICAL FUEL CELLS, ALONG WITH CONVENTIONAL POWER GENERATION MACHINERY.

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         *********       COAL CONSUMPTION IS   4991. TONS PER DAY.      *********
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         *********       NET DELIVERED POWER IS  736. MEGAWATTS.        *********
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---------------------------------------------DETAILS----------------------------------------------

COAL CONSUMPTION IS  416. THOUSANDS POUNDS PER HOUR, OR ABOUT  5314. MILLIONS BTU.
THE OXYGEN PLANT SIZE IS  3023. TONS PER DAY.
PROCESS STEAM CONSUMPTION IS   792. THOUSANDS POUNDS PER HOUR.

PROCESS MACHINERY (COMPRESSORS AND MAJOR PUMPS) NEED                         -269584. KILOWATTS
PROCESS GAS TURBO-EXPANDERS & CLOSELY-ASSOCIATED STEAM TURBINES GIVE          347155. KILOWATTS
CONDENSING STEAM TURBINE (MULTI-STAGE W. REHEAT) GIVES                          3394. KILOWATTS
THE FUEL CELL ITSELF DELIVERS                                                 655041. KILOWATTS
                (AFTER CONVERSION TO 13.6KV-AC @ 98% EFF.)
THE NET POWER DELIVERED TO BUSBAR IS                                          735908. KILOWATTS

THE VOLTAGE OF THE FUEL CELL IS          0.8350 VOLTS.
THE AREA OF THE FUEL CELL IS           5744277. SQ. FT.

OVERALL PLANT EFFICIENCY, COAL PILE TO BUSBAR, IS 47.3 PERCENT.
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                                                                                         CASE    QEX.2
QEX.2     HIGH TEMPERATURE DE-DUSTING, RAW GAS LIKE JPL4, QUENCH AFTER TURBINE.            POWER LIST
                                                                                       16 FEB 83 17:10:29

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TABULATION OF SHAFT ENERGY AVAILABLE OR NEEDED.

 102  -39544.7 KILOWATTS      AIR COMPRESSSOR FOR SUPPLY TO CRYOGENIC OXYGEN PLANT
 104  -14426.3 KILOWATTS      CRYOGENIC OXYGEN COMPRESSSOR FOR SUPPLY TO COAL GASIFIER
   2   -5240.3 KILOWATTS      2 U-GAS GASIFIER NORMAL OPERATION. STREAM #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   10626.1 KILOWATTS      34 ADD STEAM VIA TURBINE; PREVENT CARBON DROP IN FUEL CELL.
  36 -171093.5 KILOWATTS      36 COMPRESS COMBUSTION AIR TO FUEL CELL PRESSURE.
  43  668409.7 KILOWATTS      43 THE FUEL CELL, MAXIMUM CONVERSION, REFORM WITHIN ANODE.
  48   -9325.0 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     QEX.2
QEX.2     HIGH TEMPERATURE DE-DUSTING, RAW GAS LIKE JPL4, QUENCH AFTER TURBINE.                          STEAM LIST
                                                                                                   16 FEB 83 17:10:29

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TABULATION OF STEAM NEEDED IN PROCESS.

 2        550.0 PSIA          24072. MOLS STEAM  2 U-GAS GASIFIER NORMAL OPERATION. STREAM #2 = NET MAKE.
21         65.0 PSIA          11997. MOLS STEAM  21 REMOVE SULFUR FROM RAW GAS BY SELEXOL SYSTEM.
34       1500.0 PSIA           7994. MOLS STEAM  34 ADD STEAM VIA TURBINE; PREVENT CARBON DROP IN FUEL CELL.
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QEX.2      HIGH TEMPERATURE DE-DUSTING, RAW GAS LIKE JPL4, QUENCH AFTER TURBINE.                        CASE    QEX.2
                                                                                                          WASTE HEAT OUT
                                                                                                        16 FEB 83 17:10:29

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TABULATION OF WASTE HEAT AVAILABLE FROM PROCESS BOILERS.

  11      600.0 DEGREES F.     332.94 MILLIONS BTU    11 COOL SHIFT FEED GAS TO 600F, RAISE STEAM.
  13      600.0 DEGREES F.     119.32 MILLIONS BTU    13 COOL INTER-STAGE SHIFT GAS IN WHB; RAISE STEAM.
  46      925.0 DEGREES F.     334.93 MILLIONS BTU    46 COOL ANODE EXIT GAS IN WHB - RAISE STEAM.
  51     1115.0 DEGREES F.     137.43 MILLIONS BTU    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
L.THICEL  SULFUR-TOLERANT CELL, 55% CONV, INT. REFORM, NO PRECOOL, 13.5% CH4 RA                   SUMMARY
                                                                                              30 MAR 83  8:52:03
             UTILITIES TABULATION FOR SOUR GAS FUEL CELL.

  THIS IS THE FINAL SYNOPSIS OF THE ENERGY BUDGET OF AN INTEGRATED POWER PLANT USING COAL
GASIFICATION AND ELECTROCHEMICAL FUEL CELLS, ALONG WITH CONVENTIONAL POWER GENERATION MACHINERY.

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        *********         COAL CONSUMPTION IS   4991. TONS PER DAY.        *********
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        *********         NET DELIVERED POWER IS  767. MEGAWATTS.          *********
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---------------------------------------------DETAILS---------------------------------------
COAL CONSUMPTION IS  416. THOUSANDS POUNDS PER HOUR, OR ABOUT  5314. MILLIONS BTU.
THE OXYGEN PLANT SIZE IS  3023. TONS PER DAY.
PROCESS STEAM CONSUMPTION IS   434. THOUSANDS POUNDS PER HOUR.

PROCESS MACHINERY (COMPRESSORS AND MAJOR PUMPS) NEED                    -300204. KILOWATTS
PROCESS GAS TURBO-EXPANDERS & CLOSELY-ASSOCIATED STEAM TURBINES GIVE     353843. KILOWATTS
CONDENSING STEAM TURBINE (MULTI-STAGE W. REHEAT) GIVES                    74353. KILOWATTS
THE FUEL CELL ITSELF DELIVERS                                            639344. KILOWATTS
                  (AFTER CONVERSION TO 13.6KV-AC @ 98% EFF.)
THE NET POWER DELIVERED TO BUSBAR IS                                     767336. KILOWATTS

THE VOLTAGE OF THE FUEL CELL IS         0.8148 VOLTS.
THE AREA OF THE FUEL CELL IS          5745691. SQ. FT.

OVERALL PLANT EFFICIENCY, COAL PILE TO BUSBAR, IS 49.3 PERCENT.
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                                                                                                   CASE    L.THICEL
L.THICEL  SULFUR-TOLERANT CELL, 95% CONV, INT. REFORM, NO PRECOOL, 13.5% CH4 RAW                      POWER LIST
                                                                                                  30 MAR 83  8:52:03

****************************************************************************************************

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.
   5    69393.7 KILOWATTS      5 EXPAND RAW GAS TO FUEL CELL PRESSURE; COLLECT KILOWATTS.
  18    -1909.7 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.
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L.THICEL  SULFUR-TOLERANT CELL, 85% CONV, 89%, REFORM, NO PRECOOL, 13.5% CH4 RAW          CASE    L.THICEL
                                                                                              STEAM LIST
                                                                                           30 MAR 83  8:52:03

*********************************************************************************************************************

TABULATION OF STEAM NEEDED IN PROCESS.

 2        550.0 PSIA          24072. MOLS STEAM  2 U-GAS GASIFIER NORMAL OPERATION. STREAM #2 = NET MAKE.
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L.THICEL  SULFUR-TOLERANT CELL, 85% CONV, INT. REFORM, NO PRECOOL, 13.5% CH4 RAW                                CASE     L.THICEL
                                                                                                                   WASTE HEAT OUT
                                                                                                                 30 MAR 83   8:52:03

*******************************************************************************************************************************

TABULATION OF WASTE HEAT AVAILABLE FROM PROCESS BOILERS.

    9     1200.0 DEGREES F.       3.04 MILLIONS BTU     9 COOL RAW GAS IN WHB; MAKE STEAM.
   32      900.0 DEGREES F.     370.34 MILLIONS BTU    32 COOL ANODE EXIT GAS IN WHB - RAISE STEAM.
   38     1010.0 DEGREES F.     867.24 MILLIONS BTU    38 COOL CATHODE EXIT RECYCLE GAS IN WHB; RAISE STEAM.
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CASE VIII

High-Temperature CO_2 Transfer

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                                                                                                   CASE      RUN.D2
   RUN.D2  HIGH TEMP. CO2 PASS ADDED TO BASE CASE, COMPARE   AIR.4E                                     SUMMARY
                                                                                                   17 FEB 83  7:18:13
              HIGH-TEMPERATURE CO2 TRANSFER ADDED TO BASE CASE.

   THIS IS THE FINAL SYNOPSIS OF THE ENERGY BUDGET OF AN INTEGRATED POWER PLANT USING COAL
 GASIFICATION AND ELECTROCHEMICAL FUEL CELLS, ALONG WITH CONVENTIONAL POWER GENERATION MACHINERY.

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        *********         COAL CONSUMPTION IS   4991. TONS PER DAY.      *********
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        *********         NET DELIVERED POWER IS  750. MEGAWATTS.        *********
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 --------------------------------------------DETAILS---------------------------------------------

 COAL CONSUMPTION IS  416. THOUSANDS POUNDS PER HOUR, OR ABOUT  5314. MILLIONS BTU.
 THE OXYGEN PLANT SIZE IS  3023. TONS PER DAY.
 PROCESS STEAM CONSUMPTION IS   748. THOUSANDS POUNDS PER HOUR.

 PROCESS MACHINERY (COMPRESSORS AND MAJOR PUMPS) NEED                       -264691. KILOWATTS
 PROCESS GAS TURBO-EXPANDERS & CLOSELY-ASSOCIATED STEAM TURBINES GIVE        271368. KILOWATTS
 CONDENSING STEAM TURBINE (MULTI-STAGE W. REHEAT) GIVES                       50365. KILOWATTS
 THE FUEL CELL ITSELF DELIVERS                                               692848. KILOWATTS
                 (AFTER CONVERSION TO 13.6KV-AC @ 98% EFF.)
 THE NET POWER DELIVERED TO BUSBAR IS                                        750391. KILOWATTS

 THE VOLTAGE OF THE FUEL CELL IS           0.8832 VOLTS.
 THE AREA OF THE FUEL CELL IS           5744291. SQ. FT.

 OVERALL PLANT EFFICIENCY, COAL PILE TO BUSBAR, IS 48.2 PERCENT.
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 RUN.D2  HIGH TEMP. CO2 PASS ADDED TO BASE CASE, COMPARE   AIR.4E                                    CASE       RUN.D2
                                                                                                         POWER LIST
                                                                                                     17 FEB 83  7:18:13

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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 (I) EXIT GAS THROUGH POWER RECOVERY TURBINE.
  49    -3144.6 KILOWATTS      49 REMOVE SULFUR FROM RAW GAS BY SELEXOL SYSTEM.
  63      913.2 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   -17925.6 KILOWATTS      76 CATHODE RECYCLE GAS CIRCULATION FAN.
  80    -7145.8 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      RUN.02
                                                                                        STEAM LIST
   RUN.02  HIGH TEMP. CO2 PASS ADDED TO BASE CASE, COMPARE   AIR.45                 17 FEB 83  7:19:13

 ***************************************************************************************************

TABULATION OF STEAM NEEDED IN PROCESS.

  2      550.0 PSIA            24072. MOLS STEAM  2 U-GAS GASIFIER NORMAL OPERATION. STREAM #2 = NET MAKE.
 36     1200.0 PSIA             5195. MOLS STEAM  36 HIGH-TEMPERATURE CO2 REMOVAL FROM SHIFT I EXIT GAS.
 37     1200.0 PSIA             1559. MOLS STEAM  37 REGEN OF MGO 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              583. MOLS STEAM  63 ADD STEAM VIA TURBINE; PREVENT CARBON DROP IN FUEL CELL.
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RUN.D2  HIGH TEMP. CO2 PASS ADDED TO BASE CASE, COMPARE   AIR.4E                                              CASE       RUN.D2
                                                                                                                WASTE HEAT OUT
                                                                                                              17 FEB 83  7:13:13

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TABULATION OF WASTE HEAT AVAILABLE FROM PROCESS BOILERS.

   8     1420.0 DEGREES F.     253.07 MILLIONS BTU     8 COOL HOT SOUR RAW GAS IN WHB; RAISE STEAM. (ONCE TO #39)
  15      690.0 DEGREES F.     237.34 MILLIONS BTU    15 COOL QUENCHED RAW GAS IN WHB; MAKE STEAM.
  20      370.0 DEGREES F.      36.97 MILLIONS BTU    20 COOL RAW GAS; PRE-HEAT BOILER FEED WATER.
  35      750.0 DEGREES F.      42.64 MILLIONS BTU    35 COOL SHIFT I EXIT GAS IN WHB; RAISE STEAM.
  36      853.9 DEGREES F.     273.24 MILLIONS BTU    36 HIGH-TEMPERATURE CO2 REMOVAL FROM SHIFT I EXIT GAS.
  41      600.0 DEGREES F.      20.29 MILLIONS BTU    41 COOL INTER-STAGE SHIFT GAS IN WHB; RAISE STEAM.
  75      940.0 DEGREES F.     239.60 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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RUN.C3  HIGH TEMP. H2S & DUST REMOVAL, MAKE MOLTEN SULFUR FROM OXYGEN.               CASE      RUN.C3
                                                                                        SUMMARY
           HIGH-TEMPERATURE H2S TRANSFER ADDED TO BASE CASE.                         28 FEB 83 14:11:14

  THIS IS THE FINAL SYNOPSIS OF THE ENERGY BUDGET OF AN INTEGRATED POWER PLANT USING COAL
GASIFICATION AND ELECTROCHEMICAL FUEL CELLS, ALONG WITH CONVENTIONAL POWER GENERATION MACHINERY.

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        *********        COAL CONSUMPTION IS   4991. TONS PER DAY.    *********
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        *********        NET DELIVERED POWER IS  779. MEGAWATTS.      *********
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----------------------------------------DETAILS___________________________________________
COAL CONSUMPTION IS  416. THOUSANDS POUNDS PER HOUR, OR ABOUT  5314. MILLIONS BTU.
THE OXYGEN PLANT SIZE IS  3023. TONS PER DAY.
PROCESS STEAM CONSUMPTION IS   438. THOUSANDS POUNDS PER HOUR.

PROCESS MACHINERY (COMPRESSORS AND MAJOR PUMPS) NEED                       -270322. KILOWATTS
PROCESS GAS TURBO-EXPANDERS & CLOSELY-ASSOCIATED STEAM TURBINES GIVE        331935. KILOWATTS
CONDENSING STEAM TURBINE (MULTI-STAGE W. REHEAT) GIVES                       78862. KILOWATTS
THE FUEL CELL ITSELF DELIVERS                                               638305. KILOWATTS
                (AFTER CONVERSION TO 13.6KV-AC @ 98% EFF.)
THE NET POWER DELIVERED TO BUSBAR IS                                        779180. KILOWATTS

THE VOLTAGE OF THE FUEL CELL IS           0.9141 VOLTS.
THE AREA OF THE FUEL CELL IS            5745691. SQ. FT.

OVERALL PLANT EFFICIENCY, COAL PILE TO BUSBAR, IS 50.0 PERCENT.
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RUN.C3  HIGH TEMP. H2S & DUST REMOVAL, MAKE MOLTEN SULFUR FROM OXYGEN.                              CASE      RUN.C3
                                                                                                      POWER LIST
                                                                                                 28 FEB 83 14:11:14

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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 COMPRESOR 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    -1625.7 KILOWATTS      55 COMPRESS COAL FEED INJECTION GAS, A SMALL RECYCLE.
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C-4

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  RUN.C3  HIGH TEMP. H2S & DUST REMOVAL, MAKE MOLTEN SULFUR FROM OXYGEN.                     CASE      RUN.C3
                                                                                                 STEAM LIST
                                                                                             28 FEB 83 14:11:14
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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             58. MOLS STEAM  15 RECLAIM ZN SPINEL (OXIDE) BY OXYGEN; MAKE SULFUR.
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                                                                                        CASE       RUN.C3
RUN.C3  HIGH TEMP. H2S & DUST REMOVAL, MAKE MOLTEN SULFUR FROM OXYGEN.                    WASTE HEAT OUT
                                                                                        28 FEB 83 14:11:14

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TABULATION OF WASTE HEAT AVAILABLE FROM PROCESS BOILERS.

   8     1500.0 DEGREES F.     209.87 MILLIONS BTU    8 COOL HOT SOUR RAW GAS IN WHB; 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   19 COOL 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

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                                                                                                      CASE    JPL.UG.7
JPL.UG.7  FLUID BED, REFORM, 85% CONV, QUENCH & SHIFT, ANODE RECYCLE, 13.5% CH4                         SUMMARY
                                                                                                     27 JAN 83  9:19:37
             U-GAS, ANODE RECYCLE, ANODE REFORMING, 13.5% C TO CH4.

   THIS IS THE FINAL SYNOPSIS OF THE ENERGY BUDGET OF AN INTEGRATED POWER PLANT USING COAL
 GASIFICATION AND ELECTROCHEMICAL FUEL CELLS, ALONG WITH CONVENTIONAL POWER GENERATION MACHINERY.

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         *********                                                            *********
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         *********                                                            *********
         *********         COAL CONSUMPTION IS    4991. TONS PER DAY.         *********
         *********                                                            *********
         *********                                                            *********
         *********                                                            *********
         *********         NET DELIVBRED POWER IS  675. MEGAWATTS.            *********
         *********                                                            *********
         *********                                                            *********
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                                     DETAILS

 COAL CONSUMPTION IS   416. THOUSANDS POUNDS PER HOUR, OR ABOUT  5314. MILLIONS BTU.
 THE OXYGEN PLANT SIZE IS  3023. TONS PER DAY.
 PROCESS STEAM CONSUMPTION IS  1740. THOUSANDS POUNDS PER HOUR.

 PROCESS MACHINERY (COMPRESSORS AND MAJOR PUMPS) NEED                          -402806. KILOWATTS
 PROCESS GAS TURBO-EXPANDERS & CLOSELY-ASSOCIATED STEAM TURBINES GIVE           415662. KILOWATTS
 DEBIT POWER VALUE OF IMPORTED STEAM, 0.1154E+04 THOUSAND LBS/HR               -143356. KILOWATTS
 THE FUEL CELL ITSELF DELIVERS                                                  805045. KILOWATTS
                    (AFTER CONVERSION TO 13.6KV-AC @ 98% EFF.)
 THE NET POWER DELIVERED TO BUSBAR IS                                           674544. KILOWATTS

 THE VOLTAGE OF THE FUEL CELL IS          0.9751 VOLTS.
 THE AREA OF THE FUEL CELL IS        5667912. SQ. FT.

 OVERALL PLANT EFFICIENCY, COAL PILE TO BUSBAR, IS 43.3 PERCENT.
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                                                                                    CASE     JPL.UG.7
JPL.UG.7  FLUID BED, REFORM, 85% CONV, QUENCH & SHIFT, ANODE RECYCLE, 13.5% CH4        POWER LIST
                                                                                 27 JAN 83  9:19:37

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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.
   22    46127.9 KILOWATTS      22 EXPAND SHIFT (I) 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.
   40    37285.6 KILOWATTS      41 GET 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.
   45    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 FEED INJECTION GAS SIDESTREAM RECYCLE #37.
   78   -75671.4 KILOWATTS      78 COMPRESS CARBON DIOXIDE FROM SELEXOL II STEP #45.
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                                                                                          CASE     JPL.UG.7
JPL.UG.7  FLUID BED, REFORM, 85% CONV, QUENCH & SHIFT, ANODE RECYCLE, 13.5% CH4             STEAM LIST
                                                                                        27 JAN 83   9:19:37

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TABULATION OF STEAM NEEDED IN PROCESS.

 2       500.0 PSIA          24072. MOLS STEAM   2 U-GAS GASIFIER NORMAL OPERATION. STREAM #2 = NET MAKE.
40        65.0 PSIA          22142. MOLS STEAM  40 REMOVE SULFUR FROM RAW GAS BY STAGE I 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 II.
45      1500.0 PSIA              0. MOLS STEAM  46 EXPAND HP STEAM TO GET POWER & LP STEAM FOR STEP #45.
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                                                                                                   CASE    JPL.UG.7
JPL.UG.7  FLUID BED, REFORM, 85% CONV, QUENCH & SHIFT, ANODE RECYCLE, 13.5% CH4                     WASTE HEAT OUT
                                                                                                 27 JAN 83  9:19:37

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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 WHB; RAISE STEAM.
   28      600.0 DEGREES F.      69.63 MILLIONS BTU     28 COOL INTER-STAGE SHIFT (CYCLE) GAS IN WHB; 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 WHB; 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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AIR4.160  JPL4, 2-STAGE AIR COMPRESSOR, WITH AIR FEED @ 160.0% OF STOICHIOMETRIC      CASE    AIR4.160
                                                                                          SUMMARY
          METH-RICH U-GAS WITH ANODE METHANE REFORMING, 13.5% C TO C                   14 JAN 83  9:21:14

  THIS IS THE FINAL SYNOPSIS OF THE ENERGY BUDGET OF AN INTEGRATED POWER PLANT USING COAL
GASIFICATION AND ELECTROCHEMICAL FUEL CELLS, ALONG WITH CONVENTIONAL POWER GENERATION MACHINERY.

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        *********                                                          *********
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        *********                                                          *********
        *********                                                          *********
        *********         COAL CONSUMPTION IS   4991. TONS PER DAY.        *********
        *********                                                          *********
        *********                                                          *********
        *********                                                          *********
        *********         NET DELIVERED POWER IS  712. MEGAWATTS.          *********
        *********                                                          *********
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--------------------------------------------DETAILS--------------------------------------------
COAL CONSUMPTION IS  416. THOUSANDS POUNDS PER HOUR, OR ABOUT  5314. MILLIONS BTU.
THE OXYGEN PLANT SIZE IS  3023. TONS PER DAY.
PROCESS STEAM CONSUMPTION IS   792. THOUSANDS POUNDS PER HOUR.

PROCESS MACHINERY (COMPRESSORS AND MAJOR PUMPS) NEED                    -268075. KILOWATTS
PROCESS GAS TURBO-EXPANDERS & CLOSELY-ASSOCIATED STEAM TURBINES GIVE     327910. KILOWATTS
CONDENSING STEAM TURBINE (MULTI-STAGE W. REHEAT) GIVES                     1290. KILOWATTS
THE FUEL CELL ITSELF DELIVERS                                            651336. KILOWATTS
                  (AFTER CONVERSION TO 13.6KV-AC @ 98% EFF.)
THE NET POWER DELIVERED TO BUSBAR IS                                     712461. KILOWATTS

THE VOLTAGE OF THE FUEL CELL IS          0.8303 VOLTS.
THE AREA OF THE FUEL CELL IS           5744276. SQ. FT.

OVERALL PLANT EFFICIENCY, COAL PILE TO BUSBAR, IS 45.8 PERCENT.
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AIR4.160  JPL4, 2-STAGE AIR COMPRESSOR, WITH AIR FEED @ 160.0% OF STOICHIOMETRIC              CASE    AIR4.160
                                                                                                  POWER LIST
                                                                                               14 JAN 83   9:21:14

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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  -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 COAL 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

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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

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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.    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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                                      FUEL GAS CONVERSION AND CHARACTERISTICS
                                      WITH ADDED FEATURES OF:      FUEL CELL
                                                                   COAL GASIFIER
                                                                   UTILITY SUMMARY
                                      DECEMBER 1982, J W COLTON

INPUT DATA PLAY-BACK:_________________________________
AIR4.160  JPL4, 2-STAGE AIR COMPRESSOR, WITH AIR FEED @ 160.0% OF STOICHIOMETRIC                 14 JAN 83  9:21:14
 32.78  7.814  29.103  19.126  35.642  0.446  0.0  1.149  0.179  0.033

 18073
  2)  1          4               2 U-GAS GASIFIER NORMAL OPERATION. STREAM #2 = NET MAKE.
  1875.     500.     415.9        99.5    0.88
  3)  3         71               RECYCLE GAS TO INJECT COAL FEED. STREAM #3 = TOTAL EXIT GAS.
  1875.     500.
  4)  11         2               4 TEST CARBON LIMITS TO FIND NECESSARY QUENCH.
  1200.     490.
  5)  11         3               5 TEST CARBON LIMITS AT ANOTHER TEMPERATURE.
  1100.     490.
  6)  11         3               6 TEST CARBON LIMITS AT ANOTHER TEMPERATURE.
  1000.     490.
  7)  11         3               7 TEST CARBON LIMITS AT ANOTHER TEMPERATURE.
  900.      490.
  8)  306        6               8 QUENCH TO STABILIZE GAS & PREVENT CARBON DEPOSITION.
  1200.     495.
  9)  5          2               9 COOL RAW GAS TO 700 DEGREES FAHRENHEIT, MAKE STEAM.
  700.      490.
  10)  5         1               10 R10T18 COOL RAW GAS; WARM CLEAN GAS #18.
  640.      488.
  11)  5         1               11 R11T17 COOL RAW GAS; WARM CLEAN GAS #17.
  580.      486.
  12)  5         1               12 R12T16 COOL RAW GAS; WARM CLEAN GAS #16.
  520.      484.
  13)  5         1               13 R13T15 COOL RAW GAS; WARM CLEAN GAS #15.
  460.      482.
  14)  6         4               14 COOL MOIST RAW GAS; SCRUB DUST; REMOVE COND. AND NH3.
  375.      477.     0.0        -1.5
  15)  5       -13               15 R13T15 WARM SHIFT FEED GAS FROM RAW GAS #13.
  420.      475.
  16)  5       -12               16 R12T16 WARM SHIFT FEED GAS FROM RAW GAS #12.
  480.      473.
  17)  5       -11               17 R11T17 WARM SHIFT FEED GAS FROM RAW GAS #11.
  540.      471.
  18)  5       -10               18 R10T18 WARM SHIFT FEED GAS FROM RAW GAS #10.
  600.      469.
  19)  9         3               19 SHIFT CONVERSION, FIRST STAGE, HIGH TEMPERATURE & PRESS.
  870.      464.
  20)  21        1               20 EXPAND SHIFT (I) EXIT GAS THROUGH POWER RECOVERY TURBINE.
  710.      196.     92.0
  21)  5         2               21 COOL INTER-STAGE SHIFT GAS IN WHB; RAISE STEAM.
  600.      194.
  22)  9         3               22 SECOND STAGE SHIFT, MODERATE TEMP., LOWER PRES.
  695.      189.
  23)  5         1               23 R23T35 COOL SHIFTED GAS TO WARM SWEET FUEL #35.
  610.      187.
  24)  5         1               24 R24T34 COOL SHIFTED GAS TO WARM SWEET FUEL #34.
  520.      185.
  25)  5         1               25 R25T33 COOL SHIFTED GAS TO WARM SWEET FUEL #33.
  430.      183.
  26)  8         1               26 DESTROY CARBONYL SULFIDE BY HYDROLYSIS.
  430.      179.
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28)  6        1          28 COOL TO AMBIENT; DROP CONDENSATE; DUMP HEAT TO AIR & CW.
 100.    173.
29)  7        3          29 REMOVE SULFUR FROM RAW GAS BY SELEXOL SYSTEM.
 100.    168.    0.19
30)  4        1          SPLIT RECYCLE GAS #30 FOR COAL FEED; GAS #31 GOES FORWARD.
 100.    168.    0.075
32)  5      -27          32 R27T32 WARM SWEETENED GAS; COOL SHIFT EXIT #27.
 220.    166.
33)  5      -25          33 R25T33 WARM SWEETENED GAS; COOL SHIFT EXIT #25.
 340.    164.
34)  5      -24          34 R24T34 WARM SWEETENED GAS; COOL SHIFT EXIT #24.
 460.    162.
35)  5      -23          35 R23T35 WARM SWEETENED GAS; COOL SHIFT EXIT #23.
 580.    160.
36)  5      -55          36 R36F55 WARM SWEETENED GAS; COOL FUEL CELL ANODE EXIT #55.
 750.    158.
37)  7        4          37 REMOVE LAST TRACES OF SULFUR BY ZINC OXIDE METATHESIS.
 750.    153.
38)  11       2          38 TEST GAS FOR DANGER OF CARBON DEPOSITION.
 1200.   150.
39)  11       3          39 TEST CARBON DEPOSITION AT ANOTHER TEMPERATURE.
 1100.   150.
40)  11       3          40 TEST CARBON DEPOSITION AT ANOTHER TEMPERATURE.
 1000.   150.
41)  11       3          41 TEST CARBON DEPOSITION AT ANOTHER TEMPERATURE.
 900.    150.
42)  3723     3          42 ADD STEAM VIA TURBINE; PREVENT CARBON DROP IN FUEL CELL.
 1000.   1500.    92.0
43)  30       6          43 FEED AIR AT 160.0% OF STOICHIOMETRIC. (INCLUDING SCALEX).
 90.     14.3    365.61
44)  20       2          44 COMPRESS COMBUSTION AIR TO FUEL CELL PRESSURE.
 270.    164.
45)  5      -65          45 R45F65 - WARM COMBUSTION AIR FROM STACK GAS #65.
 480.    162.
46)  5        1          46 R46F62 - WARM COMBUSTION AIR FROM CATHODE RECYCLE #62.
 700.    160.
47)  5        1          47 R47F61 - WARM COMBUSTION AIR FROM CATHODE RECYCLE #61.
 910.    158.
48)  3       63          48 MERGE FRESH AIR #47 WITH RECYCLED CATHODE EXIT GAS #63.
 1000.   156.
49)  5713    48          49 BURN ANODE EXHAUST GAS #57 WITH DILUTED AIR #48.
 1200.   152.
50)  4203    49          50 JOIN THE TWO FUEL CELL FEEDS, ANODE #42, CATHODE #49.
 1150.   152.
51)  14       2          51 THE FUEL CELL, MAXIMUM CONVERSION, REFORM WITHIN ANODE.
 1200.   150.    .85     .150
52)  4        3          SEPARATE ANODE EXIT GAS, #52, FROM CATHODE EXIT GAS, #53.
 1200.   145.
54)  5205     2          54 COOL ANODE EXIT GAS IN WHB - RAISE STEAM.
 925.    143.
55)  5        1          55 R36F55 COOL ANODE EXIT GAS TO WARM ANODE FEED GAS #36.
 855.    141.
56)  26       1          56 THIS STEP DELETED.
 100.
57)  5519     1          57 ANODE EXIT GAS BOOSTER FAN, TO FEED CATALYTIC BURNER.
 885.    156.
58)  5304     1          SPLIT CATHODE EXIT. COOLED RECYCLE (#58), STACK EXIT (#59).
 1250.   142.    0.396
60)  5805     2          60 COOL CATHODE RECYCLE GAS IN WHB - RAISE STEAM.
 1115.   139.
61)  5      -47          61 R47F61 COOL CATHODE RECYCLE TO WARM COMBUSTION AIR #47.
 965.    137.
62)  5      -46          62 R46F62 COOL CATHODE RECYCLE TO WARM COMBUSTION AIR #46.
 880.    135.
63)  20       1          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        92.
65)  5         1           65 R45F65 COOL STACK GAS TO WARM COMBUSTION AIR #45.
 540.     15.0
66)  212       3           66 READ COMBUSTION FUEL VALUE OF RAW GAS FROM GASIFIER.
 100.
67)  3112      3           67 READ COMBUSTION FUEL VALUE OF SHIFTED SOUR GAS.
 100.
68)  3712      3           68 READ COMBUSTION FUEL VALUE OF SWEET ANODE FEED GAS.
 100.
69)  5412      3           69 READ 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.
 1875.
72)  12        3           72 READ COMBUSTION FUEL VALUE OF RECYCLE COAL INJECTION GAS.
 100.
73)  305       1           73 READ SENSIBLE HEATS OF REACTOR EXIT RAW GAS (INCL. #71)
 1507.8   500.
74)  25        3           METH-RICH U-GAS WITH ANODE METHANE REFORMING. 13.5% C TO CH4
 30.
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AIR4.160  JPL4, 2-STAGE AIR COMPRESSOR, WITH AIR FEED = 160.0% OF STOICHIOMETRIC          CASE    AIR4.160
                                                                                          STEP    97
                                                                                          14 JAN 83  9:21:14
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       AIR COMPRESSOR FOR SUPPLY TO CRYOGENIC OXYGEN PLANT

THIS STEP DESCRIBES A PROCESS GAS COMPRESSOR OF 2 STAGES.

                         STREAM NUMBER 97
             COMPONENT        MOLS      %

                CO2             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 (ACFM):  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     AIR4.160
AIR4.160  JPL4, 2-STAGE AIR COMPRESSOR, WITH AIR FEED @ 160.0% OF STOICHIOMETRIC   STEP     99
                                                                                  14 JAN 83  9:21:14

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          CRYOGENIC OXYGEN COMPRESSOR FOR SUPPLY TO COAL GASIFIER

THIS STEP DESCRIBES A PROCESS GAS COMPRESSOR OF 5 STAGES.

                        STREAM NUMBER 99
               COMPONENT        MOLS      %

                  CO2            0.0     0.0
                  H2O            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.      22035.
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

PRESSURE (PSIA):  SUCTION  DISCHARGE      VOLUME (ACFM):  SUCTION  DISCHARGE

    STAGE # 1        20.0      41.5                       36618.4   24099.1
    STAGE # 2        35.6      73.9                       20555.3   13527.8
    STAGE # 3        67.4     139.7                       10869.4    7153.3
    STAGE # 4       131.9     273.4                        5553.4    3654.8
    STAGE # 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.
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AIR4.160  JPL4, 2-STAGE AIR COMPRESSOR, WITH AIR FEED @ 160.0% OF STOICHIOMETRIC            CASE    AIR4.160
                                                                                            STEP    2
                                                                                            14 JAN 83  9:21:14

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        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:

                          STREAM NUMBER 99
             COMPONENT         MOLS      %

                 CO2            0.0     0.0
                 H2O            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.


THIS GASIFIER MAKES THE FOLLOWING RAW GAS (NET):

                          STREAM NUMBER  2
             COMPONENT         MOLS      %

                 H2         13633.2    25.9
                 CH4         3249.8     6.2
                 CO         12103.9    23.0
                 CO2         7954.5    15.1
                 H2O        14906.7    28.3
                 N2           185.5     0.4
                 C2H6           0.0     0.0
                 H2S          477.9     0.9
                 NH3           74.4     0.1
                 COS           13.7     0.0

                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
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AIR4.160  JPL4, 2-STAGE AIR COMPRESSOR, WITH AIR FEED @ 160.0% OF STOICHIOMETRIC           CASE    AIR4.160
                                                                                          STEP    3
                                                                                          14 JAN 83  9:21:14

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

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

                 FEED STREAM  2         FEED STREAM 71         EXIT STREAM  3
   COMPONENT       MOLS      %            MOLS      %            MOLS      %

      H2         13633.2   25.9         1872.2   51.9         15505.4   27.6
      CH4         3249.8    6.2          263.3    7.3          3513.2    6.3
      CO         12103.9   23.0          214.5    5.9         12318.5   21.9
      CO2         7954.5   15.1         1243.1   34.5          9197.6   16.4
      H2O        14906.7   28.3            0.0    0.0         14906.7   26.5
      N2           185.5    0.4           15.0    0.4           200.5    0.4
      C2H6           0.0    0.0            0.0    0.0             0.0    0.0
      H2S          477.9    0.9            0.0    0.0           477.9    0.9
      NH3           74.4    0.1            0.0    0.0            74.4    0.1
      COS           13.7    0.0            0.0    0.0            13.7    0.0

     TOTAL       52599.7  100.0         3608.3  100.0         56208.0  100.0

  TEMPERATURE        1875.0                1875.0                1875.0
     PRESSURE         500.0                  14.7                 500.0
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AIR4.160  JPL4, 2-STAGE AIR COMPRESSOR, WITH AIR FEED @ 160.0% OF STOICHIOMETRIC          CASE    AIR4.160
                                                                                          STEP    4
                                                                                          14 JAN 83  9:21:14

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        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

                                              GRAPHITE        AMORPHOUS
             GAS STREAM    GAS STREAM          LIMIT            LIMIT
                MOLS      MOL FRACTION     MOL FRACTION     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
     COS        13.7         0.0002

    TOTAL    56208.0         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.
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      8 QUENCH TO STABILIZE GAS & PREVENT CARBON DEPOSITION.

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

                       FEED STREAM  3         EXIT STREAM  8
         COMPONENT       MOLS      %             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
            COS           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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AIR4.160  JPL4, 2-STAGE AIR COMPRESSOR, WITH AIR FEED @ 160.0% OF STOICHIOMETRIC          CASE    AIR4.160
                                                                                          STEP    9
                                                                                          14 JAN 83  9121:14

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

THE DUTY OF THIS PROCESS COOLER IS:  -458.66 MILLIONS BTU PER HOUR.

                     STREAM NUMBER  9
          COMPONENT        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
             COS          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.
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AIR4.160  JPL4, 2-STAGE AIR COMPRESSOR, WITH AIR FEED @ 160.0% OF STOICHIOMETRIC           CASE    AIR4.160
                                                                                           STEP    10
                                                                                           14 JAN 83  9:21:14

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      10 R10T18 COOL RAW GAS; WARM CLEAN GAS #18.

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

                              STREAM NUMBER 10
            COMPONENT            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
               COS                13.7    0.0

              TOTAL            60830.6  100.0

PROCESS GAS TEMPERATURE,     FEED=    700.0       EXIT=   640.0  DEGREES FAHRENHEIT.
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

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        11 R11T17 COOL RAW GAS: WARM CLEAN GAS #17.

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

                      STREAM NUMBER 11
          COMPONENT        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
             COS          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
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        12 R12T16 COOL RAW GAS: WARM CLEAN GAS #16.

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

                         STREAM NUMBER 12
              COMPONENT      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
                 COS         13.7    0.0

                TOTAL     60830.6  100.0

PROCESS GAS TEMPERATURE,    FEED=   580.0      EXIT=   520.0  DEGREES FAHRENHEIT.
PROCESS GAS  ENTHALPY.      FEED=   248.20     EXIT=   217.14 MILLIONS BTU.
```

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AIR4.160  JPL4, 2-STAGE AIR COMPRESSOR, WITH AIR FEED @ 160.0% OF STOICHIOMETRIC          CASE    AIR4.160
                                                                                          STEP    13
                                                                                          14 JAN 83  9:21:14
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        13 R13T15 COOL RAW GAS; WARM CLEAN GAS #15.

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

                          STREAM NUMBER 13
             COMPONENT         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
                COS             13.7    0.0

               TOTAL         60830.6  100.0

PROCESS GAS TEMPERATURE,     FEED=   520.0        EXIT=   460.0  DEGREES FAHRENHEIT.
PROCESS GAS  ENTHALPY,       FEED=   217.14       EXIT=   186.44 MILLIONS BTU.
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AIR4.160  JPL4, 2-STAGE AIR COMPRESSOR, WITH AIR FEED @ 160.0% OF STOICHIOMETRIC          CASE    AIR4.160
                                                                                           STEP    14
                                                                                           14 JAN 83  9:21:14

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       14 COOL MOIST RAW GAS; SCRUB DUST; REMOVE COND. AND NH3.

            THIS STEP SUBTRACTS CONDENSATE FROM THE GAS AS DIRECTED BY USER.

                         FEED STREAM 13          EXIT STREAM 14
          COMPONENT       MOLS      %             MOLS      %

             H2         15505.4   25.5          15505.4   25.8
             CH4         3513.2    5.8           3513.2    5.8
             CO         12318.5   20.3          12318.5   20.5
             CO2         9197.6   15.1           9197.6   15.3
             H2O        19529.4   32.1          18905.5   31.4
             N2           200.5    0.3            200.5    0.3
             C2H6           0.0    0.0              0.0    0.0
             H2S          477.9    0.8            477.9    0.8
             NH3           74.4    0.1              0.0    0.0
             COS           13.7    0.0             13.7    0.0

            TOTAL       60830.6  100.0          60132.3  100.0

        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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AIR4.160  JPL4, 2-STAGE AIR COMPRESSOR, WITH AIR FEED @ 160.0% OF STOICHIOMETRIC          CASE    AIR4.160
                                                                                         STEP    15
                                                                                         14 JAN 83  9:21:14

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       15 R13T15 WARM SHIFT FEED GAS FROM RAW GAS #13.

THIS STEP RECUPERATES HEAT AGAINST STEP # 13
THE ENTHALPY TRANSFERRED IS:     30.70  MILLIONS BTU


                      STREAM NUMBER 15
          COMPONENT        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
             COS          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.
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AIR4.160  JPL4, 2-STAGE AIR COMPRESSOR, WITH AIR FEED @ 160.0% OF STOICHIOMETRIC        CASE    AIR4.160
                                                                                        STEP    16
                                                                                        14 JAN 83  9:21:14

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       16 R12T16 WARM SHIFT FEED GAS FROM RAW GAS #12.

THIS STEP RECUPERATES HEAT AGAINST STEP # 12
THE ENTHALPY TRANSFERRED IS:     31.07  MILLIONS BTU


                      STREAM NUMBER 16
          COMPONENT        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
             COS           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 FEED @ 160.0% OF STOICHIOMETRIC          CASE    AIR4.160
                                                                                          STEP    17
                                                                                          14 JAN 83  9:21:14
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      17 R11T17 WARM SHIFT FEED GAS FROM RAW GAS #11.

THIS STEP RECUPERATES HEAT AGAINST STEP # 11
THE ENTHALPY TRANSFERRED IS:     31.42  MILLIONS BTU


                      STREAM NUMBER 17
           COMPONENT       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
              COS           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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AIR4.160  JPL4, 2-STAGE AIR COMPRESSOR, WITH AIR FEED @ 160.0% OF STOICHIOMETRIC             CASE    AIR4.160
                                                                                             STEP    18
                                                                                             14 JAN 83  9:21:14

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        18 R10T18 WARM SHIFT FEED GAS FROM RAW GAS #10.

THIS STEP RECUPERATES HEAT AGAINST STEP # 10
THE ENTHALPY TRANSFERRED IS:    31.77  MILLIONS BTU


                    STREAM NUMBER 18
         COMPONENT        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
             COS          13.7    0.0

           TOTAL       60132.3  100.0

PROCESS GAS TEMPERATURE,    FEED=   544.1        EXIT=   605.9  DEGREES FAHRENHEIT.
PROCESS GAS  ENTHALPY,      FEED=   226.85       EXIT=   258.62 MILLIONS BTU.
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AIR4.160  JPL4. 2-STAGE AIR COMPRESSOR, WITH AIR FEED @ 160.0% OF STOICHIOMETRIC          CASE    AIR4.160
                                                                                         STEP    19
                                                                                         14 JAN 83  9:21:14

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       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

                       FEED GAS        PRODUCT GAS STREAM # 19
           COMPONENT   # 18 MOLS          MOLS       %      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
              COS          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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AIR4.160  JPL4, 2-STAGE AIR COMPRESSOR, WITH AIR FEED @ 160.0% OF STOICHIOMETRIC        CASE    AIR4.160
                                                                                         STEP    20
                                                                                         14 JAN 83  9:21:14

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        20 EXPAND SHIFT (I) EXIT GAS THROUGH POWER RECOVERY TURBINE.

THIS STEP DESCRIBES A PROCESS GAS TURBINE-EXPANDER

                         STREAM NUMBER 20
             COMPONENT        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
                C2H6          0.0    0.0
                H2S         477.9    0.8
                NH3           0.0    0.0
                COS          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:    57045.8 ACFM
```

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AIR4.160  JPL4, 2-STAGE AIR COMPRESSOR, WITH AIR FEED = 160.0% OF STOICHIOMETRIC                CASE     AIR4.160
                                                                                                STEP     21
                                                                                                14 JAN 83  9:21:14

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       21 COOL INTER-STAGE SHIFT GAS IN WHB; RAISE STEAM.

THE DUTY OF THIS PROCESS COOLER IS:   -10.06 MILLIONS BTU PER HOUR.

                    STREAM NUMBER 21
          COMPONENT       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
             C2H6          0.0    0.0
             H2S         477.9    0.8
             NH3           0.0    0.0
             COS          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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AIR4.160  JPL4, 2-STAGE AIR COMPRESSOR, WITH AIR FEED @ 160.0% OF STOICHIOMETRIC           CASE    AIR4.160
                                                                                          STEP    22
                                                                                          14 JAN 83  9:21:14
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        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

                          FEED GAS       PRODUCT GAS STREAM # 22
             COMPONENT   # 21 MOLS         MOLS        %       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       16607.8        18655.5     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
                COS          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.
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AIR4.160  JPL4, 2-STAGE AIR COMPRESSOR, WITH AIR FEED = 160.0% OF STOICHIOMETRIC                     CASE    AIR4.160
                                                                                                     STEP    23
                                                                                                     14 JAN 83  9:21:14

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        23 H23T35 COOL SHIFTED GAS TO WARM SWEET FUEL #35.

THE DUTY OF THIS PROCESS COOLER IS:   -27.13 MILLIONS BTU PER HOUR.

                     STREAM NUMBER 23
           COMPONENT       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
               COS         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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AIR4.160  JPL4, 2-STAGE AIR COMPRESSOR, WITH AIR FEED @ 160.0% OF STOICHIOMETRIC          CASE    AIR4.160
                                                                                           STEP    24
                                                                                           14 JAN 83   9:21:14

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      24 R24T34 COOL SHIFTED GAS TO WARM SWEET FUEL #34.

THE DUTY OF THIS PROCESS COOLER IS:   -48.08 MILLIONS BTU PER HOUR.

                        STREAM NUMBER 24
            COMPONENT        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
               COS            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.
```

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AIR4.160  JPL4, 2-STAGE AIR COMPRESSOR, WITH AIR FEED @ 160.0% OF STOICHIOMETRIC                   CASE     AIR4.160
                                                                                                   STEP    25
                                                                                                   14 JAN 83  9121:14
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       25 R25T33 COOL SHIFTED GAS TO WARM SWEET FUEL #33.

THE DUTY OF THIS PROCESS COOLER IS:   -47.17 MILLIONS BTU PER HOUR.

                     STREAM NUMBER 25
           COMPONENT       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
              COS         13.7    0.0

             TOTAL     60132.3  100.0

PROCESS GAS TEMPERATURE,     FEED=   520.0        EXIT=   430.0  DEGREES FAHRENHEIT.
PROCESS GAS  ENTHALPY,       FEED=   221.83       EXIT=   174.66 MILLIONS BTU.
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AIR4.160  JPL4, 2-STAGE AIR COMPRESSOR, WITH AIR FEED = 160.0% OF STOICHIOMETRIC          CASE    AIR4.160
                                                                                          STEP    26
                                                                                          14 JAN 83  9:21:14

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      26 DESTROY CARBONYL SULFIDE BY HYDROLYSIS.
           THIS STEP REACTS COS WITH WATER VAPOR TO PRODUCE H2S AND CARBON DIOXIDE.

                        FEED STREAM 25          EXIT STREAM 26
          COMPONENT       MOLS      %             MOLS      %

             H2         24953.3   41.5          24963.3   41.5
             CH4         3513.2    5.8           3513.2    5.8
             CO          2860.6    4.8           2860.6    4.8
             CO2        18655.5   31.0          18668.8   31.0
             H2O         9447.7   15.7           9434.4   15.7
             N2           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
             COS           13.7    0.0              0.5    0.0

            TOTAL       60132.3  100.0          60132.3  100.0

EQUILIBRIUM TEMPERATURE =  430.0 DEGREES F            ESSURE =  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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AIR4.160  JPL4, 2-STAGE AIR COMPRESSOR, WITH AIR FEED @ 160.0% OF STOICHIOMETRIC          CASE    AIR4.160
                                                                                          STEP    27
                                                                                          14 JAN 83   9121:14

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       27 R27T32 COOL SHIFTED GAS TO WARM SWEET FUEL #32.

THE DUTY OF THIS PROCESS COOLER IS:   -56.31 MILLIONS BTU PER HOUR.

                          STREAM NUMBER 27
           COMPONENT           MOLS      %

              H2          24963.3    41.5
              CH4          3513.2     5.8
              CO           2860.6     4.8
              CO2         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
              COS             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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AIR4.160  JPL4, 2-STAGE AIR COMPRESSOR, WITH AIR FEED @ 160.0% OF STOICHIOMETRIC          CASE    AIR4.160
                                                                                          STEP    28
                                                                                          14 JAN 83  9:21:14

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     28 COOL TO AMBIENT: DROP CONDENSATE: DUMP HEAT TO AIR & CW.

         THIS STEP COOLS THE GAS TO  100.0 F AND SUBTRACTS CONDENSATE.

                        FEED STREAM 27          EXIT STREAM 28
           COMPONENT       MOLS     %              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
              CO2       18668.8   31.0          18668.8   36.6
              H2O        9434.4   15.7            280.2    0.5
              N2          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
              COS           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 DEW POINT        244.8 F               100.0 F
         WATER SUBTRACTED            9154.2 MOLS
HEAT EXCHANGE DUTIES:
                 GAS COOLING              107.551 MILLIONS BTU
                 WATER CONDENSING         171.093 MILLIONS BTU
                 TOTAL                    278.644 MILLIONS BTU
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AIR4.160  JPL4, 2-STAGE AIR COMPRESSOR, WITH AIR FEED @ 160.0% OF STOICHIOMETRIC          CASE    AIR4.160
                                                                                         STEP    29
                                                                                         14 JAN 83  9121114

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      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.

              FEED GAS STREAM # 28         EXIT GAS STREAM # 29         EXIT GAS STREAM # 94
COMPONENT   ___MOLS_____%____PSIA_      ___MOLS_____%____PSIA_      ___MOLS_____%____PSIA_

   H2       24963.3    49.0    84.7      24962.9    51.9    87.2          0.4     0.0     0.0
   CH4       3513.2     6.9    11.9       3510.9     7.3    12.3          2.2     0.1     0.0
   CO        2860.6     5.6     9.7       2860.4     5.9    10.0          0.2     0.0     0.0
   CO2      18668.8    36.6    63.4      16574.9    34.5    57.9       2093.9    74.5    11.0
   H2O        280.2     0.5     1.0          0.3     0.0     0.0        222.8     7.9     1.2
   N2         200.5     0.4     0.7        200.5     0.4     0.7          0.0     0.0     0.0
   C2H6         0.0     0.0     0.0          0.0     0.0     0.0          0.0     0.0     0.0
   H2S        491.2     1.0     1.7          0.0     0.0     0.0        491.1    17.5     2.6
   NH3          0.0     0.0     0.0          0.0     0.0     0.0          0.0     0.0     0.0
   COS          0.5     0.0     0.0          0.3     0.0     0.0          0.2     0.0     0.0

  TOTAL     50978.1   100.0   173.0      48110.2   100.0   168.0       2810.9   100.0    14.7

 LOW-PRESSURE STEAM USEAGE IS  214. THOUSANDS POUNDS PER HOUR.

POWER CONSUMPTION IS   6958. KILOWATTS.

DRY ACID GAS SENT TO CLAUS PLANT IS 19.00 % H2S.
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          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.

                    FEED STREAM 29        EXIT STREAM 30        EXIT STREAM 31
     COMPONENT        MOLS      %           MOLS      %           MOLS      %

        H2          24962.9   51.9        1872.2   51.9        23090.6   51.9
        CH4          3510.9    7.3         263.3    7.3         3247.6    7.3
        CO           2860.4    5.9         214.5    5.9         2645.9    5.9
        CO2         16574.9   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
        COS             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 FAHRENHEIT.
THE PRESSURE OF THE TWO EXIT GAS STREAMS IS  168.0 PSIA,
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AIR4.160  JPL4, 2-STAGE AIR COMPRESSOR, WITH AIR FEED @ 160.0% OF STOICHIOMETRIC          CASE    AIR4.160
                                                                                          STEP    32
                                                                                          14 JAN 83   9:21:14
*****************************************************************************************************

       32 R27T32 WARM SWEETENED GAS: COOL SHIFT EXIT #27.

THIS STEP RECUPERATES HEAT AGAINST STEP # 27
THE ENTHALPY TRANSFERRED IS:     56.31  MILLIONS BTU


                    STREAM NUMBER 32
      COMPONENT       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
         C2H6           0.0    0.0
         H2S            0.0    0.0
         NH3            0.0    0.0
         COS            0.3    0.0

        TOTAL       44502.0  100.0

PROCESS GAS TEMPERATURE.    FEED=  100.0        EXIT=   258.4  DEGREES FAHRENHEIT.
PROCESS GAS  ENTHALPY.      FEED=    7.92       EXIT=    64.22 MILLIONS BTU.
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AIR4.160  JPL4. 2-STAGE AIR COMPRESSOR. WITH AIR FEED = 160.0% OF STOICHIOMETRIC          CASE    AIR4.160
                                                                                          STEP    33
                                                                                          14 JAN 83  9:21:14

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        33 R25T33 WARM SWEETENED GAS: COOL SHIFT EXIT #25.

THIS STEP RECUPERATES HEAT AGAINST STEP # 25
THE ENTHALPY TRANSFERRED IS:    47.17  MILLIONS BTU


                   STREAM NUMBER 33
         COMPONENT      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
            C2H6         0.0    0.0
            H2S          0.0    0.0
            NH3          0.0    0.0
            COS          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
AIR4.160  JPL4, 2-STAGE AIR COMPRESSOR, WITH AIR FEED @ 160.0% OF STOICHIOMETRIC     STEP    34
                                                                                    14 JAN 83  9:21:14

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      34 R24T34 WARM SWEETENED GAS: COOL SHIFT EXIT #24.

THIS STEP RECUPERATES HEAT AGAINST STEP # 24
THE ENTHALPY TRANSFERRED IS:     48.08  MILLIONS BTU


                     STREAM NUMBER 34
           COMPONENT      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
              C2H6         0.0     0.0
              H2S          0.0     0.0
              NH3          0.0     0.0
              COS          0.3     0.0

             TOTAL     44502.0   100.0

PROCESS GAS TEMPERATURE.     FEED=   385.4        EXIT=   510.9  DEGREES FAHRENHEIT.
PRUCESS GAS  ENTHALPY.       FEED=   111.39       EXIT=   159.47 MILLIONS BTU.
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AIR4.160  JPL4, 2-STAGE AIR COMPRESSOR, WITH AIR FEED @ 160.0% OF STOICHIOMETRIC          CASE    AIR4.160
                                                                                            STEP    35
                                                                                            14 JAN 83  9:21:14

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       35 R23T35 WARM SWEETENED GAS: COOL SHIFT EXIT #23.

THIS STEP RECUPERATES HEAT AGAINST STEP # 23
THE ENTHALPY TRANSFERRED IS:     27.13  MILLIONS BTU

                     STREAM NUMBER 35
          COMPONENT       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
             C2H6          0.0     0.0
             H2S           0.0     0.0
             NH3           0.0     0.0
             COS           0.3     0.0

            TOTAL      44502.0   100.0

PROCESS GAS TEMPERATURE.    FEED=   510.9        EXIT=   580.3  DEGREES FAHRENHEIT.
PROCESS GAS  ENTHALPY.      FEED=   159.47       EXIT=   186.60 MILLIONS BTU.
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AIR4.160  JPL4, 2-STAGE AIR COMPRESSOR, WITH AIR FEED @ 160.0% OF STOICHIOMETRIC           CASE    AIR4.160
                                                                                            STEP    36
                                                                                            14 JAN 83   9:21:14
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      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

                       STREAM NUMBER 36
          COMPONENT        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
             C2H6           0.0    0.0
             H2S            0.0    0.0
             NH3            0.0    0.0
             COS            0.3    0.0

            TOTAL       44502.0  100.0

PROCESS GAS TEMPERATURE,    FEED=   580.3        EXIT=   751.3  DEGREES FAHRENHEIT.
PROCESS GAS  ENTHALPY,      FEED=   186.60       EXIT=   255.14 MILLIONS BTU.
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                                                                                       CASE     AIR4.160
AIR4.160  JPL4, 2-STAGE AIR COMPRESSOR, WITH AIR FEED @ 160.0% OF STOICHIOMETRIC       STEP     37
                                                                                       14 JAN 83  9:21:14

*********************************************************************************************************

     37 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.

                          FEED STREAM 36          EXIT STREAM 37
              COMPONENT     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
                COS           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
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        38 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 # 37
CARBON STABILITY REPORTED AS STREAM # 38                TEMPERATURE: 1200.0 DEGREES F
                                                           PRESSURE:  150.0 PSIA

                                              GRAPHITE      AMORPHOUS
          GAS STREAM     GAS STREAM            LIMIT          LIMIT
             MOLS       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
    COS        0.0        0.0000

   TOTAL   44501.7        1.0000             1.0000         1.0000

                     H*   0.5204       H*    0.5397    H*   0.5362
                     C*   0.1867       C*    0.1561    C*   0.1623
                     O*   0.2929       O*    0.3042    O*   0.3014

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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```
AIR4.160  JPL4, 2-STAGE AIR COMPRESSOR, WITH AIR FEED @ 160.0% OF STOICHIOMETRIC          CASE    AIR4.160
                                                                                          STEP    42
                                                                                          14 JAN 83  9:21:14

********************************************************************************************************

       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

                       FEED STREAM 37         EXIT STREAM 42
          COMPONENT      MOLS      %            MOLS      %

            H2         23090.6   51.9         23090.6   44.0
            CH4         3247.6    7.3          3247.6    6.2
            CO          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
            COS            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 %.     (SET BY USER)
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 OR SUBSEQUENT FEED GAS AS STREAM NUMBER 43

                    STREAM NUMBER 43
         COMPONENT        MOLS      %

            CO2            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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```
AIR4.160  JPL4. 2-STAGE AIR COMPRESSOR. WITH AIR FEED @ 160.0% OF STUICHIOMETRIC                    CASE    AIR4.160
                                                                                                    STEP    44
                                                                                                    14 JAN 83  9:21:14

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        44 CUMPRESS COMBUSTIUN AIR TO FUEL CELL PRESSURE.

THIS STEP DESCRIBES A PROCESS GAS COMPRESSOR OF 2 STAGES.

                         STREAM NUMBER 44
           COMPONENT          MOLS      %

              CO2              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.
KILUWATTS         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:    158.12  MILLIONS BTU


                    STREAM NUMBER 45
        COMPONENT        MOLS      %

           CO2             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=   360.7        EXIT=   504.4  DEGREES FAHRENHEIT.
PROCESS GAS  ENTHALPY,       FEED=   304.74       EXIT=   462.86 MILLIONS BTU.
```

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AIR4.160  JPL4, 2-STAGE AIR COMPRESSOR, WITH AIR FEED = 160.0% OF STOICHIOMETRIC          CASE    AIR4.160
                                                                                          STEP   46
                                                                                          14 JAN 83  9:21:14

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        46 R46F62 - WARM COMBUSTION AIR FROM CATHODE RECYCLE #62.

THE DUTY OF THIS PROCESS HEATER IS:   219.06 MILLIONS BTU PER HOUR.

                          STREAM NUMBER 46
              COMPONENT        MOLS      %

                 CO2            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=   504.4        EXIT=   700.0  DEGREES FAHRENHEIT.
PROCESS GAS  ENTHALPY,      FEED=   462.86       EXIT=   681.91 MILLIONS BTU.
```

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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

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        47 R47F61 - WARM COMBUSTION AIR FROM CATHODE RECYCLE #61.

THE DUTY OF THIS PROCESS HEATER IS:   240.09 MILLIONS BTU PER HOUR.

                         STREAM NUMBER 47
           COMPONENT         MOLS      %

              CO2            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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AIR4.160  JPL4, 2-STAGE AIR COMPRESSOR, WITH AIR FEED = 160.0% OF STOICHIOMETRIC          CASE    AIR4.160
                                                                                           STEP    48
                                                                                           14 JAN 83  9:21:14

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      48 MERGE FRESH AIR #47 WITH RECYCLED CATHODE EXIT GAS #63.

THIS STEP ADDS STREAM # 47 & 63 TO PRODUCE STREAM # 48

                   FEED STREAM 47         FEED STREAM 63         EXIT STREAM 48
   COMPONENT         MOLS      %            MOLS      %            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
AIR4.160  JPL4, 2-STAGE AIR COMPRESSOR, WITH AIR FEED @ 160.0% OF STOICHIOMETRIC      STEP    49
                                                                                     14 JAN 83  9:21:14

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      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
           COS            0.0    0.0

          TOTAL       91904.7  100.0

        TEMPERATURE        883.5 DEGREES F
           PRESSURE        156.0 PSIA

                       FEED STREAM 48        EXIT STREAM 49
        COMPONENT        MOLS      %            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             1046.4 F
          PRESSURE         156.0 PSIA           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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AIR4.160  JPL4, 2-STAGE AIR COMPRESSOR, WITH AIR FEED @ 160.0% OF STOICHIOMETRIC            CASE     AIR4.160
                                                                                            STEP    50
                                                                                            14 JAN 83   9:21:14

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      50 JOIN THE TWO FUEL CELL FEEDS, ANODE #42, CATHODE #49.

THIS STEP ADDS STREAM # 42 & 49 TO PRODUCE STREAM # 50

                 FEED STREAM 42         FEED STREAM 49         EXIT STREAM 50
   COMPONENT       MOLS       %           MOLS       %           MOLS       %

      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

                 FEED STREAM 42         FEED STREAM 49         EXIT STREAM 50
   COMPONENT       MOLS       %           MOLS       %           MOLS       %

      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                1046.4                 1001.5
     PRESSURE       1500.0                 152.0                  152.0
```

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        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

                       FEED STREAM 50          EXIT STREAM 51
          COMPONENT      MOLS      %              MOLS      %

             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
             COS           0.0    0.0               0.0    0.0

            TOTAL      52516.4  100.0           91904.7  100.0

            CATHODE

                       FEED STREAM 50          EXIT STREAM 51
          COMPONENT      MOLS      %              MOLS      %

             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
             O2        35706.9    9.7           19248.0    6.1

            TOTAL     366769.9  100.0          317393.0  100.0
```

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AIR4.160  JPL4. 2-STAGE AIR COMPRESSOR, WITH AIR FEED @ 160.0% OF STOICHIOMETRIC     CASE    AIR4.160
                                                                                     STEP    51
                                                                                     14 JAN 83   9:21:14

*****************************************************************************************************

       51 THE FUEL CELL, MAXIMUM CONVERSION, REFORM WITHIN ANODE.
                          AVERAGE QUANTITIES (LOG MEAN):
         ________ANODE_________                           ________CATHODE________
           MOL          PARTIAL                              MOL        PARTIAL
         FRACTION    PRESSURE(ATM)                         FRACTION   PRESSURE(ATM)
   H2     0.1616        1.649                    CO2       0.1449        1.479
   CH4    0.0101        0.103                    H2O       0.2009        2.050
   CO     0.0372        0.379                    N2        0.5707        5.823
   CO2    0.4125        4.210                    O2        0.0776        0.791
   H2O    0.2465        2.515
   N2     0.0027        0.028

CHARACTERISTIC CELL TEMPERATURE=  922.0 DEG. K      1200.0 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:  0.3243E+10 BTU/HR. (ISOTHERMAL)
FRACTION OF POTENTIAL ENTHALPY RELEASE EXTRACTED AS POWER:  70.0 %
CELL AREA =  0.5336E+10 CM2          0.5744E+07 SQUARE FEET
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
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AIR4.160  JPL4, 2-STAGE AIR COMPRESSOR, WITH AIR FEED @ 160.0% OF STOICHIOMETRIC            CASE    AIR4.160
                                                                                            STEP   52
                                                                                            14 JAN 83  9:21:14

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       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.

                   FEED GAS       PRODUCT GAS STREAM # 52
         COMPONENT  # 51 MOLS         MOLS       %      PSIA

             H2      3326.2         3326.2     3.6       5.2
             CH4       12.4           12.4     0.0       0.0
             CO      2433.1         2433.1     2.6       3.8
             CO2    51697.6        51697.6    56.3      81.6
             H2O    34249.9        34249.9    37.3      54.0
             N2       185.5          185.5     0.2       0.3
             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
             COS        0.0            0.0     0.0       0.0

            TOTAL   91904.7        91904.7   100.0     145.0

                   FEED GAS       PRODUCT GAS STREAM # 53
         COMPONENT  # 51 MOLS         MOLS       %      PSIA

             CO2    35141.1        35141.1    11.1      16.1
             H2O    68471.4        68471.4    21.6      31.3
             N2    194532.5       194532.5    61.3      88.9
             O2     19248.0        19248.0     6.1       8.8

            TOTAL  317393.0       317393.0   100.0     145.0

              TEMPERATURE = 1259.6 DEGREES F.
              PRESSURE =      145.0 PSIA
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AIR4.160  JPL4, 2-STAGE AIR COMPRESSOR, WITH AIR FEED = 160.0% OF STOICHIOMETRIC        CASE    AIR4.160
                                                                                        STEP   54
                                                                                        14 JAN 83  9:21:14

*************************************************************************************************************

      54 COOL ANODE EXIT GAS IN WHB - RAISE STEAM.

THE DUTY OF THIS PROCESS COOLER IS:  -338.74 MILLIONS BTU PER HOUR.

                      STREAM NUMBER 54
            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
              COS             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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AIR4.160  JPL4, 2-STAGE AIR COMPRESSOR, WITH AIR FEED = 160.0% OF STOICHIOMETRIC        CASE    AIR4.160
                                                                                        STEP    55
                                                                                        14 JAN 83  9:21:14

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        55 R36F55 COOL ANODE EXIT GAS TO WARM ANODE FEED GAS #36.

THE DUTY OF THIS PROCESS COOLER IS:   -68.53 MILLIONS BTU PER HOUR.

                          STREAM NUMBER 55
           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
              COS             0.0     0.0

             TOTAL        91904.7   100.0

PROCESS GAS TEMPERATURE,     FEED=   925.0        EXIT=   855.0  DEGREES FAHRENHEIT.
PROCESS GAS  ENTHALPY,       FEED=   757.24       EXIT=   688.70 MILLIONS BTU.
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AIR4.160  JPL4, 2-STAGE AIR COMPRESSOR, WITH AIR FEED @ 160.0% OF STOICHIOMETRIC      CASE    AIR4.160
                                                                                      STEP    57
                                                                                      14 JAN 83  9:21:14

*****************************************************************************************************

        57 ANODE EXIT GAS BOOSTER FAN, TO FEED CATALYTIC BURNER.

THIS STEP DESCRIBES A PROCESS GAS COMPRESSOR

                      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
              COS            0.0    0.0

             TOTAL       91904.7  100.0

THE COMPRESSION RATIO (PER STAGE) IS 1.11


                 ENERGY REQUIREMENTS
                 PER STAGE      TOTAL
HORSEPOWER         11167.      11167.
KILOWATTS           8325.       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
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      SPLIT CATHODE EXIT. COOLED RECYCLE (#58). STACK EXIT (#59).
THIS STEP SPLITS STREAM 53  INTO SIDESTREAM 58 AND DOWNSTREAM, 59 SIDESTREAM 58 IS RECYCLED.

                FEED STREAM 53        EXIT STREAM 58        EXIT STREAM 59
   COMPONENT      MOLS      %            MOLS      %           MOLS      %

      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 COMPRESSUR, WITH AIR FEED = 160.0% OF STUICHIOMETRIC                CASE    AIR4.160
                                                                                                STEP    60
                                                                                                14 JAN 83  9:21:14
****************************************************************************************************************

        60 COOL CATHODE RECYCLE GAS IN WHB - RAISE STEAM.

THE DUTY OF THIS PROCESS COOLER IS:  -157.28 MILLIONS BTU PER HOUR.

                          STREAM NUMBER 60
            COMPONENT         MOLS      %

               CO2         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=  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 @ 160.0% OF STOICHIOMETRIC          CASE    AIR4.160
                                                                                           STEP   61
                                                                                           14 JAN 83  9:21:14
*************************************************************************************************************

        61 R47F61 COOL CATHODE RECYCLE TO WARM COMBUSTION AIR #47.

THIS STEP RECUPERATES HEAT AGAINST STEP # 47
THE ENTHALPY TRANSFERRED IS:  -240.09  MILLIONS BTU

                        STREAM NUMBER 61
              COMPONENT      MOLS      %

                  CO2       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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AIR4.160  JPL4, 2-STAGE AIR COMPRESSOR, WITH AIR FEED = 160.0% OF STOICHIOMETRIC           CASE    AIR4.160
                                                                                          STEP    62
                                                                                          14 JAN 83  9:21:14

*****************************************************************************************************************

        62 H46F62 COOL CATHODE RECYCLE TO WARM COMBUSTION AIR #46.

THIS STEP RECUPERATES HEAT AGAINST STEP # 46
THE ENTHALPY TRANSFERRED IS:  -219.06  MILLIONS BTU


                         STREAM NUMBER 62
            COMPONENT         MOLS      %

               CO2         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=    889.3        EXIT=    677.6  DEGREES FAHRENHEIT.
PROCESS GAS  ENTHALPY,       FEED=    804.25       EXIT=    585.20 MILLIONS BTU.
```

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AIR4.160  JPL4. 2-STAGE AIR COMPRESSOR. WITH AIR FEED @ 160.0% OF STOICHIOMETRIC          CASE    AIR4.160
                                                                                          STEP    63
                                                                                          14 JAN 83  9:21:14

****************************************************************************************************

       63 CATHODE RECYCLE GAS CIRCULATION FAN.

THIS STEP DESCRIBES A PROCESS GAS COMPRESSOR

                         STREAM NUMBER 63
           COMPONENT          MOLS      %

               CO2        13915.9    11.1
               H2O        27114.7    21.6
               N2         77034.9    61.3
               O2          7622.2     6.1

             TOTAL       125687.6   100.0

THE COMPRESSION RATIO (PER STAGE) IS 1.17


              ENERGY REQUIREMENTS
              PER STAGE     TOTAL
HORSEPOWER      20749.      20749.
KILOWATTS       15468.      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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AIR4.160  JPL4. 2-STAGE AIR COMPRESSOR, WITH AIR FEED # 160.0% OF STOICHIOMETRIC              CASE    AIR4.160
                                                                                              STEP    64
                                                                                              14 JAN 83  9:21:14
************************************************************************************************************

       64 EXPAND STACK GAS THROUGH TURBINE TO RECOVER POWER.

THIS STEP DESCRIBES A PROCESS GAS TURBINE-EXPANDER

                       STREAM NUMBER 64
          COMPONENT          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 DEGREES F.

INLET PRESSURE IS:      142.0      DISCHARGE IS:   17.0 PSIA

INLET VOLUME IS:    400170.4      DISCHARGE IS: 2143494.8 ACFM
```

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AIR4.160  JPL4, 2-STAGE AIR COMPRESSOR, WITH AIR FEED @ 160.0% OF STOICHIOMETRIC                 CASE    AIR4.160
                                                                                                 STEP    65
                                                                                                 14 JAN 83   9:21:14

*****************************************************************************************************************

       65 R45F65 COOL STACK GAS TO WARM COMBUSTION AIR #45.

THE DUTY OF THIS PROCESS COOLER IS:  -158.12 MILLIONS BTU PER HOUR.

                         STREAM NUMBER 65
           COMPONENT        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 FAHRENHEIT.
PROCESS GAS  ENTHALPY,       FEED=   838.61      EXIT=   680.49 MILLIONS BTU.
```

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                                                                                              CASE    AIR4.160
AIR4.160  JPL4, 2-STAGE AIR COMPRESSOR, WITH AIR FEED @ 160.0% OF STOICHIOMETRIC              STEP    66
                                                                                              14 JAN 83  9:21:14

********************************************************************************************************

      66 READ COMBUSTION FUEL VALUE OF RAW GAS FROM GASIFIER.
               THIS ROUTINE GIVES COMBUSTION HEAT VALUES.


STEP IDENTIFICATION # 66   SOURCE STREAM #  2

                TOTAL FUEL GAS         DRIED FUEL GAS          SWEETENED GAS
    COMPONENT     MOLS      %            MOLS      %             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
      C2H6          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             .
      COS          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.                137610.
            GROSS      XXXXXX             120101.                150796.

   BTU/CU FT, NET         207.               289.                   362.
            GROSS         XXX                316.                   397.

TOTAL STREAM COMBUSTION HEAT RELEASE, MILLIONS BTU:
              NET       4134.4             4124.2                 4014.4
            GROSS      XXXXXX             4518.0                 4399.1
```

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                                                                                         CASE     AIR4.160
AIR4.160  JPL4, 2-STAGE AIR COMPRESSOR, WITH AIR FEED # 160.0% OF STOICHIOMETRIC         STEP     67
                                                                                         14 JAN 83  9:21:14

****************************************************************************************************

      67 READ COMBUSTION FUEL VALUE OF SHIFTED SOUR GAS.
              THIS ROUTINE GIVES COMBUSTION HEAT VALUES.


STEP IDENTIFICATION # 67   SOURCE STREAM # 31

                   TOTAL FUEL GAS          DRIED FUEL GAS          SWEETENED GAS
    COMPONENT        MOLS      %             MOLS      %             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
      COS              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.                  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       3846.4                  3846.4                  3846.3
           GROSS       XXXXXX                  4411.6                  4411.5
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AIR4.160  JPL4, 2-STAGE AIR COMPRESSOR, WITH AIR FEED @ 160.0% OF STOICHIOMETRIC        CASE    AIR4.160
                                                                                        STEP    68
                                                                                        14 JAN 83   9:21:14

*******************************************************************************************************

      68 READ COMBUSTION FUEL VALUE OF SWEET ANODE FEED GAS.
               THIS ROUTINE GIVES COMBUSTION HEAT VALUES.

STEP IDENTIFICATION # 68   SOURCE STREAM # 37

                    TOTAL FUEL GAS           DRIED FUEL GAS            SWEETENED GAS
    COMPONENT         MOLS     %               MOLS     %                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
      COS              0.0    0.0               0.0    0.0

     TOTAL         44501.7  100.0           44501.4  100.0           29169.6  100.0

   BTU/MOL.   NET      86430.                  86431.                  131859.
            GROSS      XXXXXX                  99133.                  151238.

   BTU/CU FT. NET        227.                    227.                     347.
            GROSS        XXX                     261.                     398.

TOTAL STREAM COMBUSTION HEAT RELEASE, MILLIONS BTU:
              NET      3846.3                  3846.3                  3846.3
            GROSS      XXXXXX                  4411.5                  4411.5
```

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AIR4.160  JPL4, 2-STAGE AIR COMPRESSOR, WITH AIR FEED # 160.0% OF STOICHIOMETRIC          CASE    AIR4.160
                                                                                           STEP    69
                                                                                           14 JAN 83  9:21:14

*****************************************************************************************************

       69 READ COMBUSTION FUEL VALUE OF ANODE EXIT GAS.
                THIS ROUTINE GIVES COMBUSTION HEAT VALUES.

STEP IDENTIFICATION # 69   SOURCE STREAM # 54

                  TOTAL FUEL GAS        DRIED FUEL GAS         SWEETENED GAS
      COMPONENT     MOLS     %            MOLS     %             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
        COS          0.0    0.0            0.0    0.0

       TOTAL     91904.7  100.0        57654.8  100.0          5957.2  100.0

   BTU/MOL,   NET       7037.               11217.                108558.
            GROSS     XXXXXX                12327.                119306.

   BTU/CU FT, NET         19.                  30.                   286.
            GROSS        XXX                   32.                   314.

TOTAL STREAM COMBUSTION HEAT RELEASE, MILLIONS BTU
              NET      646.7                646.7                  646.7
            GROSS     XXXXXX                710.7                  710.7
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AIR4.160  JPL4. 2-STAGE AIR COMPRESSOR, WITH AIR FEED @ 160.0% OF STOICHIOMETRIC          CASE    AIR4.160
                                                                                          STEP    70
                                                                                          14 JAN 83  9:21:14

******************************************************************************************************

      70 COMPRESS RECYCLE GAS TO INJECT FEED COAL TO GASIFIER.

THIS STEP DESCRIBES A PROCESS GAS COMPRESSOR OF 2 STAGES.

                          STREAM NUMBER 70
              COMPONENT         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
                 COS             0.0     0.0

                TOTAL         3608.3   100.0

THE COMPRESSION RATIO (PER STAGE) IS 1.84


                ENERGY REQUIREMENTS
                PER STAGE      TOTAL
HORSEPOWER         1206.       2411.
KILOWATTS           899.       1797.

ISENTROPIC 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 # 1      168.0     309.6                         2072.4    1335.8
    STAGE # 2      298.4     550.0                         1166.7     752.0
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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

                   TOTAL FUEL GAS           DRIED FUEL GAS            SWEETENED GAS
    COMPONENT       MOLS      %              MOLS      %               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
       COS            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
AIR4.160  JPL4, 2-STAGE AIR COMPRESSOR, WITH AIR FEED @ 160.0% OF STOICHIOMETRIC  STEP    73
                                                                                  14 JAN 83  9:21:14

**********************************************************************************************

       73 READ SENSIBLE HEATS OF REACTOR EXIT RAW GAS (INCL. #71)

THE DUTY OF THIS PROCESS COOLER IS:  -208.24 MILLIONS BTU PER HOUR.

                     STREAM NUMBER 73
          COMPONENT        MOLS     %

             H2        15505.4   27.6
             CH4        3513.2    6.3
             CO        12318.5   21.9
             CO2        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
             COS          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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******JOB EXITED ON 14 JAN 83  9:21:37   ******

******CPU TIME=       0MINS    7.44SECS  ******

******CONNECT TIME=       0MINS   36SECS  ******
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APPENDIX Ce. Printouts of Gasifier Balance for Non-Confidential Fluidized-Bed Cases

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IDENTIFICATION - U-GAS DESIGN FOR H2 PRODUCTION FOR FUEL CELL - FLEMING   [illegible]U 6   CASE
                                   U-GAS GASIFIER MATERIAL AND ENERGY BALANCE
                                   ------------------------------------------
                                                                                                    RECYCLE      PRODUCT
      SOLIDS FLOW,MOLES     FEED        FINES    AGGLOMERATES     GAS FLOW,MOLES     FEED1       FEED2        GAS          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
                                                                        COS         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
                                   ------------------------------------------
                                                                                                    RECYCLE      PRODUCT
        SOLIDS FLOW,LBS     FEED        FINES    AGGLOMERATES     GAS FLOW,LBS       FEED1       FEED2        GAS          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      0.0000      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
                                                                        COS         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       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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IDENTIFICATION - U-GAS DESIGN FOR H2 PRODUCTION FOR FUEL CELL - FLEMING     ILL NO 6   CASE 1
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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  (    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       CO2       H2       H2O       CH4      O2       N2       H2S      COS      NH3      HCN
                                        25.127   14.364   29.545   25.925    3.717   0.000   0.350   0.821   0.023   0.128   0.000
               H2O AND NH3 FREE:        33.979   19.425   39.954     -       5.027   0.000   0.473   1.110   0.032     -     0.000
   MOLECULAR WEIGHT = 19.6355   (GAS VOLUME =   7013.403 CU FT : AT REACTOR TEMP AND PRESSURE)
RECYCLE GAS COMPOSITION, MOLE PERCENT:     CO       CO2       H2       H2O       CH4      O2       N2       H2S      COS      NH3      HCN
                                        34.371   19.649   40.415    0.000    5.085   0.000   0.480   0.000   0.000   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 HHV       =  0.8242

   CARBON GASIFICATION EFFICIENCY     = 98.49 %
   CARBON UTILIZATION EFFICIENCY      = 96.88 %
   MAF COAL GASIFICATION EFFICIENCY   = 98.76 %
   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        = 0.3525 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)
   COAL 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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IDENTIFICATION - U-GAS DESIGN FOR H2 PRODUCTION FOR FUEL CELL - FLEMING      ILL NO 6   CASE 2

                                                          U-GAS GASIFIER MATERIAL AND ENERGY BALANCE
                                                          ------------------------------------------
                                                                                                  RECYCLE      PRODUCT
     SOLIDS FLOW,MOLES       FEED       FINES   AGGLOMERATES    GAS FLOW,MOLES     FEED1      FEED2       GAS          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            O(2)       18.8314     0.0000     0.0000       0.0000
                                                                      N(2)        0.0946     0.0000     0.0376       0.4834
            MOIST.         1.4233      0.0000       0.0000            H(2)S       0.0000     0.0000     0.0000       1.1488
                                                                      COS         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      97.9026            TOTAL       18.9261    57.8803     7.5000     133.9705

TEMPERATURE, DEGREE F         80.       1875.        1875.   TEMPERATURE, DEGREE F   250.       570.       600.        1875.


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 * FLOW RATE IS IN LBS
                                                          U-GAS GASIFIER MATERIAL AND ENERGY BALANCE
                                                          ------------------------------------------
                                                                                                  RECYCLE      PRODUCT
        SOLIDS FLOW,LBS      FEED       FINES   AGGLOMERATES     GAS FLOW,LBS      FEED1      FEED2       GAS          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  1042.7136     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            O(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
                                                                      COS         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      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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IDENTIFICATION - U-GAS DESIGN FOR H2 PRODUCTION FOR FUEL CELL - FLEMING      ILL NO 6   CASE 2
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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      CO2      H2      H2O      CH4      O2      N2      H2S     COS     NH3     HCN
                                        23.549  15.475  26.525  26.754    6.321   0.000   0.361   0.858   0.025   0.133   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      H2O      CH4      O2      N2      H2S     COS     NH3     HCN
                                        32.609  21.413  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       = 96.88 %
   MAF COAL GASIFICATION EFFICIENCY    = 98.73 %
   MAF COAL UTILIZATION EFFICIENCY     = 97.40 %
   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    0.0823 CU FT/LB (AT  515.0 PSIA AND  80. F)
   RECYCLE GAS/COAL (MF)               = 0.1563 LB/LB    0.0844 CU FT/LB (AT  515.0 PSIA AND  80. F)
   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    37.4504
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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 (Coal Feed - 4991 Tons/Day)

Case	System	Power Output Net Mech	Fuel Cell	Steam	Net	Eff, %	Comments
		MW	MW	MW	MW		
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_2S				806-808		(2 Cases)
K	As I, Burn Part of Anode Exhaust to Cathode Exhaust				816-817		(2 Cases)
L	As I, Low ΔP 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 CO_2 to Cathode

— As O but Minimum Cell Thickness, Other Future Improvements.

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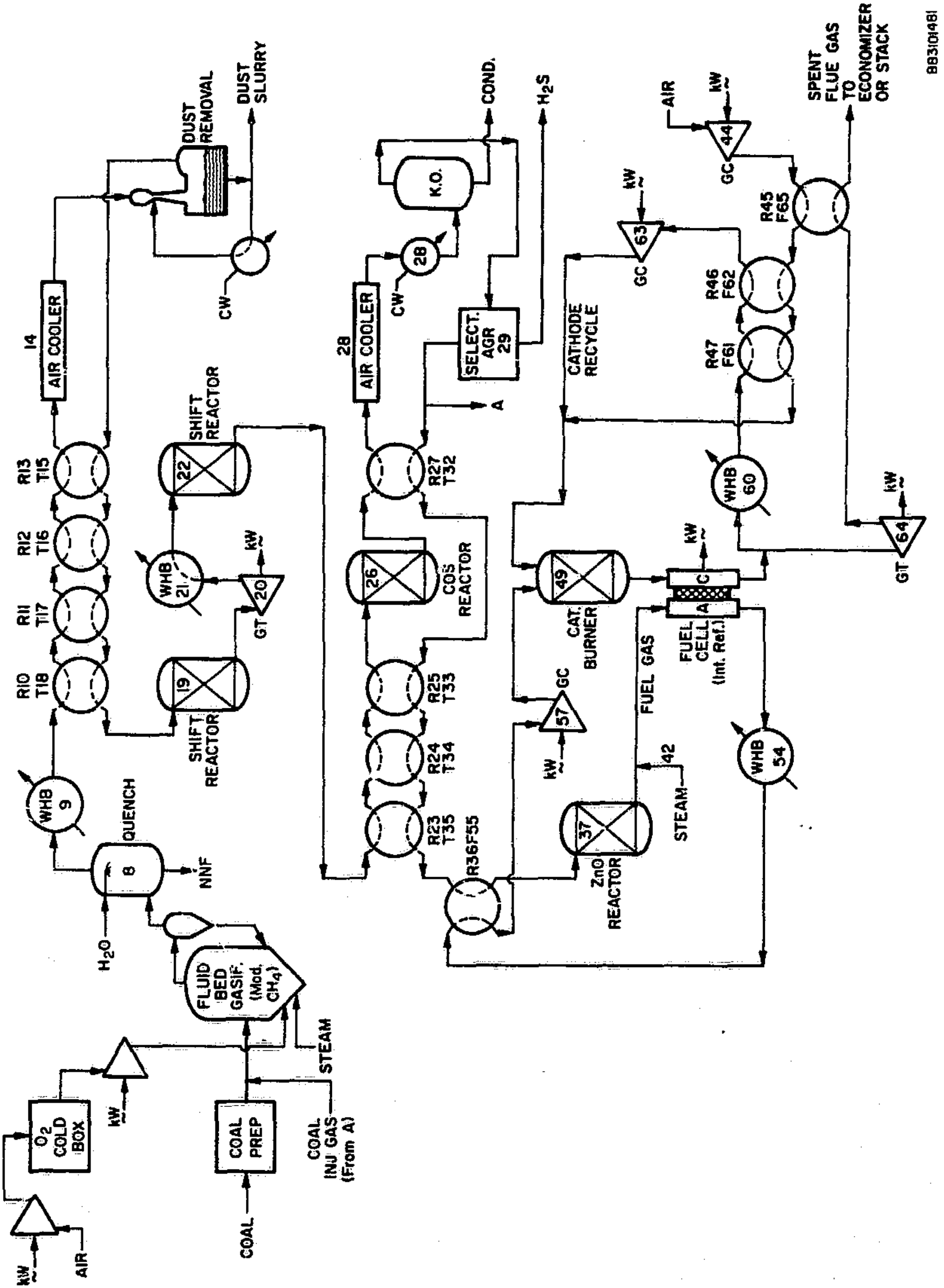

Figure Da-1. PROCESS FLOW DIAGRAM FOR THE GENERAL SYSTEM EMPLOYED IN ANALYSIS OF INITIAL ASSUMPTIONS

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% H_2S; 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 that 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 H_2S concentration. This approach removes more CO_2 with the H_2S 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 CO_2, 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 SO_2 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 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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JPL.UG.4  FEED TO BED. REFORM. H2S CONV. QUENCH. PRE-SHIFT. COSA. 13.5% CH4 RAW            CASE    JPL.UG.4
                                                                                                SUMMARY
                                                                                          4 JAN 83 14135119
             METHANE-RICH O-GAS WITH ANODE METHANE REFORMING. 13.5% C TO C

  THIS IS THE FINAL SYNOPSIS OF THE ENERGY BUDGET OF AN INTEGRATED POWER PLANT USING COAL
GASIFICATION AND ELECTROCHEMICAL FUEL CELLS. ALONG WITH CONVENTIONAL POWER GENERATION MACHINERY.

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        *********         COAL CONSUMPTION IS   4391. TONS PER DAY.        *********
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        *********         NET DELIVERED POWER IS  706. MEGAWATTS.          *********
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                                             DETAILS

COAL CONSUMPTION IS  416. THOUSANDS POUNDS PER HOUR. OR ABOUT  5314. MILLIONS BTU.
THE OXYGEN PLANT SIZE IS  3023. TONS PER DAY.
PROCESS STEAM CONSUMPTION IS   742. THOUSANDS POUNDS PER HOUR.

PROCESS MACHINERY (COMPRESSORS AND MAJOR PUMPS) NEED                    -269030. KILOWATTS
PROCESS GAS TURBO-EXPANDERS & CLOSELY-ASSOCIATED STEAM TURBINES GIVE     321056. KILOWATTS
CONDENSING STEAM TURBINE (MULTI-STAGE W. REHEAT) GIVES                     3653. KILOWATTS
THE FUEL CELL ITSELF DELIVERS                                            650614. KILOWATTS
                     (AFTER CONVERSION TO 13.8KV-AC @ 98% EFF.)
THE NET POWER DELIVERED TO BUSBAR IS                                     706294. KILOWATTS

THE VOLTAGE OF THE FUEL CELL IS            0.8294 VOLTS.
THE AREA OF THE FUEL CELL IS            5744276. SQ. FT.

OVERALL PLANT EFFICIENCY. COAL PILE TO BUSBAR. IS 45.4 PERCENT.
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JPL.UG.4  FLUID BED, REFORM, 15% CONV, QUENCH, PRE-SHIFT, COSX, 13.5% CH4 RAW          CASE    JPL.UG.4
                                                                                        POWER LIST
                                                                                     4 JAN 83 14:35:19

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TABULATION OF SHAFT ENERGY AVAILABLE OR NEEDED.

  97   -39644.7 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    -6457.4 KILOWATTS      29 REMOVE SULFUR FROM RAW GAS BY SELEXOL SYSTEM.
  42    10667.5 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   663892.0 KILOWATTS      51 THE FUEL CELL, MAXIMUM CONVERSION, REFORM WITHIN ANODE.
  57    -6324.6 KILOWATTS      57 ANODE EXIT GAS BOOSTER FAN, TO FEED CATALYTIC BURNER.
  63   -19446.7 KILOWATTS      63 CATHODE RECYCLE GAS CIRCULATION FAN.
  64   278632.7 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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JPL.UG.4  FLUID BED, REFORM, P39 CONV, QUENCH, PRE-SHIFT, COSA, 13.5% CH4 RAW              CASE    JPL.UG.4
                                                                                               STEAM LIST
                                                                                          4 JAN 83 14:35:19
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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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JPL.UG.4  FLUID BED. REFORM. H34 CONV. QUENCH. PRE-SHIFT. COSX. 13.5% CH4 RAW                  CASE    JPL.UG.4
                                                                                                  WASTE HEAT OUT
                                                                                                  6 JAN 83 14:35:19

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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.      10.00 MILLIONS BTU    21 COOL INTER-STAGE SHIFT GAS IN WHBT RAISE STEAM.
  54     925.0 DEGREES F.     319.00 MILLIONS BTU    54 COOL ANODE EXIT GAS IN WHB - RAISE STEAM.
  60    1090.0 DEGREES F.     199.06 MILLIONS BTU    60 COOL CATHODE RECYCLE GAS IN WHB - RAISE STEAM.
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CASE B

Utilize Two-Stage Air Compression

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                                                                                        CASE    JPL.2S.4
JPL.2S.4  FLUID BED. REFORM. 85% CONV. QUENCH. PRE-SHIFT. COSX. 13.5% CH4 RAW                SUMMARY
                                                                                        13 JAN 83 10:14:26
          METH-RICH U-GAS WITH ANODE METHANE REFORMING. 13.5% C TO C

  THIS IS THE FINAL SYNOPSIS OF THE ENERGY BUDGET OF AN INTEGRATED POWER PLANT USING COAL
GASIFICATION AND ELECTROCHEMICAL FUEL CELLS. ALONG WITH CONVENTIONAL POWER GENERATION MACHINERY.

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        *********       COAL CONSUMPTION IS   4991. TONS PER DAY.    *********
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        *********       NET DELIVERED POWER IS  710. MEGAWATTS.      *********
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_____________________________________DETAILS___________________________________________

COAL CONSUMPTION IS  416. THOUSANDS POUNDS PER HOUR. OR ABOUT  5314. MILLIONS BTU.
THE OXYGEN PLANT SIZE IS  3023. TONS PER DAY.
PROCESS STEAM CONSUMPTION IS   792. THOUSANDS POUNDS PER HOUR.

PROCESS MACHINERY (COMPRESSORS AND MAJOR PUMPS) NEED                    -270046. KILOWATTS
PROCESS GAS TURBO-EXPANDERS & CLOSELY-ASSOCIATED STEAM TURBINES GIVE     324181. KILOWATTS
CONDENSING STEAM TURBINE (MULTI-STAGE W. REHEAT) GIVES                     4933. KILOWATTS
THE FUEL CELL ITSELF DELIVERS                                            650614. KILOWATTS
                 (AFTER CONVERSION TO 13.6KV-AC @ 98% EFF.)
THE NET POWER DELIVERED TO BUSBAR IS                                     709682. KILOWATTS

THE VOLTAGE OF THE FUEL CELL IS         0.8294 VOLTS.
THE AREA OF THE FUEL CELL IS          5744276. SQ. FT.

OVERALL PLANT EFFICIENCY. COAL PILE TO BUSBAR. IS 45.6 PERCENT.
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JPL.2S.4  FLUID BED, REFORM, 85% CONV, QUENCH, PRE-SHIFT, COSX, 13.5% CH4 RAW                    CASE     JPL.2S.4
                                                                                                    POWER LIST
                                                                                                 13 JAN 83 10:14:26

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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 -171083.5 KILOWATTS      44 COMPRESS COMBUSTION AIR TO FUEL CELL PRESSURE.
  51  663891.7 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  -20371.1 KILOWATTS      63 CATHODE RECYCLE GAS CIRCULATION FAN.
  64  281757.5 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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JPL.2S.4  FLUID BED, REFORM, 85% CONV, QUENCH, PRE-SHIFT, COSX, 13.5% CH4 RAW                    CASE    JPL.2S.4
                                                                                                    STEAM LIST
                                                                                                13 JAN 83 10:14:26
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TABULATION OF STEAM NEEDED IN PROCESS.

 2        500.0 PSIA            24072. MULS STEAM  2 U-GAS GASIFIER NORMAL OPERATION. STREAM #2 = NET MAKE.
29         65.0 PSIA            11889. MULS STEAM  29 REMOVE SULFUR FROM RAW GAS BY SELEXOL SYSTEM.
42       1500.0 PSIA             8015. MULS STEAM  42 ADD STEAM VIA TURBINE: PREVENT CARBON DROP IN FUEL CELL.
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JPL.2S.4  FLUID BED, REFORM, 85% CONV, QUENCH, PRE-SHIFT, COSX, 13.5% CH4 RAW                    CASE     JPL.2S.4
                                                                                                  WASTE HEAT OUT
                                                                                                13 JAN 83 10:14:26

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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.     338.82 MILLIONS BTU    54 COOL ANODE EXIT GAS IN WHB - 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

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AIR4.160  JPL4, 2-STAGE AIR COMPRESSOR, WITH AIR FEED @ 160.0% OF STOICHIOMETRIC          CASE     AIR4.160
                                                                                                   SUMMARY
             METH-RICH U-GAS WITH ANODE METHANE REFORMING, 13.5% C TO C                   14 JAN 83  9:21:14

  THIS IS THE FINAL SYNOPSIS OF THE ENERGY BUDGET OF AN INTEGRATED POWER PLANT USING COAL
GASIFICATION AND ELECTROCHEMICAL FUEL CELLS, ALONG WITH CONVENTIONAL POWER GENERATION MACHINERY.

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        *********          COAL CONSUMPTION IS   4991. TONS PER DAY.         *********
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        *********          NET DELIVERED POWER IS  712. MEGAWATTS.           *********
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____________________________________________DETAILS___________________________________________

COAL CONSUMPTION IS  416. THOUSANDS POUNDS PER HOUR, OR ABOUT  5314. MILLIONS BTU.
THE OXYGEN PLANT SIZE IS  3023. TONS PER DAY.
PROCESS STEAM CONSUMPTION IS   792. THOUSANDS POUNDS PER HOUR.

PROCESS MACHINERY (COMPRESSORS AND MAJOR PUMPS) NEED                     -268075. KILOWATTS
PROCESS GAS TURBO-EXPANDERS & CLOSELY-ASSOCIATED STEAM TURBINES GIVE      327910. KILOWATTS
CONDENSING STEAM TURBINE (MULTI-STAGE W. REHEAT) GIVES                      1290. KILOWATTS
THE FUEL CELL ITSELF DELIVERS                                             651336. KILOWATTS
                  (AFTER CONVERSION TO 13.6KV-AC @ 98% EFF.)
THE NET POWER DELIVERED TO BUSBAR IS                                      712461. KILOWATTS

THE VOLTAGE OF THE FUEL CELL IS         0.8303 VOLTS.
THE AREA OF THE FUEL CELL IS          5744276. SQ. FT.

OVERALL PLANT EFFICIENCY, COAL PILE TO BUSBAR, IS 45.8 PERCENT.
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AIR4.160  JPL4, 2-STAGE AIR COMPRESSOR, WITH AIR FEED @ 160.0% OF STOICHIOMETRIC        CASE    AIR4.160
                                                                                           POWER LIST
                                                                                        14 JAN 83   9:21:14

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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 -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 COAL 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

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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
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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.     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

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                                                                                              CASE    RUN4.1
RUN4.1    JPL.UG.4 BASE CASE, FC REFORM, PRE-SHIFT, 13.5% CARB EX COAL TO CH4.                     SUMMARY
                                                                                               4 FEB 83 14:52:02
             NEW BASE CASE WITH PRE-QUENCH RAW GAS HEAT RECOVERY.

  THIS IS THE FINAL SYNOPSIS OF THE ENERGY BUDGET OF AN INTEGRATED POWER PLANT USING COAL
GASIFICATION AND ELECTROCHEMICAL FUEL CELLS, ALONG WITH CONVENTIONAL POWER GENERATION MACHINERY.

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        *********         COAL CONSUMPTION IS   4991. TONS PER DAY.        *********
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        *********         NET DELIVERED POWER IS  728. MEGAWATTS.          *********
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____________________________________________DETAILS_________________________________________

COAL CONSUMPTION IS  416. THOUSANDS POUNDS PER HOUR, OR ABOUT  5314. MILLIONS BTU.
THE OXYGEN PLANT SIZE IS  3023. TONS PER DAY.
PROCESS STEAM CONSUMPTION IS   791. THOUSANDS POUNDS PER HOUR.

PROCESS MACHINERY (COMPRESSORS AND MAJOR PUMPS) NEED                     -280188. KILOWATTS
PROCESS GAS TURBO-EXPANDERS & CLOSELY-ASSOCIATED STEAM TURBINES GIVE      340004. KILOWATTS
CONDENSING STEAM TURBINE (MULTI-STAGE W. REHEAT) GIVES                     15356. KILOWATTS
THE FUEL CELL ITSELF DELIVERS                                             653246. KILOWATTS
                  (AFTER CONVERSION TO 13.6KV-AC @ 98% EFF.)
THE NET POWER DELIVERED TO BUSBAR IS                                      728418. KILOWATTS

THE VOLTAGE OF THE FUEL CELL IS          0.8327 VOLTS.
THE AREA OF THE FUEL CELL IS           5744276. SQ. FT.

OVERALL PLANT EFFICIENCY, COAL PILE TO BUSBAR, IS 46.8 PERCENT.
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RUN4.1    JPL.UG.4 BASE CASE, FC REFORM, PRE-SHIFT, 13.5% CARB EX COAL TO CH4.            CASE    RUN4.1
                                                                                          POWER LIST
                                                                                       4 FEB 83 14:52:02

****************************************************************************************************

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.
  29    43214.6 KILOWATTS      29 EXPAND SHIFT (I) 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    RUN4.1
RUN4.1    JPL.UG.4 BASE CASE, FC REFORM, PRE-SHIFT, 13.5% CARB EX COAL TO CH4.                STEAM LIST
                                                                                         4 FEB 83 14:52:02

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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 SELEXOL SYSTEM.
52      1500.0 PSIA           8002. MOLS STEAM  52 ADD STEAM VIA TURBINE; 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     RUN4.1
                                                                                                     WASTE HEAT OUT
                                                                                                    4 FEB 83 14:52:02

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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 WHB; RAISE STEAM.
  70      880.0 DEGREES F.     434.77 MILLIONS BTU     70 COOL CATHODE RECYCLE GAS IN WHB - RAISE STEAM.
  81      880.0 DEGREES F.     164.56 MILLIONS BTU     81 COOL ANODE EXIT GAS; RAISE STEAM.
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CASE E

Utilize Expansion Turbines on Low-Pressure Steam Used

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RUN4.2    JPL.UG.4 PLUS BACK-PRESSURE STEAM TURBINES TO GET LP STEAM.                                   4.2
                                                                                                    SUMMARY
                                                                                              5 FEB 83 10:34:00
                      NEW BASE CASE PLUS LP STEAM SUPPLY VIA BACK-PRES TURBS.

   THIS IS THE FINAL SYNOPSIS OF THE ENERGY BUDGET OF AN INTEGRATED POWER PLANT USING COAL
GASIFICATION AND ELECTROCHEMICAL FUEL CELLS. ALONG WITH CONVENTIONAL POWER GENERATION MACHINERY.

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            *********      COAL CONSUMPTION IS   4991. TONS PER DAY.      *********
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            *********      NET DELIVERED POWER IS  764. MEGAWATTS.       *********
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____________________________________________DETAILS____________________________________________

COAL CONSUMPTION IS  416. THOUSANDS POUNDS PER HOUR. OR ABOUT  5314. MILLIONS BTU.
THE OXYGEN PLANT SIZE IS  3023. TONS PER DAY.
PROCESS STEAM CONSUMPTION IS   791. THOUSANDS POUNDS PER HOUR.

PROCESS MACHINERY (COMPRESSORS AND MAJOR PUMPS) NEED                         -280179. KILOWATTS
PROCESS GAS TURBO-EXPANDERS & CLOSELY-ASSOCIATED STEAM TURBINES GIVE          375954. KILOWATTS
CONDENSING STEAM TURBINE (MULTI-STAGE W. REHEAT) GIVES                         15343. KILOWATTS
THE FUEL CELL ITSELF DELIVERS                                                 653248. KILOWATTS
            (AFTER CONVERSION TO 13.6KV-AC @ 98% EFF.)
THE NET POWER DELIVERED TO BUSBAR IS                                          764365. KILOWATTS

THE VOLTAGE OF THE FUEL CELL IS          0.8327 VOLTS.
THE AREA OF THE FUEL CELL IS           5744276. SQ. FT.

OVERALL PLANT EFFICIENCY. COAL PILE TO BUSBAR. IS 49.1 PERCENT.
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RUN4.2     JPL.UG.4 PLUS BACK-PRESSURE STEAM TURBINES TO GET LP STEAM.                              CASE    RUN4.2
                                                                                                          POWER LIST
                                                                                                     5 FEB 83 10:34:00

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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.
   2    16031.3 KILOWATTS      3 GET POWER BY EXPANDING GASIFIER STEAM NEEDS FROM 1500 PSI.
  29    43214.6 KILOWATTS      29 EXPAND SHIFT (I) 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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RUN4.2    JPL.UG.4 PLUS BACK-PRESSURE STEAM TURBINES TO GET LP STEAM.                    CASE    RUN4.2
                                                                                            STEAM LIST
                                                                                       5 FEB 83 10:34:00

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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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RUN4.2    JPL.UG.4 PLUS BACK-PRESSURE STEAM TURBINES TO GET LP STEAM.                         CASE     RUN4.2
                                                                                                  WASTE HEAT OUT
                                                                                                 5 FEB 83 10:34:00

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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 WHB; RAISE STEAM.
  70      880.0 DEGREES F.     434.58 MILLIONS BTU     70 COOL CATHODE RECYCLE GAS IN WHB - RAISE STEAM.
  81      880.0 DEGREES F.     164.62 MILLIONS BTU     81 COOL ANODE EXIT GAS; RAISE STEAM.
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CASE F

Employ Economizer on Final
Tail Gas to 250°F

```
RUN4.3    JPL.UG.4 + LP ST-TURB'S. STACK GAS ECONOMIZER TO 250 DEGREES FAHR.                      SUMMARY
                                                                                                  5 FEB 83 10:40:36
             MANY CREDIBLE IMPROVEMENTS BUT NO RISK OF CARBON FORMATION

  THIS IS THE FINAL SYNOPSIS OF THE ENERGY BUDGET OF AN INTEGRATED POWER PLANT USING COAL
GASIFICATION AND ELECTROCHEMICAL FUEL CELLS, ALONG WITH CONVENTIONAL POWER GENERATION MACHINERY.

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        *********         COAL CONSUMPTION IS   4991. TONS PER DAY.         *********
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        *********         NET DELIVERED POWER IS  798. MEGAWATTS.           *********
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______________________________________________DETAILS_________________________________________

COAL CONSUMPTION IS  416. THOUSANDS POUNDS PER HOUR, OR ABOUT  5314. MILLIONS BTU.
THE OXYGEN PLANT SIZE IS  3023. TONS PER DAY.
PROCESS STEAM CONSUMPTION IS   791. THOUSANDS POUNDS PER HOUR.

PROCESS MACHINERY (COMPRESSORS AND MAJOR PUMPS) NEED                      -280179. KILOWATTS
PROCESS GAS TURBO-EXPANDERS & CLOSELY-ASSOCIATED STEAM TURBINES GIVE       372982. KILOWATTS
CONDENSING STEAM TURBINE (MULTI-STAGE W. REHEAT) GIVES                      52218. KILOWATTS
THE FUEL CELL ITSELF DELIVERS                                              653248. KILOWATTS
                   (AFTER CONVERSION TO 13.6KV-AC @ 98% EFF.)
THE NET POWER DELIVERED TO BUSBAR IS                                       798269. KILOWATTS

THE VOLTAGE OF THE FUEL CELL IS          0.8327 VOLTS.
THE AREA OF THE FUEL CELL IS           5744276. SQ. FT.

OVERALL PLANT EFFICIENCY, COAL PILE TO BUSBAR, IS 51.3 PERCENT.
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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.
   2    16031.3 KILOWATTS      3 GET POWER BY EXPANDING GASIFIER STEAM NEEDS FROM 1500 PSI.
  29    43214.6 KILOWATTS      29 EXPAND SHIFT (I) 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   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.
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                                                                                CASE    RUN4.3
RUN4.3    JPL.UG.4 + LP ST-TURB'S. STACK GAS ECONOMIZER TO 250 DEGREES FAHR.          STEAM LIST
                                                                               5 FEB 83 10:40:36

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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          11823. 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           8002. MOLS STEAM  52 ADD STEAM VIA TURBINE: PREVENT CARBON DROP IN FUEL CELL.
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RUN4.3     JPL.UG.4 • LP ST-TURB'S, STACK GAS ECONOMIZER TO 250 DEGREES FAHR.                          CASE      RUN4.3
                                                                                                         WASTE HEAT OUT
                                                                                                        5 FEB 83 10:40:36

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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 WHB; 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.
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CASE G

Employ Minimum Quenching to Amorphous Carbon

```
RUN4.4    JPL.UG.4 + LP ST-TURB'S, ECON 250, MINIMUM HUMIDITY TO AMORPHOUS LIMIT                CASE     RUN4.4
                                                                                                  SUMMARY
                                                                                               2 FEB 83  8:05:21
              ALL CREDIBLE IMPROVEMENTS EXCEPT HIGH-TEMPERATURE FU-CEL.

   THIS IS THE FINAL SYNOPSIS OF THE ENERGY BUDGET OF AN INTEGRATED POWER PLANT USING COAL
 GASIFICATION AND ELECTROCHEMICAL FUEL CELLS, ALONG WITH CONVENTIONAL POWER GENERATION MACHINERY.

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          *********         COAL CONSUMPTION IS   4991. TONS PER DAY.         *********
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          *********         NET DELIVERED POWER IS  805. MEGAWATTS.           *********
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 ----------------------------------------DETAILS----------------------------------------
 COAL CONSUMPTION IS  416. THOUSANDS POUNDS PER HOUR, OR ABOUT  5314. MILLIONS BTU.
 THE OXYGEN PLANT SIZE IS  3023. TONS PER DAY.
 PROCESS STEAM CONSUMPTION IS   765. THOUSANDS POUNDS PER HOUR.

 PROCESS MACHINERY (COMPRESSORS AND MAJOR PUMPS) NEED                     -280028. KILOWATTS
 PROCESS GAS TURBO-EXPANDERS & CLOSELY-ASSOCIATED STEAM TURBINES GIVE      369862. KILOWATTS
 CONDENSING STEAM TURBINE (MULTI-STAGE W. REHEAT) GIVES                     59065. KILOWATTS
 THE FUEL CELL ITSELF DELIVERS                                             655992. KILOWATTS
                   (AFTER CONVERSION TO 13.6KV-AC @ 98% EFF.)
 THE NET POWER DELIVERED TO BUSBAR IS                                      804891. KILOWATTS

 THE VOLTAGE OF THE FUEL CELL IS         0.8362 VOLTS.
 THE AREA OF THE FUEL CELL IS          5744273. SQ. FT.

 OVERALL PLANT EFFICIENCY, COAL PILE TO BUSBAR, IS 51.7 PERCENT.
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RUN4.4     JPL.UG.4 + LP ST-TURB'S, ECON 250, MINIMUM HUMIDITY TO AMORPHOUS LIMIT                    CASE     RUN4.4
                                                                                                        POWER LIST
                                                                                                      2 FEB 83   8:05:21

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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.
   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.
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                                                                                        CASE    RUN4.4
RUN4.4    JPL.UG.4 + LP ST-TURB'S, ECON 250, MINIMUM HUMIDITY TO AMORPHOUS LIMIT          STEAM LIST
                                                                                        2 FEB 83  8:05:21

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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.
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RUN4.4    JPL.UG.4 + LP ST-TURB'S, ECON 250, MINIMUM HUMIDITY TO AMORPHOUS LIMIT                    CASE     RUN4.4
                                                                                                     WASTE HEAT OUT
                                                                                                    2 FEB 83  8:05:21

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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 WHB; RAISE STEAM.
  70      880.0 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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CASE H

Change Gasifier to
More Optimistic Design Point

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                                                                                              CASE    RUN5
RUN5      600PSI U-GAS, LP ST-TURB'S, ECON 250, MIN. H2O TO AMORF LIM., FC 1260                    SUMMARY
                                                                                               2 FEB 83  7:55:43
             ALL CREDIBLE IMPROVEMENTS EXCEPT HIGH-TEMPERATURE FU-CEL.

  THIS IS THE FINAL SYNOPSIS OF THE ENERGY BUDGET OF AN INTEGRATED POWER PLANT USING COAL
GASIFICATION AND ELECTROCHEMICAL FUEL CELLS, ALONG WITH CONVENTIONAL POWER GENERATION MACHINERY.

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        *********         COAL CONSUMPTION IS   4991. TONS PER DAY.         *********
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        *********         NET DELIVERED POWER IS  803. MEGAWATTS.           *********
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---------------------------------------------DETAILS----------------------------------------

COAL CONSUMPTION IS  416. THOUSANDS POUNDS PER HOUR, OR ABOUT  5311. MILLIONS BTU.
THE OXYGEN PLANT SIZE IS  2914. TONS PER DAY.
PROCESS STEAM CONSUMPTION IS   661. THOUSANDS POUNDS PER HOUR.

PROCESS MACHINERY (COMPRESSORS AND MAJOR PUMPS) NEED                     -280402. KILOWATTS
PROCESS GAS TURBO-EXPANDERS & CLOSELY-ASSOCIATED STEAM TURBINES GIVE      366759. KILOWATTS
CONDENSING STEAM TURBINE (MULTI-STAGE W. REHEAT) GIVES                     51641. KILOWATTS
THE FUEL CELL ITSELF DELIVERS                                             665244. KILOWATTS
                 (AFTER CONVERSION TO 13.6KV-AC @ 98% EFF.)
THE NET POWER DELIVERED TO BUSBAR IS                                      803273. KILOWATTS

THE VOLTAGE OF THE FUEL CELL IS          0.8361 VOLTS.
THE AREA OF THE FUEL CELL IS           5526377. SQ. FT.

OVERALL PLANT EFFICIENCY, COAL PILE TO BUSBAR, IS 51.6 PERCENT.
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RUN5      600PSI U-GAS, LP ST-TURB'S, ECON 250, MIN. H2O TO AMORF LIM., FC 1260                CASE     RUN5
                                                                                                 POWER LIST
                                                                                              2 FEB 83  7:55:43

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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.
  29   46653.4 KILOWATTS      29 EXPAND SHIFT (I) EXIT GAS THROUGH POWER RECOVERY TURBINE.
  38   -6433.0 KILOWATTS      38 REMOVE SULFUR FROM RAW GAS BY SELEXOL SYSTEM.
  39   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  678920.7 KILOWATTS      NORMAL FUEL CELL, 1200 F INPUT SETS VOLTAGE, 1260 F EXIT.
  71  -23498.9 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.
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                                                                                               CASE    RUNS
_RUNS ..... 600PSI U-GAS, LP ST-TURB'S, ECON 250, MIN. H2O TO AMORF LIM., FC 1260                  STEAM LIST
                                                                                               2 FEB 83  7:55:43

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 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.
 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.
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RUNS        600PSI U-GAS, LP ST-TURB'S, ECON 250, MIN. H2O TO AMORF LIM., FC 1260                   CASE     RUNS
                                                                                                  WASTE HEAT OUT
                                                                                                 2 FEB 83  7:55:43

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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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CASE I

As Case G, but
Change Quench System From
Liquid Water to Steam

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                                                                                          CASE     RUN5A
RUN5A     STEAM INJECTION IN PLACE OF RAW GAS QUENCH, OTHERWISE SAME AS CASE 5              SUMMARY
                                                                                     2 FEB 83 11:08:43
              600# U-GAS, STEAM INJECT TO RAW GAS, OTHER IMPROVEMENTS.

  THIS IS THE FINAL SYNOPSIS OF THE ENERGY BUDGET OF AN INTEGRATED POWER PLANT USING COAL
GASIFICATION AND ELECTROCHEMICAL FUEL CELLS, ALONG WITH CONVENTIONAL POWER GENERATION MACHINERY.

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       *********                                                           *********
       *********         COAL CONSUMPTION IS   4991. TONS PER DAY.         *********
       *********                                                           *********
       *********                                                           *********
       *********                                                           *********
       *********         NET DELIVERED POWER IS  822. MEGAWATTS.           *********
       *********                                                           *********
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---------------------------------------------DETAILS---------------------------------------------
COAL CONSUMPTION IS  416. THOUSANDS POUNDS PER HOUR, OR ABOUT  5311. MILLIONS BTU.
THE OXYGEN PLANT SIZE IS  2914. TONS PER DAY.
PROCESS STEAM CONSUMPTION IS   822. THOUSANDS POUNDS PER HOUR.

PROCESS MACHINERY (COMPRESSORS AND MAJOR PUMPS) NEED                    -280555. KILOWATTS
PROCESS GAS TURBO-EXPANDERS & CLOSELY-ASSOCIATED STEAM TURBINES GIVE     379411. KILOWATTS
CONDENSING STEAM TURBINE (MULTI-STAGE W. REHEAT) GIVES                    57644. KILOWATTS
THE FUEL CELL ITSELF DELIVERS                                            665217. KILOWATTS
                 (AFTER CONVERSION TO 13.6KV-AC @ 98% EFF.)
THE NET POWER DELIVERED TO BUSBAR IS                                     821716. KILOWATTS

THE VOLTAGE OF THE FUEL CELL IS          0.8360 VOLTS.
THE AREA OF THE FUEL CELL IS            5926377. SQ. FT.

OVERALL PLANT EFFICIENCY, COAL PILE TO BUSBAR, IS 52.8 PERCENT.
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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    5274.5 KILOWATTS     13 STEAM INJECTION TO PREVENT CARBON DEPOSITION. (VIA TURB)
  29   53998.3 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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                                                                                                  CASE     RUN5A
RUN5A     STEAM INJECTION IN PLACE OF RAW GAS QUENCH. OTHERWISE SAME AS CASE 5                        STEAM LIST
                                                                                                2 FEB 83 11:05:43

**********************************************************************************************************

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           8940. 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.
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RUNSA     STEAM INJECTION IN PLACE OF RAW GAS QUENCH. OTHERWISE SAME AS CASE 5                CASE     RUNSA
                                                                                                WASTE HEAT OUT
                                                                                                2 FEB 83 11:08:43

*****************************************************************************************************************

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 WHB; 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.    337.07 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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CASE J

Remove More CO_2 With H_2S

```
RUN6      LIKE RUN5A BUT REMOVE MORE CO2 @ #38 & RETIRE TURBINE #52.                 CASE    RUN6
                                                                                         SUMMARY
               600# U-GAS, STEAM INJECT TO STABILIZE RAW GAS, DILUTE H2S.           2 FEB 83 15:58:41

   THIS IS THE FINAL SYNOPSIS OF THE ENERGY BUDGET OF AN INTEGRATED POWER PLANT USING COAL
 GASIFICATION AND ELECTROCHEMICAL FUEL CELLS, ALONG WITH CONVENTIONAL POWER GENERATION MACHINERY.

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      *********                                                          *********
      *********         COAL CONSUMPTION IS   4991. TONS PER DAY.        *********
      *********                                                          *********
      *********                                                          *********
      *********                                                          *********
      *********         NET DELIVERED POWER IS  808. MEGAWATTS.          *********
      *********                                                          *********
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 ------------------------------------------DETAILS_____________________________________

 COAL CONSUMPTION IS  416. THOUSANDS POUNDS PER HOUR, OR ABOUT  5311. MILLIONS BTU.
 THE OXYGEN PLANT SIZE IS  2914. TONS PER DAY.
 PROCESS STEAM CONSUMPTION IS   747. THOUSANDS POUNDS PER HOUR.

 PROCESS MACHINERY (COMPRESSORS AND MAJOR PUMPS) NEED                      -274937. KILOWATTS
 PROCESS GAS TURBO-EXPANDERS & CLOSELY-ASSOCIATED STEAM TURBINES GIVE       355806. KILOWATTS
 CONDENSING STEAM TURBINE (MULTI-STAGE W. REHEAT) GIVES                      65261. KILOWATTS
 THE FUEL CELL ITSELF DELIVERS                                              661826. KILOWATTS
                    (AFTER CONVERSION TO 13.6KV-AC @ 98% EFF.)
 THE NET POWER DELIVERED TO BUSBAR IS                                       807957. KILOWATTS

 THE VOLTAGE OF THE FUEL CELL IS          0.8506 VOLTS.
 THE AREA OF THE FUEL CELL IS           5697838. SQ. FT.

 OVERALL PLANT EFFICIENCY, COAL PILE TO BUSBAR, IS 51.9 PERCENT.
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RUN6       LIKE RUN5A BUT REMOVE MORE CO2 @ #38 & RETIRE TURBINE #52.                                                CASE     RUN6
                                                                                                                          POWER LIST
                                                                                                                       2 FEB 83 15:55:41

****************************************************************************************************************************

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     5134.7 KILOWATTS      13 STEAM INJECTION TO PREVENT CARBON DEPOSITION. (VIA TURB)
   29    53359.5 KILOWATTS      29 EXPAND SHIFT (I) 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.
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                                                                                                          CASE     RUN6
RUN6       LIKE RUN5A BUT REMOVE MORE CO2 @ #39 & RETIRE TURBINE #52.                                         STEAM LIST
                                                                                                      2 FEB 83 15:58:41

*****************************************************************************************************************************

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           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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                                                                                                    CASE     RUN6
RUN6      LIKE RUN5A BUT REMOVE MORE CO2 @ #38 & RETIRE TURBINE #52.                                  WASTE HEAT OUT
                                                                                                     2 FEB 83 15:59:41

 ************************************************************************************************************************

TABULATION OF WASTE HEAT AVAILABLE FROM PROCESS BOILERS.

   15      690.0 DEGREES F.     239.96 MILLIONS BTU    15 COOL RAW GAS TO 690 DEGREES FAHRENHEIT; 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 WHB; RAISE STEAM.
   70      880.0 DEGREES F.     490.66 MILLIONS BTU    70 COOL CATHODE RECYCLE GAS IN WHB - RAISE STEAM.
   77      250.0 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    81 COOL ANODE EXIT GAS; RAISE STEAM.
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CASE K

As Case I, but Burn Part of
Anode Exhaust into Cathode Exhaust

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      N7    TW     RNS,    U     , L     TUR     EC     50,       H2     AM     IP.                CASE     RUN7  90%
                                                                                                       SUP
                                                                                                    3 FEB 83 11:20:21
                90 % OF ANODE EXIT BURNED TO CATHODE FEED.   10 % BYPASS.

     THIS IS THE FINAL SYNOPSIS OF THE ENERGY BUDGET OF AN INTEGRATED POWER PLANT USING COAL
   GASIFICATION AND ELECTROCHEMICAL FUEL CELLS, ALONG WITH CONVENTIONAL POWER GENERATION MACHINERY.

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           *********                                          *********
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           *********                                          *********
           *********                                          *********
           *********       COAL CONSUMPTION IS   4991. TONS PER DAY.       *********
           *********                                          *********
           *********                                          *********
           *********                                          *********
           *********       NET DELIVERED POWER IS  816. MEGAWATTS.      *********
           *********                                          *********
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 -------------------------------------------DETAILS--------------------------------------

 COAL CONSUMPTION IS  416. THOUSANDS POUNDS PER HOUR, OR ABOUT  5311. MILLIONS BTU.
 THE OXYGEN PLANT SIZE IS  2914. TONS PER DAY.
 PROCESS STEAM CONSUMPTION IS   822. THOUSANDS POUNDS PER HOUR.

 PROCESS MACHINERY (COMPRESSORS AND MAJOR PUMPS) NEED                     -280178. KILOWATTS
 PROCESS GAS TURBO-EXPANDERS & CLOSELY-ASSOCIATED STEAM TURBINES GIVE      375140. KILOWATTS
 CONDENSING STEAM TURBINE (MULTI-STAGE W. REHEAT) GIVES                     60852. KILOWATTS
 THE FUEL CELL ITSELF DELIVERS                                            660212. KILOWATTS
                (AFTER CONVERSION TO 13.6KV-AC @ 98% EFF.)
 THE NET POWER DELIVERED TO BUSBAR IS                                     816026. KILOWATTS

 THE VOLTAGE OF THE FUEL CELL IS          0.8298 VOLTS.
 THE AREA OF THE FUEL CELL IS            5826377. SQ. FT.

 OVERALL PLANT EFFICIENCY, COAL PILE TO BUSBAR, IS 52.4 PERCENT.
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RUN7  90% TWO BURNS, 600# U-GAS, LP ST-TURB'S, ECON 250, MIN. H2O TO AMORF LIM.          CASE     RUN7  90%
                                                                                              POWER LIST
                                                                                          3 FEB 83 11:20:21

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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     5514.3 KILOWATTS      13 INJECT STEAM VIA POWER RECOVERY TURBINE TO STABILIZE GAS.
  29    53957.2 KILOWATTS      29 EXPAND SHIFT (I) EXIT GAS THROUGH POWER RECOVERY TURBINE.
  38    -6439.0 KILOWATTS      38 REMOVE SULFUR FROM RAW GAS BY SELEXOL SYSTEM.
  38    19372.1 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, 1260 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.
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                                                                                           CASE     RUN7  90%
RUN7  90% TWO BURNS, 600# U-GAS, LP ST-TURB'S, ECON 250, MIN. H2O TO AMORF LIM.                STEAM LIST
                                                                                         3 FEB 93 11:20:21

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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           8940. 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%
RUN7  90% TWO BURNS, 600# U-GAS, LP ST-TURB'S, ECON 250, MIN. H20 TO AMORF LIM.                    WASTE HEAT OUT
                                                                                                  3 FEB 83 11:20:21

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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.    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      830.0 DEGREES F.    160.15 MILLIONS BTU    30 COOL ANODE EXIT GAS; RAISE STEAM.
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CASE L

As Case I, Employ Low-Pressure
Drops Throughout the System

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RUN8      LO-DP, 600# U-GAS, LP ST-TURB'S, STM-STBLE W. MIN. H2O, FC 1260                     CASE     RUN8
                                                                                                  SUMMARY
             LOW DELTA-P'S, 600# U-GAS, STEAM STABILIZED, COMPARE RUN5A                      3 FEB 83 14:46:27

  THIS IS THE FINAL SYNOPSIS OF THE ENERGY BUDGET OF AN INTEGRATED POWER PLANT USING COAL
GASIFICATION AND ELECTROCHEMICAL FUEL CELLS, ALONG WITH CONVENTIONAL POWER GENERATION MACHINERY.

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        *********         COAL CONSUMPTION IS   4991. TONS PER DAY.       *********
        *********                                                         *********
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        *********                                                         *********
        *********         NET DELIVERED POWER IS  841. MEGAWATTS.         *********
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---------------------------------------------DETAILS_______________________________________

COAL CONSUMPTION IS  416. THOUSANDS POUNDS PER HOUR, OR ABOUT  5311. MILLIONS BTU.
THE OXYGEN PLANT SIZE IS  2914. TONS PER DAY.
PROCESS STEAM CONSUMPTION IS   823. THOUSANDS POUNDS PER HOUR.

PROCESS MACHINERY (COMPRESSORS AND MAJOR PUMPS) NEED                    -260032. KILOWATTS
PROCESS GAS TURBO-EXPANDERS & CLOSELY-ASSOCIATED STEAM TURBINES GIVE     387974. KILOWATTS
CONDENSING STEAM TURBINE (MULTI-STAGE W. REHEAT) GIVES                    47649. KILOWATTS
THE FUEL CELL ITSELF DELIVERS                                            665497. KILOWATTS
                  (AFTER CONVERSION TO 13.6KV-AC @ 98% EFF.)
THE NET POWER DELIVERED TO BUSBAR IS                                     841087. KILOWATTS

THE VOLTAGE OF THE FUEL CELL IS          0.9364 VOLTS.
THE AREA OF THE FUEL CELL IS           5926377. SQ. FT.

OVERALL PLANT EFFICIENCY, COAL PILE TO BUSBAR, IS 54.0 PERCENT.
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                                                                                    CASE    RUN8
RUN8      LO-DP, 600# U-GAS, LP ST-TURB'S, STM-STBLE W. MIN. H2O, FC 1260              POWER LIST
                                                                                  3 FEB 83 14:46:29

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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     5255.2 KILOWATTS     13 STEAM INJECTION TO PREVENT CARBON DEPOSITION. (VIA TURB)
  29    60403.0 KILOWATTS     29 EXPAND SHIFT (I) EXIT GAS THROUGH POWER RECOVERY TURBINE.
  38    -6440.2 KILOWATTS     38 REMOVE SULFUR FROM RAW GAS BY SELEXOL SYSTEM.
  38    20448.0 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  679073.2 KILOWATTS     NORMAL 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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                                                                                              CASE     RUN8
RUN8      LO-DP, 600# U-GAS, LP ST-TURB'S, STM-STELE W. MIN. H2O, FC 1250                          STEAM LIST
                                                                                               3 FEB 83 14:46:29

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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           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           6834. MOLS STEAM  52 ADD STEAM VIA TURBINE; PREVENT CARBON DROP IN FUEL CELL.
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                                                                                        CASE     RUN8
RUN8      LO-DP, 600# U-GAS, LP ST-TURB'S, STM-STBLE W. MIN. H2O, FC 1260                 WASTE HEAT OUT
                                                                                         3 FEB 83 14:46:29

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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.      43.69 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      880.0 DEGREES F.     426.55 MILLIONS BTU    70 COOL CATHODE RECYCLE GAS IN WHB - RAISE STEAM.
  77      250.0 DEGREES F.     343.66 MILLIONS BTU    77 FINAL ECONOMIZER ON STACK GAS TO 250. DEG. F.
  81      880.0 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

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RUN9     LO-DP, 600# U-GAS, LP ST-TURB'S, STM-STABLE W. MIN. H2O, FC 1300 EXIT                SUMMARY
                                                                                        3 FEB 83 15:26:42
            HI-TEMP FC, LO D-P'S, 600# U-GAS, STM-STABLE, COMPARE RUNB

   THIS IS THE FINAL SYNOPSIS OF THE ENERGY BUDGET OF AN INTEGRATED POWER PLANT USING COAL
 GASIFICATION AND ELECTROCHEMICAL FUEL CELLS, ALONG WITH CONVENTIONAL POWER GENERATION MACHINERY.

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         **********         COAL CONSUMPTION IS   4991. TONS PER DAY.          **********
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         **********                                                            **********
         **********         NET DELIVERED POWER IS  848. MEGAWATTS.            **********
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--------------------------------------------DETAILS______________________________________

COAL CONSUMPTION IS  416. THOUSANDS POUNDS PER HOUR, OR ABOUT  5311. MILLIONS BTU.
THE OXYGEN PLANT SIZE IS  2914. TONS PER DAY.
PROCESS STEAM CONSUMPTION IS   328. THOUSANDS POUNDS PER HOUR.

PROCESS MACHINERY (COMPRESSORS AND MAJOR PUMPS) NEED                   -257177. KILOWATTS
PROCESS GAS TURBO-EXPANDERS & CLOSELY-ASSOCIATED STEAM TURBINES GIVE    394546. KILOWATTS
CONDENSING STEAM TURBINE (MULTI-STAGE W. REHEAT) GIVES                   43896. KILOWATTS
THE FUEL CELL ITSELF DELIVERS                                           666610. KILOWATTS
              (AFTER CONVERSION TO 13.6KV-AC @ 99% EFF.)
THE NET POWER DELIVERED TO BUSBAR IS                                    847875. KILOWATTS

THE VOLTAGE OF THE FUEL CELL IS         0.8378 VOLTS.
THE AREA OF THE FUEL CELL IS          5826377. SQ. FT.

OVERALL PLANT EFFICIENCY, COAL PILE TO BUSBAR, IS 54.5 PERCENT.
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                                                                          CASE    RUN9
RUN9      LO-DP, 600# U-GAS, LP ST-TURB'S, STM-STBLE W. MIN. H2O, FC 1300 EXIT    POWER LIST
                                                                       3 FEB 83 15:26:42

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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    -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 (I) EXIT GAS THROUGH POWER RECOVERY TURBINE.
 38    -6440.2 KILOWATTS     33 REMOVE SULFUR FROM RAW GAS BY SELEXOL SYSTEM.
 38    20448.0 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  289138.3 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
RUN9      LO-DP, 600# U-GAS, LP ST-TURB'S, STM-STBLE W. MIN. H2O, FC 1300 EXIT          STEAM LIST
                                                                                    3 FEB 83 15:26:42

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TABULATION OF STEAM NEEDED IN PROCESS.

 2       650.0 PSIA          19054. 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           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           6834. MOLS STEAM  52 ADD STEAM VIA TURBINE; PREVENT CARBON DROP IN FUEL CELL.
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RUN9      LO-DP, 600# U-GAS, LP ST-TURB'S, STM-STPLE W. MIN. H2O, FC 1300 EXIT                    CASE    RUN9
                                                                                                     WASTE HEAT OUT
                                                                                                    3 FEB 83 15:26:42

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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.     303.32 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.
  91      830.0 DEGREES F.     196.61 MILLIONS BTU    91 COOL ANODE EXIT GAS; RAISE STEAM.
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